
Handbook of Industrial Drying

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Arun S. Mujumdar (editor)

Third Edition

Foreword to the First Edition

The *Handbook of Industrial Drying* fills an important need and is of immeasurable value in the field of drying. Academics, students, and industry people—from sales to research—can learn much from the combination of principles and practices used throughout. The presentation of principles does not overwhelm the coverage of equipment and systems. More appropriate theories will develop as a result of the description of equipment and systems. For example, a description of dryers, particularly industrial dryers, is lacking in many research articles; this handbook provides such information.

The authors have distilled much information from extensive literature to provide generic information as contrasted with details of a specific drying system of a particular manufacturer. The users can extrapolate the use of drying systems, by design and management, to a variety of products. As a special feature, a complete listing of books written on the subject of drying is included.

The authors, a blend of students, faculty, and those in industry, represent experience with different kinds

of drying systems, different applications of principles, and different products. The book provides excellent coverage of the cross-disciplinary nature of drying by utilizing well-known authors from many countries of the world. Dr. Mujumdar and his associates have assembled an excellent up-to-date handbook.

The common thread throughout the book is the movement of heat and moisture as well as the movement and handling of products. Also included are instrumentation, sensors, and controls that are important for quality control of products and efficiency of operation. The emphasis on the design of equipment to expedite these processes in an economical manner is appropriate and useful.

The word *handbook* is sometimes used disparagingly to describe a reference for quick answers to limited questions or problems. In that sense this book is more than a handbook—the knowledge base provided permits the user to build different systems for products other than those covered.

Carl W. Hall

Foreword to the Second Edition

The second edition of the *Handbook of Industrial Drying* continues the tradition of the editor and the publisher as international leaders in providing information in the field of industrial drying. The authors are knowledgeable of the subjects and have been chosen from among the world's authorities in industry, academia, government, and consulting. Some 50 authors from 15 countries have written 43 chapters plus 3 appendices. There are 21 new chapters, plus 2 new appendices. All chapters have been updated or revised. There is over 60% new material, making this edition practically a new volume.

The mark of an outstanding handbook is that it provides current information on a subject—in this case multidisciplinary in nature—understandable to a broad audience. A balanced approach of covering principles and practices provides a sound basis for the presentations. Students, academics, consultants, and industry people can find information to meet their needs. Researchers, designers, manufacturers, and sales people can benefit from the book as they consider elements or components related to drying as well as the system itself.

New material has been added to provide the latest information on minimizing environmental impacts, increasing energy efficiency, maintaining quality control, improving safety of operation, and improving the control of drying systems. New sections or chapters have been added to cover in detail microwave drying; infrared drying; impinging stream dryers; use of superheated steam and osmotic dehydration; and drying of biotechnological materials, tissue and towels, peat, coal, and fibrous materials.

The information in this book can be categorized as product related, equipment related, and the relationship between the two—the system of drying. For products not specifically covered, or for the design of dryers not detailed, users can select closely related applicable information to meet many needs. The user may want to pursue a subject in considerably more detail. Pertinent references, but not voluminous overwhelming bibliographies, are included at the end of each chapter. An appendix devoted to an annotated bibliography is also included.

Carl W. Hall

Foreword to the Third Edition

The *Handbook of Industrial Drying*, as a result of the great success of its first and second editions, has gained high reputation among readers interested in the process of drying. In the last three decades we have observed a growing interest in the multidisciplinary subject of drying which had resulted in a major increase of research activity, publication of several monographs, book series, technical papers, international journals, several drying conference series in almost all continents, etc. Today drying R&D continues worldwide at a pace unmatched in any earlier period. To keep abreast with all these scattered sources of information in a broad area like drying is extremely difficult for most readers in academia and industry alike.

So, the third edition of the *Handbook*, nearly a decade after the second edition, will play a very important role in providing comprehensive, updated information and a view of the current state of the art in industrial drying as a more cohesive whole.

This third edition continues the style of the two previous ones; the authors are international leaders and generally recognized world authorities from academia, industry, and R&D laboratories from many countries. It maintains the essential interdisciplinary character addressing a broad academic and industrial readership. This book gives the possibility for self-study and of finding a clear overview of the fundamentals and practical information in broad aspects and problems of drying technology. It is like having one's own private "consultant on the desk."

The topics chosen are constructed to give a quick and clear overview of the fundamental principles and

many practical data referring to the selection of industrial dryers, description of drying equipment, industrial drying technologies, recent developments in R&D in drying as well as future trends. Over 60% of the chapters are new and some 40% revised. A few chapters have been deleted from the second edition due to space limitations. New sections have been added to encompass the latest data on drying of several materials (foods, wood, herbal medicines, sludge, grain, nano size products, fish and seafood, etc.); some dryer types (rotary, indirect, drum, fluidized, flush and pneumatic, etc.) with a strong general approach to energy, environmental safety, control and quality aspects. So practically, this edition can be treated as a truly new *Handbook of Industrial Drying* based on the latest achievements in the drying area.

Finally, having in mind the international character of the authors, this Handbook gives readers a chance to get acquainted in considerable detail with the literature sources published not only in English but also in other languages. Key relevant references are included at the end of each chapter.

I am confident that this third edition of the Handbook will be of great help to the broad audience from academia and in the application, progress and future trends in drying R&D on a global scale.

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Preface to the First Edition

Drying of solids is one of the oldest and most common unit operations found in diverse processes such as those used in the agricultural, ceramic, chemical, food, pharmaceutical, pulp and paper, mineral, polymer, and textile industries. It is also one of the most complex and least understood operations because of the difficulties and deficiencies in mathematical descriptions of the phenomena of simultaneous—and often coupled and multiphase—transport of heat, mass, and momentum in solid media. Drying is therefore an amalgam of science, technology, and art (or know-how based on extensive experimental observations and operating experience) and is likely to remain so, at least for the foreseeable future.

Industrial as well as academic interest in solids drying has been on the rise for over a decade, as evidenced by the continuing success of the Biennial Industrial Drying Symposia (IDS) series. The emergence of several book series and an international journal devoted exclusively to drying and related areas also demonstrates the growing interest in this field. The significant growth in research and development activity in the western world related to drying and dewatering was no doubt triggered by the energy crunch of the early 1970s, which increased the cost of drying several-fold within only a few years. However, it is worth noting that continued efforts in this area will be driven not only by the need to conserve energy, but also by needs related to increased productivity, better product quality, quality control, new products and new processes, safer and environmentally superior operation, etc.

This book is intended to serve both the practicing engineer involved in the selection or design of drying systems and the researcher as a reference work that covers the wide field of drying principles, various commonly used drying equipment, and aspects of drying in important industries. Since industrial dryers can be finely categorized into over 200 variants and, furthermore, since they are found in practically all major industrial sectors, it is impossible within limited space to cover all aspects of drying and dryers. We have had to make choices. In view of the availability of such publications as *Advances in Drying* and the *Proceedings of the International Drying Symposia*, which emphasize research and development in solids drying, we decided to concentrate on various practical aspects of commonly used industrial dryers following a brief introduction to the basic principles, classifica-

tion and selection of dryers, process calculation schemes, and basic experimental techniques in drying. For detailed information on the fundamentals of drying, the reader is referred to various textbooks in this area.

The volume is divided into four major parts. Part I covers the basic principles, definitions, and process calculation methods in a general but concise fashion. The second part is devoted to a series of chapters that describe and discuss the more commonly used industrial dryers. Novel and less prevalent dryers have been excluded from coverage; the reader will find the necessary references in Appendix B, which lists books devoted to drying and related areas in English as well as other languages. Part III is devoted to the discussion of current drying practices in key industrial sectors in which drying is a significant if not necessarily dominant operation. Some degree of repetition was unavoidable since various dryers are discussed under two possible categories. Most readers will, however, find such information complementary as it is derived from different sources and generally presented in different contexts.

Because of the importance of gas humidity measurement techniques, which can be used to monitor and control the convective drying operation, Part IV includes a chapter that discusses such techniques. Energy savings in drying via the application of energy recovery techniques, and process and design modifications, optimization and control, and new drying techniques and nonconventional energy sources are also covered in some depth in the final part of the book.

Finally, it is my pleasant duty to express my sincerest gratitude to the contributors from industry and academia, from various parts of the world, for their continued enthusiasm and interest in completing this major project. The comments and criticisms received from over 25 reviewers were very valuable in improving the contents within the limitations of space. Many dryer manufacturers assisted me and the contributors directly or indirectly, by providing nonproprietary information about their equipment. Dr. Maurits Dekker, Chairman of the Board, Marcel Dekker, Inc., was instrumental in elevating the level of my interest in drying so that I was able to undertake the major task of compiling and editing a handbook in a truly multidisciplinary area whose advancement depends on closer industry-academia interaction and cooperation. My heartfelt thanks go to Chairman Mau for his kindness, continuous

encouragement, and contagious enthusiasm throughout this project.

Over the past four years, many of my graduate students provided me with enthusiastic assistance in connection with this project. In particular, I wish to thank Mainul Hasan and Victor Jariwala for their help and support. In addition, Purnima and Anita Mujumdar kindly word-processed countless drafts

of numerous chapters. Without the assistance of my coauthors, it would have been impossible to achieve the degree of coverage attained in this book. I wish to record my appreciation of their efforts. Indeed, this book is a result of the combined and sustained efforts of everyone involved.

Arun S. Mujumdar

Preface to the Second Edition

The second edition of the *Handbook of Industrial Drying* is a testimonial to the success of the first edition published in 1987. Interest in the drying operation has continued to increase on a truly global scale over the past decade. For example, over 1500 papers have been presented at the biennial International Drying Symposia (IDS) since its inception in 1978. *Drying Technology—An International Journal* published some 2000 pages in seven issues in 1993 compared with just over 300, only a decade earlier. The growth in drying R&D is stimulated by the need to design and operate dryers more efficiently and produce products of higher quality.

A handbook is expected to provide the reader with critical information and advice on appropriate use of such information compiled in a readily accessible form. It is intended to bring together widely scattered information and know-how in a coherent format. Since drying of solids is a multidisciplinary field—indeed, a discipline by itself—it is necessary to call on the expertise of individuals from different disciplines, different industrial sectors, and several countries. A quick perusal of the list of contributors will indicate a balanced blend of authorship from industry as well as academia. An attempt has been made to provide the key elements of fundamentals along with details of industrial dryers and special aspects of drying in specific industries, e.g., foods, pulp and paper, and pharmaceuticals.

The first edition contained 29 chapters and 2 appendixes; this one contains 43 chapters and 3 appendixes. Aside from the addition of new chapters to cover topics missing from the first one, a majority of earlier chapters have been updated—some fully rewritten with new authorship. This edition contains over 60% new updated material. Thus, this book will be a valuable addition even to the bookshelves that already hold the first edition.

This revised and expanded edition follows the same general organization as the first with additions

made to each of the four parts to eliminate some of the weaknesses of the first edition. For example, an extensive chapter is added in Part I on transport properties needed for dryer calculations. Chapters on infrared drying and the novel impinging stream dryers are added to Part II. Part III contains the largest enhancement with ten new chapters while Part IV is completely new except for the chapter on humidity measurements.

A two-volume set of this magnitude must depend on the direct and indirect contributions of a large number of individuals and organizations. Clearly it is impossible to name them all. I am grateful to all the contributors for the valuable time and effort they devoted to this project. The companies and publishers who have permitted us to reproduce some of their copyrighted artwork are acknowledged for their support. Appropriate credits are given in the text where applicable. Exergex Corporation, Brossard, Quebec, Canada provided all the secretarial and related assistance over a three-year period. Without it this revision would have been nearly impossible.

Over the past two years most of my graduate students and postdoctoral fellows of McGill University have provided me with very enthusiastic assistance in various forms in connection with this project. In particular, I wish to express my thanks to Dr. T. Kudra for his continued help in various ways. Purnima, Anita, and Amit Mujumdar kindly word-processed numerous chapters and letters, and helped me keep track of the incredible paperwork involved. The encouragement I received from Dr. Carl W. Hall was singularly valuable in keeping me going on this project while handling concurrently the editorial responsibilities for *Drying Technology—An International Journal* and a host of other books. Finally, the staff at Marcel Dekker, Inc., have been marvellous; I sincerely appreciate their patience and faith in this project.

Arun S. Mujumdar

Preface to the Third Edition

From the success of the second edition of the *Handbook of Industrial Drying* the need for an updated and enhanced edition is realized at this time. Interest in industrial drying operations has been growing continuously over the last three decades and still shows no signs of abatement. This unit operation is central to almost all industrial sectors while exposure to its fundamentals and applications is minimal in most engineering and applied science curricula around the world. The escalating interest in drying is evidenced by the large number of international, regional, and national conferences being held regularly around the world, which are devoted exclusively to thermal and nonthermal dehydration and drying. Although deceptively simple, the processes involved are still too complex to be described confidently in mathematical terms. This means that the design and analyses of industrial dryers remain a combination of science, engineering, and art. It is necessary to have both know-how and know-why of the processes involved to improve the design and operation of dryers. This book represents a comprehensive compendium of collected knowledge of experts from around the world. We are grateful to them for contributing to this effort.

As in the earlier editions, we have a blend of academic and industry-based authors. The academics were carefully selected to ensure they also have industrial background so that readers can reliably utilize the knowledge embedded in this book. Nevertheless, we need to include information and resources available in the public domain; despite our best intentions and high degree of selectivity, we cannot assume responsibility for validity of all the data and information given in this book. Readers must exercise due diligence before using the data in an industrial design or operation.

About two thirds of this book contains new material written by new authors using recent literature. A few topics from the second chapter are deleted. Numerous chapters are totally rewritten with new authorship. At least ten new chapters have been added to make the coverage encyclopedic. I believe that individuals and libraries who have the second edition in their collection should keep that as an independent reference. The material in it is still relevant since the shelf-life of drying technologies is rather long—several decades!

As some 50,000 materials are estimated to require drying on varying scales, it is obvious that it is impossible to pretend to cover all possible dryer types

and products in any single resource. However, I believe we have covered most of the commonly used drying equipment and ancillaries, as well as addressed industrial sectors where drying is a key operation. In this edition for the first time we have covered several new topics relevant to drying, e.g., risk analysis, crystallization, and frying. We have also covered new and emerging drying technologies in adequate detail.

This book is organized in much the same way as the earlier editions. The main difference is the wider coverage of topics. Once again, a deliberate attempt is made to cover most industrial sectors and make the content useful to industry as well as academia. Students and instructors in many disciplines will find the content useful for teaching, design, and research. It is particularly useful for researchers who wish to make their findings relevant to real-world needs.

As energy costs escalate and environmental impact becomes a serious issue in the coming decade, it is clear that the significance of drying for industry will rise. It is hoped that industry will encourage academia to include the study of drying, both as a basic and as an applied subject, as an essential part of engineering and technical curricula. Industry–university cooperation and active collaboration is essential to gaining in-depth knowledge of drying and dryers. I believe that the rising energy costs and demand for enhanced product quality will drive drying R&D. Although no truly disruptive drying technology appears on the horizon today, it is likely to happen within the next decade. This book addresses some of the new technologies that have the potential to be disruptive.

Production of a massive handbook such as this one is a collective effort of scores of dedicated and enthusiastic individuals from around the globe. Indeed, this book embodies a result of globalization. Aside from the authors and referees, numerous staff members initially at Marcel Dekker, New York, and then at Taylor & Francis, Philadelphia, have helped move this project along over a period of nearly five years. Purnima Mujumdar, as usual, played a pivotal part in bringing this project to a successful closure. Without her enthusiastic volunteer effort it is highly unlikely this book would have seen the proverbial end of the tunnel. A number of my postgraduate students at McGill, National University of Singapore, and indeed many overseas institutions also assisted in various ways for which I want express my gratitude.

The encouragement I received regularly from Dr. Carl Hall was instrumental in keeping the project alive and kicking over very long periods, especially since it competed for my leisure time used to edit *Drying Technology—An International Journal* and several other books, as well as organizational effort for many drying-related conferences such as IDS, ADC,

NDC, IWSID, etc. I thank the authors for their patience and effort in making this third edition a valuable reference work.

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Editor

Arun S. Mujumdar is currently professor of mechanical engineering at the National University of Singapore, Singapore, and adjunct professor of chemical as well as agricultural and biosystems engineering at McGill University, Montreal, Canada. Until 2000, he was professor of chemical engineering at McGill. He earned his B.Chem.Eng. with distinction from UDCT, University of Mumbai, India, and his M.Eng. and Ph.D., both in chemical engineering, from McGill. He has published over 300 refereed publications in heat/mass transfer and drying. He has worked on experimental and modeling projects involving almost all physical forms of wet products to be dried in at least 20 different drying configurations, many of which were his original ideas that were later carried forward by others. He has supervised over 40 Ph.D. students and over 30 postdoctoral researchers at McGill, National University of Singapore, as well as in several other countries. Dr. Mujumdar has won numerous international awards and honors for his distinguished contributions to chemical engineering in general, and to drying as well as heat and mass transfer in particular. Founder/program chairman of the International Drying Symposium (IDS) and cofounder of the sister symposia ADC, IADC, NDC series, he is a frequent keynote speaker at major international conferences and a consultant in drying technology for numerous multinational companies. He serves as the editor-in-chief of the premier archival

journal *Drying Technology—An International Journal*. He is also the editor of over 50 books including the widely acclaimed *Handbook of Industrial Drying* (Marcel Dekker, New York) now undergoing third enhanced edition. His recent book, *Mujumdar's Practical Guide to Industrial Drying*, has already been translated into several languages including Chinese, Indonesian, French, Vietnamese, and Hungarian. Dr. Mujumdar has lectured in 38 countries across 4 continents. He has also given professional development courses to industrial and academic audiences in the United States, Canada, Japan, China, and India. Details of his research activities and interests in drying can be found at www.geocities.com/AS_Mujumdar.

He has been instrumental in developing the then-neglected field of drying into a major multi- and interdisciplinary field on a truly global scale. Thanks to his missionary efforts, often carried out single-handedly before the field received worldwide recognition, engineers and scientists around the world have been able to pursue their interests in this exciting field, which provides a kaleidoscope of challenging research opportunities for innovation. He is aptly called the Drying Guru—a label he was first given during the presentation of the esteemed Joseph Janus Medal of the Czech Academy of Sciences in Prague in 1990 to honor his countless contributions to chemical engineering and drying technologies.

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Part I

Fundamental Aspects

1 Principles, Classification, and Selection of Dryers

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1.1 INTRODUCTION

Drying commonly describes the process of thermally removing volatile substances (moisture) to yield a solid product. Moisture held in loose chemical combination, present as a liquid solution within the solid or even trapped in the microstructure of the solid, which exerts a vapor pressure less than that of pure liquid, is called *bound* moisture. Moisture in excess of bound moisture is called *unbound* moisture.

When a wet solid is subjected to thermal drying, two processes occur simultaneously:

1. Transfer of energy (mostly as heat) from the surrounding environment to evaporate the surface moisture
2. Transfer of internal moisture to the surface of the solid and its subsequent evaporation due to process 1

The rate at which drying is accomplished is governed by the rate at which the two processes proceed. Energy transfer as heat from the surrounding environment to the wet solid can occur as a result of convection, conduction, or radiation and in some cases as a result of a combination of these effects. Industrial dryers differ in type and design, depending on the principal method of heat transfer employed. In most cases heat is transferred to the surface of the wet solid and then to the interior. However, in dielectric, radio frequency (RF), or microwave freeze drying, energy is supplied to generate heat internally within the solid and flows to the exterior surfaces.

Process 1, the removal of water as vapor from the material surface, depends on the external conditions of temperature, air humidity and flow, area of exposed surface, and pressure.

Process 2, the movement of moisture internally within the solid, is a function of the physical nature of the solid, the temperature, and its moisture content. In a drying operation any one of these processes may be the limiting factor governing the rate of drying, although they both proceed simultaneously throughout the drying cycle. In the following sections we shall discuss the terminology and some of the basic concepts behind the two processes involved in drying.

The separation operation of drying converts a solid, semisolid, or liquid feedstock into a solid product by evaporation of the liquid into a vapor phase through application of heat. In the special case of freeze drying, which takes place below the triple point of the liquid that is removed, drying occurs by sublimation of the solid phase directly into the vapor phase. This definition thus excludes conversion of a liquid phase into a concentrated liquid phase

(evaporation), mechanical dewatering operations such as filtration, centrifugation, sedimentation, supercritical extraction of water from gels to produce extremely high porosity aerogels (extraction) or so-called drying of liquids and gases by the use of molecular sieves (adsorption). Phase change and production of a solid phase as end product are essential features of the drying process. Drying is an essential operation in the chemical, agricultural, biotechnology, food, polymer, ceramics, pharmaceutical, pulp and paper, mineral processing, and wood processing industries.

Drying is perhaps the oldest, most common and most diverse of chemical engineering unit operations. Over 400 types of dryers have been reported whereas over 100 distinct types are commonly available. It competes with distillation as the most energy-intensive unit operation due to the high latent heat of vaporization and the inherent inefficiency of using hot air as the (most common) drying medium. Several studies report national energy consumption for industrial drying operations ranging from 10–15% for United States, Canada, France, and U.K. to 20–25% for Denmark and Germany. The latter figures have been obtained recently based on mandatory energy audit data supplied by industry and hence are more reliable.

Energy consumption in drying ranges from a low value of under 5% for the chemical process industries to 35% for the papermaking operations. In the United States, for example, capital expenditures for dryers are estimated to be in the order of only \$800 million per annum. Thus, the major costs for dryers are in their operation rather than in their initial investment costs.

Drying of various feedstocks is needed for one or several of the following reasons: need for easy-to-handle free-flowing solids, preservation and storage, reduction in cost of transportation, achieving desired quality of product, etc. In many processes, improper drying may lead to irreversible damage to product quality and hence a nonsalable product.

Before proceeding to the basic principles, it is useful to note the following unique features of drying, which make it a fascinating and challenging area for research and development (R&D):

- Product size may range from microns to tens of centimeters (in thickness or depth)
- Product porosity may range from 0 to 99.9%
- Drying times range from 0.25 s (drying of tissue paper) to 5 months (for certain hardwood species)
- Production capacities may range from 0.10 kg/h to 100 tons/h
- Product speeds range from 0 (stationary) to 2000 m/min (tissue paper)
- Drying temperatures range from below the triple point to above the critical point of the liquid

- Operating pressure may range from fraction of a millibar to 25 atm
- Heat may be transferred continuously or intermittently by convection, conduction, radiation, or electromagnetic fields

Clearly, no single design procedure that can apply to all or even several of the dryer variants is possible. It is therefore essential to revert to the fundamentals of heat, mass and momentum transfer coupled with knowledge of the material properties (quality) when attempting design of a dryer or analysis of an existing dryer. Mathematically speaking, all processes involved, even in the simplest dryer, are highly nonlinear and hence scale-up of dryers is generally very difficult. Experimentation at laboratory and pilot scales coupled with field experience and know how for it is essential to the development of a new dryer application. Dryer vendors are necessarily specialized and normally offer only a narrow range of drying equipment. The buyer must therefore be reasonably conversant with the basic knowledge of the wide assortment of dryers and be able to come up with an informal preliminary selection before going to the vendors with notable exceptions. In general, several different dryers may be able to handle a given application.

Drying is a complex operation involving transient transfer of heat and mass along with several rate processes, such as physical or chemical transformations, which, in turn, may cause changes in product quality as well as the mechanisms of heat and mass transfer. Physical changes that may occur include shrinkage, puffing, crystallization, and glass transitions. In some cases, desirable or undesirable chemical or biochemical reactions may occur, leading to changes in color, texture, odor, or other properties of the solid product. In the manufacture of catalysts, for example, drying conditions can yield significant differences in the activity of the catalyst by changing the internal surface area.

Drying occurs by effecting vaporization of the liquid by supplying heat to the wet feedstock. As noted earlier, heat may be supplied by convection (direct dryers), by conduction (contact or indirect dryers), radiation or volumetrically by placing the wet material in a microwave or RF electromagnetic field. Over 85% of industrial dryers are of the convective type with hot air or direct combustion gases as the drying medium. Over 99% of the applications involve removal of water. All modes except the dielectric (microwave and RF) supply heat at the boundaries of the drying object so that the heat must diffuse into the solid primarily by conduction. The liquid must travel to the boundary of the material

before it is transported away by the carrier gas (or by application of vacuum for nonconvective dryers).

Transport of moisture within the solid may occur by any one or more of the following mechanisms of mass transfer:

- Liquid diffusion, if the wet solid is at a temperature below the boiling point of the liquid
- Vapor diffusion, if the liquid vaporizes within material
- Knudsen diffusion, if drying takes place at very low temperatures and pressures, e.g., in freeze drying
- Surface diffusion (possible although not proven)
- Hydrostatic pressure differences, when internal vaporization rates exceed the rate of vapor transport through the solid to the surroundings
- Combinations of the above mechanisms

Note that since the physical structure of the drying solid is subject to change during drying, the mechanisms of moisture transfer may also change with elapsed time of drying.

1.2 EXTERNAL CONDITIONS (PROCESS 1)

Here the essential external variables are temperature, humidity, rate and direction of airflow, the physical form of the solid, the desirability of agitation, and the method of supporting the solid during the drying operation [1]. External drying conditions are especially important during the initial stages of drying when unbound surface moisture is removed. In certain cases, for example, in materials like ceramics and timber in which considerable shrinkage occurs, excessive surface evaporation after the initial free moisture has been removed sets up high moisture gradients from the interior to the surface. This is liable to cause over-drying and excessive shrinkage and consequently high tension within the material, resulting in cracking and warping. In these cases surface evaporation should be retarded through the employment of high air relative humidities while maintaining the highest safe rate of internal moisture movement by heat transfer.

Surface evaporation is controlled by the diffusion of vapor from the surface of the solid to the surrounding atmosphere through a thin film of air in contact with the surface. Since drying involves the interphase transfer of mass when a gas is brought in contact with a liquid in which it is essentially insoluble, it is necessary to be familiar with the equilibrium characteristics of the wet solid. Also, since the mass transfer is usually accompanied by the simultaneous transfer of heat, due consideration must be given to the enthalpy characteristics.

1.2.1 VAPOR–LIQUID EQUILIBRIUM AND ENTHALPY FOR A PURE SUBSTANCE

VAPOR–PRESSURE CURVE

When a liquid is exposed to a dry gas, the liquid evaporates, that is, forms vapor and passes into the gaseous phase. If m_W is the mass of vapor in the gaseous phase, then this vapor exerts a pressure over the liquid, the *partial pressure*, which, assuming ideal gas behavior for the vapor, is given by

$$P_W V = \frac{m_W}{M_W} RT \quad \text{or} \quad P_W V_W = RT \quad (1.1)$$

The maximum value of P_W that can be reached at any temperature is the saturated vapor pressure P_W^0 . If the vapor pressure of a substance is plotted against temperature, a curve such as TC of Figure 1.1 is obtained. Also plotted in the figure are the solid–liquid equilibrium curve (melting curve) and the solid–vapor (sublimation) curve. The point T in the graph at which all three phases can coexist is called the *triple point*. For all conditions along the curve TC , liquid and vapor may coexist, and these points correspond with the *saturated liquid* and the *saturated vapor state*. Point C is the critical point at which distinction between the liquid and vapor phases disappears, and all properties of the liquid, such as density, viscosity, and refractive index, are identical with those of the vapor. The substance above the critical temperature is called a gas, the temperature corresponding to a pressure at each point on the curve TC is the *boiling point*, and that corresponding to a pressure of 101.3 kPa is the *normal boiling point*.

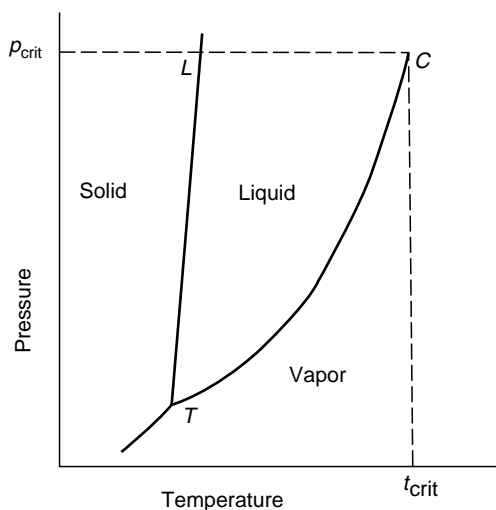


FIGURE 1.1 Vapor pressure of a pure liquid.

1.2.1.1 The Clausius–Clapeyron Equation

Comprehensive tables of vapor-pressure data of common liquids, such as water, common refrigerants, and others, may be found in Refs. [2,3]. For most liquids, the vapor–pressure data are obtained at a few discrete temperatures, and it might frequently be necessary to interpolate between or extrapolate beyond these measurement points. At a constant pressure, the Clausius–Clapeyron equation relates the slope of the vapor pressure–temperature curve to the latent heat of vaporization through the relation

$$\frac{dP_W^0}{dT} = \frac{\Delta H_W}{T(V_W - V_L)} \quad (1.2)$$

where V_W and V_L are the specific molar volumes of saturated vapor and saturated liquid, respectively, and ΔH_W is the molar latent heat of vaporization. Since the molar volume of the liquid is very small compared with that of the vapor, we neglect V_L and substitute for V_W from Equation 1.1 to obtain

$$d \ln P_W^0 = \frac{\Delta H_W}{RT^2} dT \quad (1.3)$$

Since ΔH_W could be assumed to be a constant over short temperature ranges, Equation 1.3 can be integrated to

$$\ln P_W^0 = -\frac{\Delta H_W}{RT} + \text{constant} \quad (1.4)$$

and this equation can be used for interpolation. Alternatively, reference-substance plots [6] may be constructed. For the reference substance,

$$d \ln P_R^0 = \frac{\Delta H_R}{RT^2} dT \quad (1.5)$$

Dividing Equation 1.3 by Equation 1.5 and integrating provides

$$\ln P_W^0 = \frac{M_W \Delta H_W}{M_R \Delta H_R} \ln P_R^0 + \text{constant} \quad (1.6)$$

The reference substance chosen is one whose vapor pressure data are known.

1.2.1.2 Enthalpy

All substances have an internal energy due to the motion and relative position of the constituent atoms and molecules. Absolute values of the internal

energy, u , are unknown, but numerical values relative to an arbitrarily defined baseline at a particular temperature can be computed. In any steady flow system there is an additional energy associated with forcing streams into a system against a pressure and in forcing streams out of the system. This flow work per unit mass is PV , where P is the pressure and V is the specific volume. The internal energy and the flow work per unit mass have been conveniently grouped together into a composite energy called the enthalpy H . The enthalpy is defined by the expression

$$H = u + PV \quad (1.7)$$

and has the units of energy per unit mass (J/kg or N m/kg).

Absolute values of enthalpy of a substance like the internal energy are not known. Relative values of enthalpy at other conditions may be calculated by arbitrarily setting the enthalpy to zero at a convenient reference state. One convenient reference state for zero enthalpy is liquid water under its own vapor pressure of 611.2 Pa at the triple-point temperature of 273.16 K (0.01°C).

The isobaric variation of enthalpy with temperature is shown in Figure 1.2. At low pressures in the gaseous state, when the gas behavior is essentially ideal, the enthalpy is almost independent of the pressure, so the isobars nearly superimpose on each other. The curves marked “saturated liquid” and “saturated

vapor,” however, cut across the constant pressure lines and show the enthalpies for these conditions at temperatures and pressures corresponding to the equilibrium vapor pressure relationship for the substance. The distance between the saturated vapor and saturated liquid curves, such as the distance $V-L$ corresponds to the latent heat of vaporization at a temperature T . Both T and $V-L$ are dependent on pressure, the distance $V-L$ decreases and becomes zero at the critical temperature T_C . Except near the critical temperature, the enthalpy of the liquid is almost independent of pressure until exceedingly high pressures are reached.

1.2.1.3 Heat Capacity

The *heat capacity* is defined as the heat required to raise the temperature of a unit mass of substance by a unit temperature. For a constant pressure process, the heat capacity C_P is given by

$$C_P = \left(\frac{\partial Q}{\partial T} \right)_P \quad (1.8)$$

where the heat flow Q is the sum of the internal energy change ∂u and the work done against pressure $P \partial V$. Equation 1.8 may be expanded as follows:

$$C_P = \left(\frac{\partial u}{\partial T} \right)_P + P \left(\frac{\partial V}{\partial T} \right)_P = \left(\frac{\partial H}{\partial T} \right)_P \quad (1.9)$$

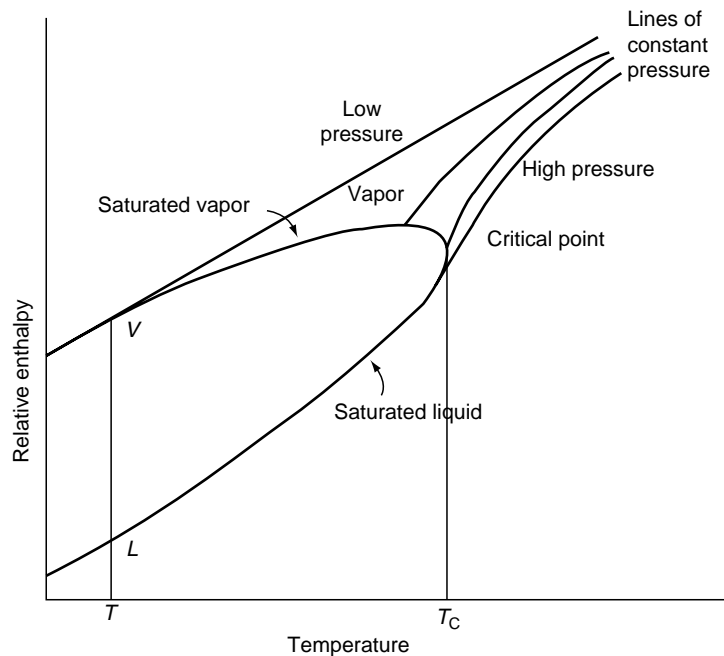


FIGURE 1.2 Typical enthalpy–temperature diagram for a pure substance.

The slope of the isobars of Figure 1.2 yields the heat capacities.

In drying calculation, it is more convenient to use the mean values of heat capacity over a finite temperature step:

$$\bar{C}_P = \left(\frac{\Delta Q}{\Delta T} \right)_P = \frac{1}{(T_2 - T_1)} \int_{T_1}^{T_2} C_P dT \quad (1.10)$$

Second-order polynomials in temperature have been found to adequately describe the variation of C_P with temperature in the temperature range 300–1500 K [4], but for the temperature changes normally occurring in drying the quadratic term can be neglected.

Thus if

$$C_P = a + bT \quad (1.11)$$

then from Equation 1.10,

$$\bar{C}_P = a + \frac{1}{2}b(T_1 + T_2) = C_P(T_{av}) \quad (1.12)$$

The mean heat capacity is the heat capacity evaluated at the arithmetic mean temperature T_{av} .

From Equation 1.9 and Equation 1.10, the enthalpy of the pure substance can be estimated from its heat capacity by

$$H = \bar{C}_P \theta \quad (1.13)$$

where θ denotes the temperature difference or excess over the zero enthalpy reference state. Heat capacity data for a large number of liquids and vapors are found in Ref. [5].

1.2.2 VAPOR–GAS MIXTURES

When a gas or gaseous mixture remains in contact with a liquid surface, it will acquire vapor from the liquid until the partial pressure of the vapor in the gas mixture equals the vapor pressure of the liquid at the existing temperature. In drying applications, the gas frequently used is air and the liquid used is water. Although common concentration units (partial pressure, mole fraction, and others) based on total quantity of gas and vapor are useful, for operations that involve changes in vapor content of a vapor–gas mixture without changes in the amount of gas, it is more convenient to use a unit based on the unchanging amount of gas.

Humid air is a mixture of water vapor and gas, composed of a mass m_W of water vapor and a mass m_G of gas (air). The *moisture content* or *absolute humidity* can be expressed as

$$Y = \frac{m_W}{m_G} \quad (1.14)$$

The total mass can be written in terms of Y and m_G as

$$m_G + m_W = m_G(1 + Y) \quad (1.15)$$

Using the gas law for vapor and air fractions at constant total volume V and temperature T ,

$$m_G = \frac{P_G V}{RT} M_G \quad \text{and} \quad m_W = \frac{P_W V}{RT} M_W \quad (1.16)$$

Thus,

$$Y = \frac{P_W}{P_G} \frac{M_W}{M_G} \quad (1.17)$$

Using Dalton's law of partial pressures,

$$P = P_W + P_G \quad (1.18)$$

and

$$Y = \frac{P_W}{P - P_W} \frac{M_W}{M_G} \quad (1.19)$$

When the partial pressure of the vapor in the gas equals the vapor pressure of the liquid, an equilibrium is reached and the gas is said to be *saturated* with vapor. The ideal saturated absolute humidity is then

$$Y_S = \frac{P_W}{P - P_W^0} \frac{M_W}{M_G} \quad (1.20)$$

The *relative humidity* ψ of a vapor–gas mixture is a measure of its fractional saturation with moisture and is defined as the ratio of the partial pressure of the vapor P_W to the saturated pressure P_W^0 at the same temperature. Thus ψ is given by

$$\psi = \frac{P_W}{P_W^0} \quad (1.21)$$

Equation 1.19 may now be written as

$$Y = \frac{M_W}{M_G} \frac{\psi P_W^0}{P - \psi P_W^0} \quad (1.22)$$

For water vapor and air when $M_W = 18.01$ kg/kmol and $M_G = 28.96$ kg/kmol, respectively, Equation 1.22 becomes

$$Y = 0.622 \frac{\psi P_W^0}{P - \psi P_W^0} \quad (1.23)$$

1.2.3 UNSATURATED VAPOR-GAS MIXTURES: PSYCHROMETRY IN RELATION TO DRYING

If the partial pressure of the vapor in the vapor-gas mixture is for any reason less than the vapor pressure of the liquid at the same temperature, the vapor-gas mixture is said to be unsaturated. As mentioned earlier, two processes occur simultaneously during the thermal process of drying a wet solid, namely, heat transfer to change the temperature of the wet solid and to evaporate its surface moisture and the mass transfer of moisture to the surface of the solid and its subsequent evaporation from the surface to the surrounding atmosphere. Frequently, the surrounding medium is the drying medium, usually heated air or combustion gases. Consideration of the actual quantities of air required to remove the moisture liberated by evaporation is based on psychrometry and the use of humidity charts. The following are definitions of expressions used in psychrometry [6].

1.2.3.1 Dry Bulb Temperature

This is the temperature of a vapor-gas mixture as ordinarily determined by the immersion of a thermometer in the mixture.

1.2.3.2 Dew Point

This is the temperature at which a vapor-gas mixture becomes saturated when cooled at a constant total pressure out of contact with a liquid (i.e., at constant absolute humidity). The concept of the dew point is best illustrated by referring to Figure 1.3, a plot of the absolute humidity versus temperature for a fixed pressure and the same gas. If an unsaturated mixture initially at point F is cooled at constant pressure out of contact of liquid, the gas saturation increases until the point G is reached, when the gas is fully saturated. The temperature at which the gas is fully saturated is called the dew point T_D . If the temperature is

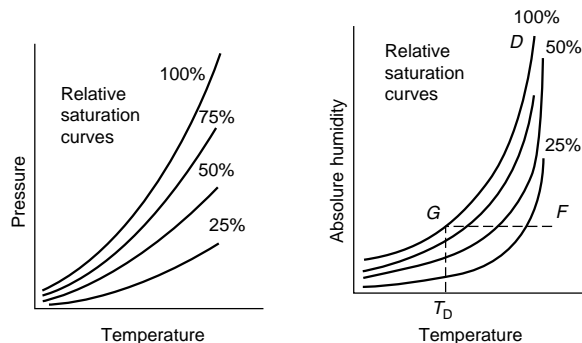


FIGURE 1.3 Two forms of psychrometric charts.

reduced to an infinitesimal amount below T_D , the vapor will condense and the process follows the saturation curve.

While condensation occurs the gas always remains saturated. Except under specially controlled circumstances, supersaturation will not occur and no vapor-gas mixture whose coordinates lie to the left of the saturation curve will result.

1.2.3.3 Humid Volume

The humid volume V_H of a vapor-gas mixture is the volume in cubic meters of 1 kg of dry gas and its accompanying vapor at the prevailing temperature and pressure. The volume of an ideal gas or vapor at 273 K and 1 atm (101.3 kPa) is 22.4 m³/kg mol. For a mixture with an absolute humidity Y at T_G (K) and P (atm), the ideal gas law gives the humid volume as

$$V_H = \left(\frac{1}{M_G} + \frac{Y}{M_W} \right) 22.4 \frac{T}{273.14} \frac{1}{P}$$

$$V_H = 0.082 \left(\frac{1}{M_G} + \frac{Y}{M_W} \right) \frac{T}{P} \quad (1.24)$$

When the mass of dry gas in the vapor-gas mixture is multiplied by the humid volume, the volume of the vapor-gas mixture is obtained. The humid volume at saturation is computed with $Y = Y_S$, and the specific volume of the dry gas can be obtained by substituting $Y = 0$. For partially saturated mixtures, V_H may be interpolated between values for 0 and 100% saturation at the same temperature and pressure.

1.2.3.4 Enthalpy

Since the enthalpy is an extensive property, it could be expected that the enthalpy of a humid gas is the sum of the partial enthalpies of the constituents and a term to take into account the heat of mixing and other effects. The humid enthalpy I_G is defined as the enthalpy of a unit mass of dry gas and its associated moisture. With this definition of enthalpy,

$$I_G = H_{GG} + YH_{GW} + \Delta H_{GM} \quad (1.25)$$

where H_{GG} is the enthalpy of dry gas, H_{GW} is the enthalpy of moisture, and ΔH_{GM} is the residual enthalpy of mixing and other effects. In air saturated with water vapor, this residual enthalpy is only -0.63 kJ/kg at 60°C (333.14 K) [3] and is only 1% of H_{GG} ; thus it is customary to neglect the influences of this residual enthalpy.

It is sometimes convenient to express the enthalpy in terms of specific heat. Analogous to Equation 1.13, we could express the enthalpy of the vapor–gas mixture by

$$I_G = \bar{C}_{PY}\theta + \Delta H_{V0}Y \quad (1.26)$$

\bar{C}_{PY} is called the *humid heat*, defined as the heat required to raise the temperature of 1 kg of gas and its associated moisture by 1 K at constant pressure. For a mixture with absolute humidity Y ,

$$\bar{C}_{PY} = \bar{C}_{PG} + \bar{C}_{PW}Y \quad (1.27)$$

where \bar{C}_{PG} and \bar{C}_{PW} are the mean heat capacities of the dry gas and moisture, respectively.

The path followed from the liquid to the vapor state is described as follows. The liquid is heated up to the dew point T_D , vaporized at this temperature, and superheated to the dry bulb temperature T_G . Thus

$$H_{GW} = \bar{C}_{LW}(T_D - T_0) + \Delta H_{VD} + \bar{C}_{PW}(T_G - T_D) \quad (1.28)$$

However, since the isothermal pressure gradient $(\Delta H/\Delta P)_T$ is negligibly small, it could be assumed that the final enthalpy is independent of the vaporization path followed. For the sake of convenience it could be assumed that vaporization occurs at 0°C (273.14 K), at which the enthalpy is zero, and then directly superheated to the final temperature T_G . The enthalpy of the vapor can now be written as

$$H_{GW} = \bar{C}_{PW}(T_G - T_0) + \Delta H_{V0} \quad (1.29)$$

and the humid enthalpy given by

$$I_G = \bar{C}_{PG}(T_G - T_0) + Y(\bar{C}_{PW}(T_G - T_0) + \Delta H_{V0}) \quad (1.30)$$

Using the definition for the humid heat capacity, Equation 1.30 reduces to

$$I_G = \bar{C}_{PY}(T_G - T_0) + \Delta H_{V0}Y \quad (1.31)$$

In Equation 1.31 the humid heat is evaluated at $(T_G + T_0)/2$ and ΔH_{V0} , the latent heat of vaporization at 0°C (273.14 K). Despite its handiness, the use of Equation 1.31 is not recommended above a humidity of 0.05 kg/kg. For more accurate work, it is necessary to resort to the use of Equation 1.28 in conjunction with Equation 1.25. In Equation 1.28 it should be noted that \bar{C}_{LW} is the mean capacity of liquid moisture between

T_0 and T_D , \bar{C}_{PW} is the mean capacity of the moisture vapor evaluated between T_D and T_G , and ΔH_{VD} is the latent heat of vaporization at the dew point T_D . The value of ΔH_{VD} can be approximately calculated from a known latent heat value at temperature T_0 by

$$\frac{\Delta H_{VD}}{\Delta H_{V0}} = \left(\frac{T_D - T_C}{T_0 - T_C} \right)^{1/3} \quad (1.32)$$

where T_C is the critical temperature. Better and more accurate methods of estimating ΔH_{VD} are available in Refs. [5,7].

1.2.4 ENTHALPY–HUMIDITY CHARTS

Using Equation 1.23, Equation 1.25, and Equation 1.28, the enthalpy–humidity diagram for unsaturated air ($\psi < 1$) can be constructed using the parameters ψ and θ . In order to follow the drying process we need access to enthalpy–humidity values. There seems to be no better, convenient, and cheaper way to store these data than in graphic form. The first of these enthalpy–humidity charts is attributed to Mollier. Mollier's original enthalpy–humidity chart was drawn with standard rectangular coordinates (Figure 1.4), but in order to extend the area over which it can be read, an oblique-angle system of coordinates is chosen for $I_G = f(Y)$.

In the unsaturated region, it can be seen from Equation 1.30 that I_G varies linearly with the humidity Y and the temperature T_G . If zero temperature (0°C) is taken as the datum for zero enthalpy, then

$$I_G = \bar{C}_{PG}\theta + Y(\bar{C}_{PW}\theta + \Delta H_{V0}) \quad (1.33)$$

where θ is the temperature in degree Celsius.

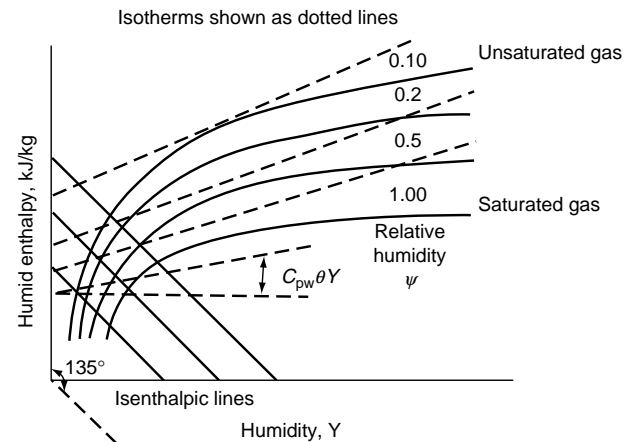


FIGURE 1.4 An enthalpy–humidity diagram for a moist gas.

The isotherms ($\theta = \text{constant}$) cut the ordinate ($Y = 0$) at a value $\bar{C}_{PG}\theta$ (the dry gas enthalpy). If the isenthalpic lines ($I_G = \text{constant}$) are so inclined that they fall with a slope $-\Delta H_{V0}$, and if only $\Delta H_{V0}Y$ were taken into account in the contribution of vapor to the vapor–gas enthalpy, then the isotherms would run horizontally, but because of the contribution of $\bar{C}_{PW}\theta Y$, they increase with Y for $\theta < 0^\circ\text{C}$ and decrease with Y for $\theta > 0^\circ\text{C}$. Contours of relative humidity ψ are also plotted. The region above the curve $\psi = 1$ at which air is saturated corresponds to an unsaturated moist gas; the region below the curve corresponds to fogging conditions. At a fixed temperature air cannot take up more than a certain amount of vapor. Liquid droplets then precipitate due to oversaturation, and this is called the *cloud* or *fog state*. Detailed enthalpy–humidity diagrams are available elsewhere in this handbook and in Ref. [10].

A humidity chart is not only limited to a specific system of gas and vapor but is also limited to a particular total pressure. The thermophysical properties of air may be generally used with reasonable accuracy for diatomic gases [3], so that charts developed for mixtures in air can be used to describe the properties of the same moisture vapor in a gas such as nitrogen. Charts other than those of moist air are often required in the drying of fine chemicals and pharmaceutical products. These are available in Refs. [3,8,9].

1.2.4.1 Adiabatic Saturation Curves

Also plotted on the psychrometric chart are a family of adiabatic saturation curves. The operation of adiabatic saturation is indicated schematically in Figure 1.5. The entering gas is contacted with a liquid and as a result of mass and heat transfer between the gas and liquid the gas leaves at conditions of humidity and

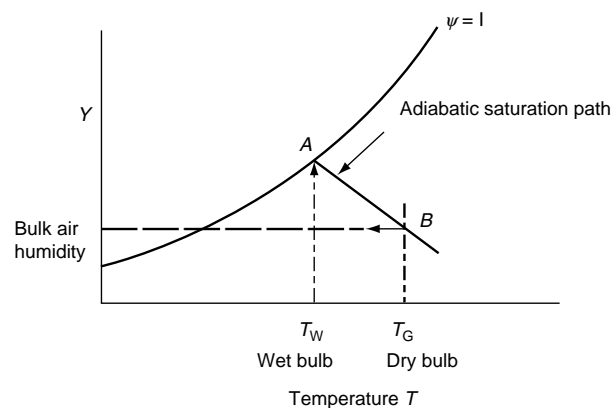


FIGURE 1.5 A temperature–humidity diagram for moist air.

temperature different from those at the entrance. The operation is adiabatic as no heat is gained or lost by the surroundings. Doing a mass balance on the vapor results in

$$G_V = G_G(Y_{\text{out}} - Y_{\text{in}}) \quad (1.34)$$

The enthalpy balance yields

$$I_{G_{\text{in}}} + (Y_{\text{out}} - Y_{\text{in}})I_{LW} = I_{G_{\text{out}}} \quad (1.35)$$

Substituting for I_G from Equation 1.31, we have

$$\begin{aligned} \bar{C}_{PY_{\text{in}}}(T_{\text{in}} - T_0) + \Delta H_{V0}Y_{\text{in}} + (Y_{\text{out}} - Y_{\text{in}})\bar{C}_{LW}(T_L - T_0) \\ = \bar{C}_{PY_{\text{out}}}(T_{\text{out}} - T_0) + \Delta H_{V0}Y_{\text{out}} \end{aligned} \quad (1.36)$$

Now, if a further restriction is made that the gas and the liquid phases reach equilibrium when they leave the system (i.e., the gas–vapor mixture leaving the system is saturated with liquid), then $T_{\text{out}} = T_{GS}$, $I_{G_{\text{out}}} = I_{GS}$, and $Y_{\text{out}} = Y_{GS}$ where T_{GS} is the adiabatic saturation temperature and Y_{GS} is the absolute humidity saturated at T_{GS} . Still further, if the liquid enters at the adiabatic saturation temperature T_{GS} , that is, $T_L = T_{GS}$, Equation 1.36 becomes

$$\begin{aligned} \bar{C}_{PY_{\text{in}}}(T_{\text{in}} - T_0) + \Delta H_{V0}Y_{\text{in}} \\ + (Y_{GS} - Y_{\text{in}})\bar{C}_{LW}(T_{GS} - T_0) \\ = \bar{C}_{PY_{GS}}(T_{GS} - T_0) + \Delta H_{V0}Y_{GS} \end{aligned} \quad (1.37)$$

or substituting for \bar{C}_{PG} from Equation 1.27

$$\begin{aligned} \bar{C}_{PY_{\text{in}}}(T_{\text{in}} - T_0) + Y_{\text{in}}\bar{C}_{PW_{\text{in}}}(T_{\text{in}} - T_0) + \Delta H_{V0}Y_{\text{in}} \\ + (Y_{GS} - Y_{\text{in}})\bar{C}_{LW}(T_{GS} - T_0) \\ = \bar{C}_{PG_{GS}}(T_{GS} - T_0) + \bar{C}_{PW_{GS}}Y_{GS}(T_{GS} - T_0) \\ + \Delta H_{V0}Y_{GS} \end{aligned} \quad (1.38)$$

Assuming that the heat capacities are essentially constant over the temperature range involved, $\bar{C}_{PG_{\text{in}}} = \bar{C}_{PG_{GS}} = \bar{C}_{PG}$ and $\bar{C}_{PW_{\text{in}}} = \bar{C}_{PW_{GS}} = \bar{C}_{PW}$. Further subtracting $Y_{\text{in}}\bar{C}_{PW}T_{GS}$ from both sides of Equation 1.38 and simplifying, we have

$$\begin{aligned} \bar{C}_{PY}(T_{\text{in}} - T_{GS}) \\ = (Y_{GS} - Y_{\text{in}}) \\ \times [(\bar{C}_{PW}(T_{GS} - T_0) + \Delta H_{V0} - \bar{C}_{LW}(T_{GS} - T_0))] \end{aligned} \quad (1.39)$$

From Figure 1.2 the quantity in square brackets is equal to ΔH_{VS} , and thus,

$$\bar{C}_{PY}(T_{in} - T_{GS}) = (Y_{GS} - Y_{in})\Delta H_{VS} \quad (1.40)$$

or

$$T_{in} - T_{GS} = (Y_{GS} - Y_{in}) \frac{\Delta H_{VS}}{\bar{C}_{PY}} \quad (1.41)$$

Equation 1.41 represents the “adiabatic saturation curve” on the psychrometric chart, which passes through the points $A(Y_{GS}, T_{GS})$ on the 100% saturation curve ($\psi = 1$) and $B(Y_{in}, T_{in})$, the initial condition. Since the humid heat contains the term Y_{in} , the curve is not straight but is curved slightly concave upward. Knowing the adiabatic saturation temperature and the actual gas temperature, the actual gas humidity can be easily obtained as the absolute humidity from the saturation locus. Equation 1.40 indicates that the sensible heat given up by the gas in cooling equals the latent heat required to evaporate the added vapor. It is important to note that, since Equation 1.41 is derived from the overall mass and energy balances between the initial gas conditions and the adiabatic saturation conditions, it is applicable only at these points and may not describe the path followed by the gas as it becomes saturated. A family of these adiabatic saturation curves for the air–water system are contained in the psychrometric charts [10].

1.2.4.2 Wet Bulb Temperature

One of the oldest and best-known methods of determining the humidity of a gas is to measure its “wet bulb temperature” and its dry bulb temperature. The wet bulb temperature is the steady temperature reached by a small amount of liquid evaporating into a large amount of rapidly moving unsaturated vapor–gas mixture. It is measured by passing the gas rapidly past a thermometer bulb kept wet by a saturated wick and shielded from the effects of radiation. If the gas is unsaturated, some liquid is evaporated from the wick into the gas stream, carrying with it the associated latent heat. This latent heat is taken from within the liquid in the wick, and the wick is cooled. As the temperature of the wick is lowered, sensible heat is transferred by convection from the gas stream and by radiation from the surroundings. At steady state, the net heat flow to the wick is zero and the temperature is constant.

The heat transfer to the wick can be written as

$$q = (h_C + h_R)A(T_G - T_W) \quad (1.42)$$

where h_C and h_R are the convective and radiative heat transfer coefficients, respectively, T_G is the gas

temperature, T_W is the temperature indicated by thermometer. By using h_R , it is assumed that radiant heat transfer can be approximated:

$$q_R = h_g A(T_G - T_W) \quad (1.43)$$

The rate of mass transfer from the wick is

$$N_G = KA(Y_W - Y_G) \quad (1.44)$$

An amount of heat given by

$$q = N_G \Delta H_{VW} \quad (1.45)$$

is associated with this mass transfer. Since under steady conditions all the heat transferred to the wick is utilized in mass transfer, from Equation 1.42, Equation 1.44, and Equation 1.45 we have

$$T_G - T_W = \frac{K \Delta H_{VW}}{h_C + h_R} (Y_W - Y_G) \quad (1.46)$$

The quantity $T_G - T_W$ is called the *wet bulb depression*. In order to determine the humidity Y_G from Equation 1.46, predictable values of $K \Delta H_{VW}/(h_C + h_R)$ must be obtained. This ratio of coefficients depends upon the flow, boundary, and temperature conditions encountered. In measuring the wet bulb temperature, several precautions are taken to ensure reproducible values of $K \Delta H_{VW}/(h_C + h_R)$. The contribution by radiation is minimized by shielding the wick. The convective heat transfer can be enhanced by making the gas movement past the bulb rapid, often by swinging the thermometer through the gas, as in the sling psychrometer, or by inserting the wet bulb thermometer in a constriction in the gas flow path. Under these conditions Equation 1.46 reduces to

$$T_G - T_W = \frac{K \Delta H_{VW}}{h_C} (Y_W - Y_G) \quad (1.47)$$

For turbulent flow past a wet cylinder, such as a wet bulb thermometer, the accumulated experimental data give

$$\frac{h_C}{K} = 35.53 \left(-\frac{\mu}{\rho D} \right)^{0.56} \text{ J/mol } ^\circ\text{C} \quad (1.48)$$

when air is the noncondensable gas and

$$\frac{h_C}{K} = \bar{C}_{PY} \left(\frac{Sc}{Pr} \right)^{0.56} \quad (1.49)$$

for other gases. Equation 1.49 is based on heat and mass transfer experiments with various gases flowing normal to cylinders. For pure air, $Sc \cong Pr \cong 0.70$ and $h_C/K = 29.08 \text{ J/mol } ^\circ\text{C}$ from Equation 1.48 and Equation 1.49. Experimental data for the air–water system yield values of h_C/K ranging between 32.68 and 28.54 J/mol $^\circ\text{C}$. The latter figure is recommended [11]. For the air–water system, the h_C/K value can be replaced by \bar{C}_{PY} within moderate ranges of temperature and humidity, provided flow is turbulent. Under these conditions, Equation 1.47 becomes identical to the adiabatic saturation curve Equation 1.41 and thus the adiabatic saturation temperature is the same as the wet bulb temperature for the air–water system. For systems other than air–water, they are not the same, as can be seen from the psychrometric charts given by Perry [7].

It is worthwhile pointing out here that, although the adiabatic saturation curve equation does not reveal anything of the enthalpy–humidity path of either the liquid phase or gas phase at various points in the contacting device (except for the air–water vapor system), each point within the system must conform with the wet bulb relation, which requires that the heat transferred be exactly consumed as latent heat of vaporization of the mass of liquid evaporated. The identity of h_C/K with \bar{C}_{PY} was first found empirically by Lewis and hence is called the Lewis relation. The treatment given here on the wet bulb temperature applies only in the limit of very mild drying conditions when the vapor flux becomes directly proportional to the humidity potential ΔY . This is the case in most drying operations.

A more detailed treatment using a logarithmic driving force for vapor flux and the concept of the humidity potential coefficient ϕ while accounting for the influence of the moisture vapor flux on the transfer of heat to the surface, namely, the Ackermann correction ϕ_E , has been given in Ref. [3]. The concept of Luikov number Lu , which is essentially the ratio of the Prandtl number Pr to the Schmidt number Sc , has also been introduced.

1.2.5 TYPES OF PSYCHROMETRIC REPRESENTATION

As stated previously, two processes occur simultaneously during the thermal process of drying a wet solid: heat transfer, to change the temperature of the wet solid, and mass transfer of moisture to the surface of a solid accompanied by its evaporation from the surface to the surrounding atmosphere, which in convection or direct dryers is the drying medium. Consideration of the actual quantities of air required to remove the moisture liberated by evaporation is based on psychrometry and the use of humidity charts. This

procedure is extremely important in the design of forced convection, pneumatic, and rotary dryers. The definitions of terms and expressions involved in psychrometry have been discussed in Section 1.2.3.

There are different ways of plotting humidity charts. One procedure involves plotting the absolute humidity against the dry bulb temperature. A series of curves is obtained for different percentage humidity values from saturation downward (Figure 1.3). On this chart, the saturation humidities are plotted from vapor pressure data with the help of Equation 1.23 to give curve GD . The curve for humidities at 50% saturation is plotted at half the ordinate of curve GD . All curves at constant percentage saturation reach infinity at the boiling point of the liquid at the prevailing pressure.

Another alternative is the graphic representation of conditions of constant relative saturation on a vapor pressure–temperature chart (Figure 1.3). The curve for 50% relative saturation shows a partial pressure equal to one-half of the equilibrium vapor pressure at any temperature. A common method of portraying humidity charts is by using the enthalpy–humidity chart indicated earlier [10].

1.3 INTERNAL CONDITIONS (PROCESS 2)

After having discussed the factors and definitions related to the external conditions of air temperature and humidity, attention will now be paid to the solid characteristics.

As a result of heat transfer to a wet solid, a temperature gradient develops within the solid while moisture evaporation occurs from the surface. This produces a migration of moisture from within the solid to the surface, which occurs through one or more mechanisms, namely, diffusion, capillary flow, internal pressures set up by shrinkage during drying, and, in the case of indirect (conduction) dryers, through a repeated and progressive occurring vaporization and recondensation of moisture to the exposed surface. An appreciation of this internal movement of moisture is important when it is the controlling factor, as it occurs after the critical moisture content, in a drying operation carried to low final moisture contents. Variables such as air velocity and temperature, which normally enhance the rate of surface evaporation, are of decreasing importance except to promote the heat transfer rates. Longer residence times, and, where permissible, higher temperatures become necessary. In the case of such materials as ceramics and timber, in which considerable shrinkage occurs, excessive surface evaporation sets up high moisture gradients from the interior toward the

surface, which is liable to cause overdrying, excessive shrinkage, and, consequently, high tension, resulting in cracking or warping. In such cases, it is essential not to incur too high moisture gradients by retarding surface evaporation through the employment of high air relative humidities while maintaining the highest safe rate of internal moisture movement by virtue of heat transfer. The temperature gradient set up in the solid will also create a vapor–pressure gradient, which will in turn result in moisture vapor diffusion to the surface; this will occur simultaneously with liquid moisture movement.

1.3.1 MOISTURE CONTENT OF SOLIDS

The moisture contained in a wet solid or liquid solution exerts a vapor pressure to an extent depending upon the nature of moisture, the nature of solid, and the temperature. A wet solid exposed to a continuous supply of fresh gas continues to lose moisture until the vapor pressure of the moisture in the solid is equal to the partial pressure of the vapor in the gas. The solid and gas are then said to be in *equilibrium*, and the moisture content of the solid is called the *equilibrium moisture content* under the prevailing conditions. Further exposure to this air for indefinitely long periods will not bring about any additional loss of moisture. The moisture content in the solid could be reduced further by exposing it to air of lower relative humidity. Solids can best be classified as follows [12]:

Nonhygroscopic capillary-porous media, such as sand, crushed minerals, nonhygroscopic crystals, polymer particles, and some ceramics. The defining criteria are as follows. (1) There is a clearly recognizable pore space; the pore space is filled with liquid if the capillary-porous medium is completely saturated and is filled with air when the medium is completely dry. (2) The amount of physically bound moisture is negligible; that is, the material is nonhygroscopic. (3) The medium does not shrink during drying.

Hygroscopic-porous media, such as clay, molecular sieves, wood, and textiles. The defining criteria are as follows. (1) There is a clearly recognizable pore space. (2) There is a large amount of physically bound liquid. (3) Shrinkage often occurs in the initial stages of drying. This category was further classified into (a) hygroscopic capillary-porous media (micropores and macropores, including bi-disperse media, such as wood, clays, and textiles) and (b) strictly hygroscopic media (only micropores, such as silica gel, alumina, and zeolites).

Colloidal (nonporous) media, such as soap, glue, some polymers (e.g., nylons), and various food

products. The defining criteria are as follows: (1) there is no pore space (evaporation can take place only at the surface); (2) all liquid is physically bound.

It should be noted that such classifications are applicable only to homogeneous media that could be considered as continua for transport.

As a wet solid is usually swollen compared with its condition when free of moisture and its volume changes during the drying process, it is not convenient to express moisture content in terms of volume. The moisture content of a solid is usually expressed as the moisture content by weight of bone-dry material in the solid, X . Sometimes a wet basis moisture content W , which is the moisture–solid ratio based on the total mass of wet material, is used. The two moisture contents are related by the expression

$$X = \frac{W}{1 - W} \quad (1.50)$$

Water may become bound in a solid by retention in capillaries, solution in cellular structures, solution with the solid, or chemical or physical adsorption on the surface of the solid. Unbound moisture in a hygroscopic material is the moisture in excess of the equilibrium moisture content corresponding to saturation humidity. All the moisture content of a nonhygroscopic material is unbound moisture. Free moisture content is the moisture content removable at a given temperature and may include both bound and unbound moisture.

In the immediate vicinity of the interface between free water and vapor, the vapor pressure at equilibrium is the saturated vapor pressure. Very moist products have a vapor pressure at the interface almost equal to the saturation vapor pressure. If the concentration of solids is increased by the removal of water, then the dissolved hygroscopic solids produce a fall in the vapor pressure due to osmotic forces. Further removal of water finally results in the surface of the product dried. Water now exists only in the interior in very small capillaries, between small particles, between large molecules, and bound to the molecules themselves. This binding produces a considerable lowering of vapor pressure. Such a product can therefore be in equilibrium only with an external atmosphere in which the vapor pressure is considerably decreased.

1.3.2 MOISTURE ISOTHERMS [10]

A dry product is called hygroscopic if it is able to bind water with a simultaneous lowering of vapor pressure. Different products vary widely in their hygroscopic

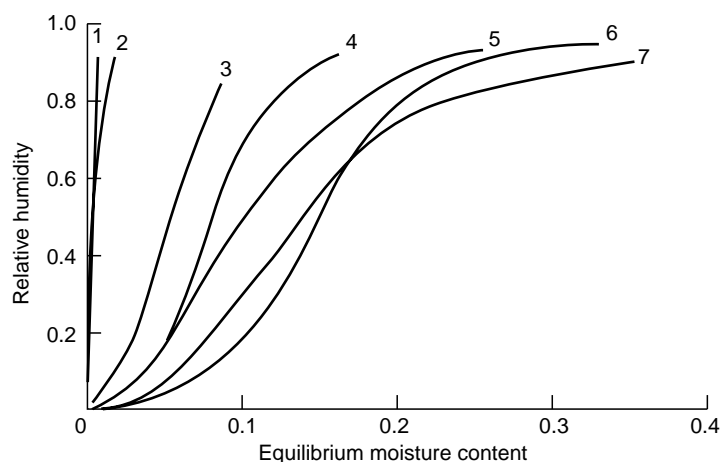


FIGURE 1.6 Typical equilibrium moisture isotherms at room temperature for selected substances: (1) asbestos fiber, (2) PVC (50°C), (3) wood charcoal, (4) Kraft paper, (5) jute, (6) wheat, (7) potatoes.

properties. The reason for this is their molecular structure, their solubility, and the extent of reactive surface.

Sorption isotherms measured experimentally under isothermal conditions are used to describe the hygroscopic properties of a product. A graph is constructed in which the moisture bound by sorption per unit weight is plotted against relative humidity, and vice versa. Such isotherms are shown in Figure 1.6 and Figure 1.7. From Figure 1.7 it is seen that molecular sieves are highly hygroscopic but polyvinyl chloride (PVC) powder is mildly hygroscopic. Potatoes and milk exhibit intermediate hygroscopicity.

Figure 1.8 shows the shape of the sorption isotherm characteristic of many dry food products. If the partial pressure of the external atmosphere P_w is nearly zero, then the equilibrium moisture inside the dry product will also be almost zero. Section A of the curve represents a region in which the monomolecular layers are formed, although there may be multimolecular layers in some places toward the end of A.

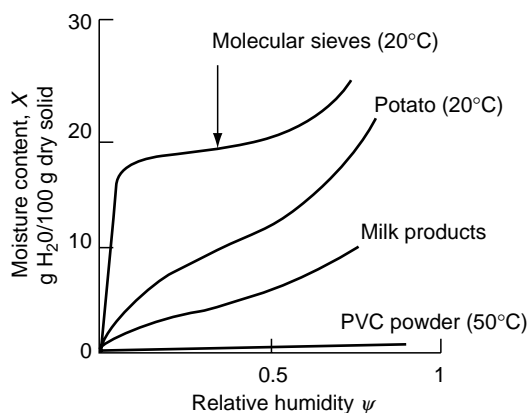


FIGURE 1.7 Shapes of sorption isotherms for materials of varying hygroscopicity.

Section B is a transitional region in which double and multiple layers are mainly formed. Capillary condensation could also have taken place. In section C the slope of the curve increases again, which is attributed mainly to increasing capillary condensation and swelling. The maximum hygroscopicity X_{\max} is achieved when the solid is in equilibrium with air saturated with moisture ($\psi = 1$).

1.3.2.1 Sorption–Desorption Hysteresis

The equilibrium moisture content of a product may be different depending on whether the product is wetted (sorption or absorption) or dried (desorption) (Figure 1.9). These differences are observed to varying degrees in almost all hygroscopic products.

One of the hypotheses used to explain hysteresis is to consider a pore connected to its surroundings by a small capillary [10]. During absorption, as the relative humidity rises, the capillary begins to fill while the pore is empty. Only when the partial pressure of the vapor in air is greater than the vapor pressure of the liquid in the capillary will the moisture move into the pore. Starting from saturation the pore is full of liquid. This fluid can only escape when the partial

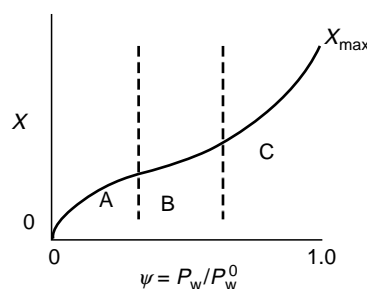


FIGURE 1.8 A typical isotherm (see text for explanation of areas within dashed lines).

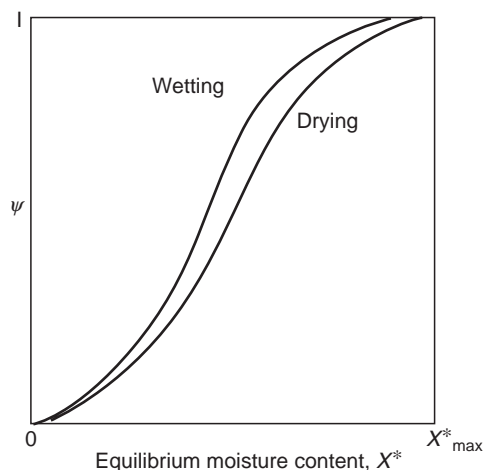


FIGURE 1.9 Wetting and drying isotherms for a typical hygroscopic solid.

pressure of the surrounding air falls below the vapor pressure of the liquid in the capillary. Since the system of pores has generally a large range of capillary diameters, it follows that differences between adsorption and desorption will be observed. This theory assumes that the pore is a rigid structure. This is not true for foods or synthetic materials, although these show hysteresis. The explanation is that contraction and swelling are superimposed on the drying and wetting processes, producing states of tension in the interior of the products and leading to varying equilibrium moisture contents depending on whether desorption or absorption is in progress.

1.3.2.2 Temperature Variations and Enthalpy of Binding

Moisture isotherms pertain to a particular temperature. However, the variation in equilibrium moisture content for small changes of temperature ($<10^{\circ}\text{C}$) is neglected [3]. To a first approximation, the temperature coefficient of the equilibrium moisture content is proportional to the moisture content at a given relative humidity:

$$\left(\frac{\partial X^*}{\partial T}\right)_{\psi} = -AX^* \quad (1.51)$$

The coefficient A lies between 0.005 and 0.01 per kelvin for relative humidities between 0.1 and 0.9 for such materials as natural and synthetic fibers, wood, and potatoes. A could be taken to increase linearly with ψ . So for $\psi = 0.5$ there is a 0.75% fall in moisture content for each degree kelvin rise in temperature. The extent of absorption–desorption hysteresis becomes smaller with increasing temperature.

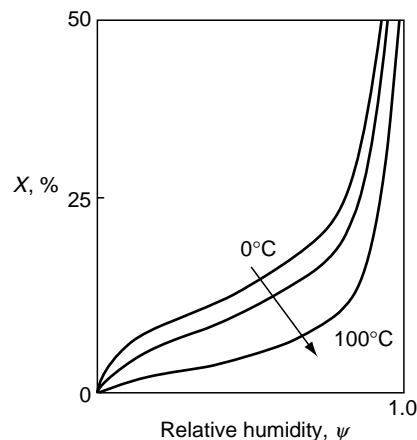


FIGURE 1.10 Sorption isotherms for potatoes.

Figure 1.10 shows moisture isotherms at various temperatures. The binding forces decrease with increasing temperature; that is, less moisture is absorbed at higher temperatures at the same relative humidity. Kessler [10] has shown that the slope of a plot of $\ln(P_w/P_w^0)$ versus $1/T$ at constant X (Figure 1.11) gives the enthalpy of binding. The variation of enthalpy of binding versus moisture content is shown in Figure 1.12. From the figure it is seen that in the region where monomolecular layers are formed, enthalpies of binding are very high.

1.3.3 DETERMINATION OF SORPTION ISOTHERMS [10]

The sorption isotherms are established experimentally starting mostly with dry products. The initial humidity of the air with which the product is in equilibrium should be brought to extremely low values using either concentrated sulfuric acid or phosphorus pentoxide, so that the moisture content of the product is close to zero at the beginning. The product is then exposed to successively greater humidities in a thermostatically controlled atmosphere. Sufficient time must be allowed for equilibrium between the air and solid to be attained. Using thin slices of the

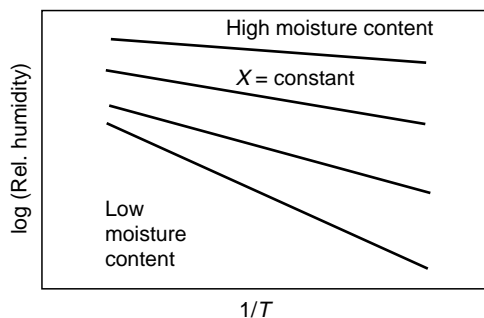


FIGURE 1.11 Determination of the heat of sorption from sorption isotherms.

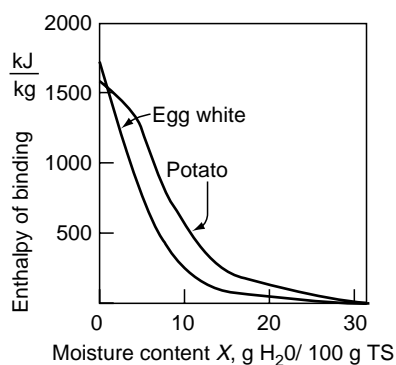


FIGURE 1.12 Enthalpy of sorption as a function of the hygroscopic moisture content. (Egg white data by Nemitz; potato data by Krischer.)

product, moving air and especially vacuum help to establish equilibrium quickly. This is especially important for foodstuffs: there is always the danger of spoilage. There are severe problems associated with the maintenance of constant humidity and temperature. These problems could be alleviated by using sulfuric acid–water mixtures and saturated salt solutions to obtain different relative humidities [10,13]. Figure 1.13 depicts the absorption isotherms of a range of food products. Further information on solid moisture characteristics, enthalpy of wetting, and sorption isotherms are available in Refs. [3,10].

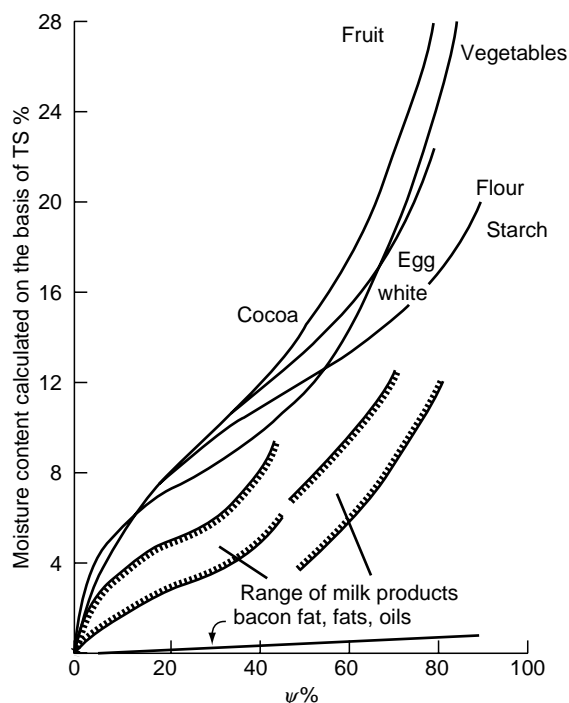


FIGURE 1.13 Range of sorption isotherms of various foods at room temperature.

1.4 MECHANISM OF DRYING

As mentioned above, moisture in a solid may be either unbound or bound. There are two methods of removing unbound moisture: evaporation and vaporization. Evaporation occurs when the vapor pressure of the moisture on the solid surface is equal to the atmospheric pressure. This is done by raising the temperature of the moisture to the boiling point. This kind of phenomenon occurs in roller dryers.

If the material dried is heat sensitive, then the temperature at which evaporation occurs, that is, the boiling point, could be lowered by lowering the pressure (vacuum evaporation). If the pressure is lowered below the triple point, then no liquid phase can exist and the moisture in the product is frozen. The addition of heat causes sublimation of ice directly to water vapor as in the case of freeze drying.

Second, in vaporization, drying is carried out by convection, that is, by passing warm air over the product. The air is cooled by the product, and moisture is transferred to the air by the product and carried away. In this case the saturation vapor pressure of the moisture over the solid is less than the atmospheric pressure.

A preliminary necessity to the selection of a suitable type of dryer and design and sizing thereof is the determination of the drying characteristics. Information also required are the solid-handling characteristics, solid moisture equilibrium, and material sensitivity to temperature, together with the limits of temperature attainable with the particular heat source. These will be considered later and in other sections of this book.

The drying behavior of solids can be characterized by measuring the moisture content loss as a function of time. The methods used are humidity difference, continuous weighing, and intermittent weighing. Descriptions of these methods are available in Refs. [3,13].

Figure 1.14 qualitatively depicts a typical drying rate curve of a hygroscopic product. Products that contain water behave differently on drying according to their moisture content. During the first stage of drying the drying rate is constant. The surface contains free moisture. Vaporization takes place from there, and some shrinkage might occur as the moisture surface is drawn back toward the solid surface. In this stage of drying the rate-controlling step is the diffusion of the water vapor across the air–moisture interface and the rate at which the surface for diffusion is removed. Toward the end of the constant rate period, moisture has to be transported from the inside of the solid to the surface by capillary forces and the drying rate may still be constant. When the

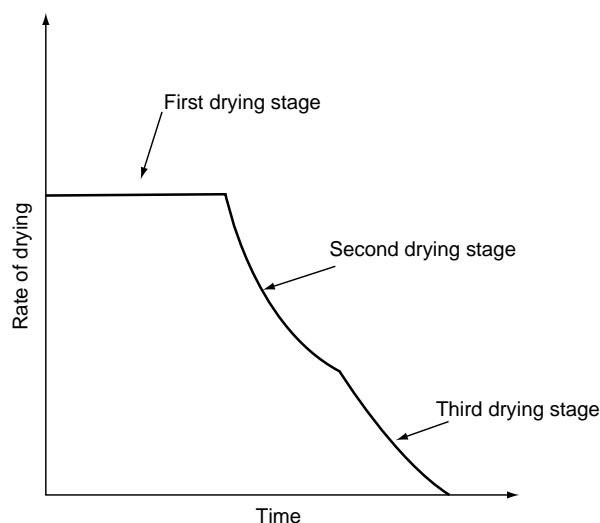


FIGURE 1.14 Typical rate-of-drying curve, constant drying conditions.

average moisture content has reached the critical moisture content X_{cr} , the surface film of moisture has been so reduced by evaporation that further drying causes dry spots to appear upon the surface. Since, however, the rate is computed with respect to the overall solid surface area, the drying rate falls even though the rate per unit wet solid surface area remains constant. This gives rise to the second drying stage or the first part of the falling rate period, the period of unsaturated surface drying. This stage proceeds until the surface film of liquid is entirely evaporated. This part of the curve may be missing entirely, or it may constitute the whole falling rate period.

On further drying (the second falling rate period or the third drying stage), the rate at which moisture may move through the solid as a result of concentration gradients between the deeper parts and the surface is the controlling step. The heat transmission now consists of heat transfer to the surface and heat conduction in the product. Since the average depth of the moisture level increases progressively and the heat conductivity of the dry external zones is very small, the drying rate is increasingly influenced by the heat conduction. However, if the dry product has a relatively high bulk density and a small cavity volume with very small pores, drying is determined not so much by heat conduction but by a rather high resistance to diffusion within the product. The drying rate is controlled by diffusion of moisture from the inside to the surface and then mass transfer from the surface. During this stage some of the moisture bound by sorption is removed. As the moisture concentration is lowered by drying, the rate of internal movement of

moisture decreases. The rate of drying falls even more rapidly than before and continues until the moisture content falls down to the equilibrium value X^* for the prevailing air humidity and then drying stops. The transition from one drying stage to another is not sharp, as indicated in Figure 1.14.

In actual practice, the original feedstock may have a high moisture content and the product may be required to have a high residual moisture content so that all the drying may occur in the constant rate period. In most cases however both phenomena exist, and for slow-drying materials most of the drying may occur in the falling rate period. As mentioned earlier, in the constant rate period the rate of drying is determined by the rate of evaporation. When all the exposed surface of the solid ceases to be wetted, vapor movement by diffusion and capillarity from within the solid to the surface are the rate-controlling steps. Whenever considerable shrinkage occurs, as in the drying of timber, pressure gradients are set up within the solid and these may assume importance. In this case, as in the case of materials that “caseharden,” that is, form a hard impermeable skin, it is essential to retard evaporation and bring it in step with the rate of moisture movement from the interior. This could be achieved by increasing the relative humidity of the drying air. With solids, in which the initial moisture content is relatively low and the final moisture content required is extremely low, the falling rate period becomes important. Dryness times are long. Air velocities will be important only to the extent to which they enhance heat transfer rates. Air temperature, humidity, material thickness, and bed depth all become important. When the rate of diffusion is the controlling factor, particularly when long drying periods are required to attain low moisture contents, the rate of drying during the falling rate period varies as the square of the material thickness, which indicates the desirability of granulating the feedstock using agitation or using thin layers in case of cross-flow tray dryers. Thus the drying characteristics of the solid are extremely important in dryer design.

1.4.1 CHARACTERISTIC DRYING RATE CURVE [14]

When the drying rate curves are determined over a range of conditions for a given solid, the curves appear to be geometrically similar and are simply a function of the extent to which drying has occurred. If these curves were normalized with respect to the initial drying rate and average moisture content, then all the curves could often be approximated to a single curve, “characteristic” of a particular substance. This is the characteristic drying curve. The normalized variables, the characteristic drying rate

f and the characteristic moisture content ϕ , are defined as follows:

$$f = \frac{N_v}{N_w}$$

and

$$\phi = \frac{\bar{X} - X^*}{X_{cr} - X^*}$$

where N_v is the rate of drying for a unit surface, N_w is the rate when the body is fully saturated or the initial drying rate, \bar{X} is the average moisture content in the body, X_{cr} is the corresponding critical point value, and X^* is the equilibrium moisture content.

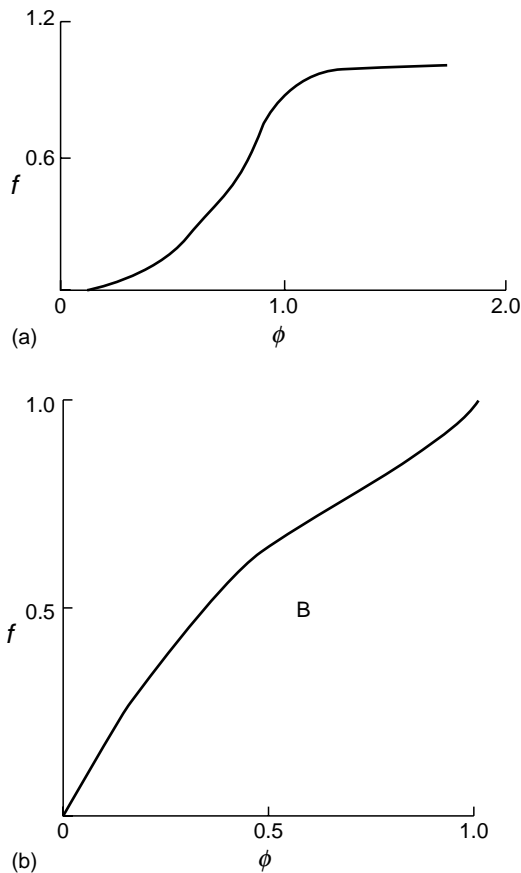


FIGURE 1.15 Experimental drying rates for (a) molecular sieves 13-X particles with $X_0 = 0.27$, diameter = 2.2 mm, air velocity = 4.4 m/s, $T_G \cong 36.5\text{--}97^\circ\text{C}$; (b) silica-gel particles with diameter = 3.0 mm, air velocity = 1 m/s, $T_G = 54\text{--}68^\circ\text{C}$, $T_w = 25\text{--}29^\circ\text{C}$.

If a solid's drying behavior is to be described by the characteristic curve, then its properties must satisfy the following two criteria:

1. The critical moisture content \bar{X}_{cr} is invariant and independent of initial moisture content and external conditions
2. All drying curves for a specific substance are geometrically similar so that the shape of the curve is unique and independent of external conditions

These criteria are restrictive, and it is quite unlikely that any solid will satisfy them over an exhaustive range of conditions; however, the concept is widely used and often utilized for interpolation and prediction of dryer performance [3,17]. The use of the mean moisture content as an index of the degree of drying contains the implicit assumption that the extent of drying at a mean moisture content will also depend on the relative extensiveness of the exposed surface per unit volume of material. Thus, similar drying behavior may be expected only in the case of materials that are unchanged in form. A typical characteristic drying curve is shown in Figure 1.15.

Further information on the characteristic drying curve, extrapolation procedures used, and the theoretical developments in examining the range of validity of the characteristic drying rate model are available [3,13,25]. The various types of characteristic drying curves have been depicted schematically in Figure 1.16.

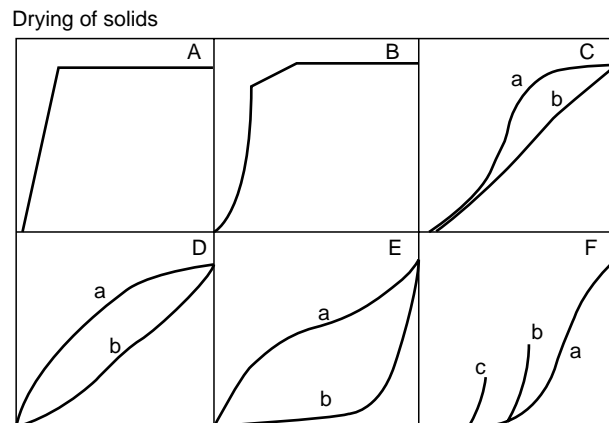


FIGURE 1.16 Examples of normalized drying rate curves for different types of media: (A) sand, clay, silica gel, paper pulp, leather; (B) sand, plastic-clay mix, silica-brick mix, ceramic plate, leather; (C) [a] fir wood and [b] Cyprus wood; (D) [a] paper, wool, and [b] potatoes, tapioca tuber, rice flour; (E) [a] rye bread, yeast, and [b] butter and margarine; (F) [a] wheat corns, [b] and [c] represent curves at lower values of initial moisture.

1.5 CLASSIFICATION AND SELECTION OF DRYERS

With a very few exceptions, most products from today's industry undergo drying at some stage or another. A product must be suitable for either subsequent processing or sale. Materials need to have a particular moisture content for processing, molding, or pelleting. Powders must be dried to suitable low moisture contents for satisfactory packaging. Whenever products are heated to high temperatures, as in ceramic and metallurgical processes, predrying at lower temperatures ahead of firing kilns is advantageous for energy savings. Cost of transport (as in the case of coal) depends on the moisture content of the product, and a balance must be struck between the cost of conveying and the cost of drying. Excessive drying is wasteful; not only is more heat, that is, expense, involved than is necessary, but often overdrying results in a degraded product, as in the case of paper and timber. Consideration must be given to methods involved in energy savings in dryers [15]. Examples of products and the types of precautions that need to be taken during drying are highlighted. Thermal drying is an essential stage in the manufacture of colors and dyes. Many inorganic colors and most organic dyes are heat sensitive; the time-temperature effect in drying may be critical in arriving at a correct shade of color or in the elimination of thermal degradation. Drying is normally accomplished at low temperatures and in the absence of air. The most widely used dryer is the recirculation type truck and tray compartment dryer. Most pharmaceuticals and fine chemicals require drying before packaging. Large turbo-tray and through circulation dryers are employed. Excessively heat-sensitive products, such as antibiotics and blood plasma, call for special treatment, like freeze drying or high vacuum tray drying. Continuous rotary dryers are usually used to handle large tonnages of natural ores, minerals, and heavy chemicals.

The largest demand for drying equipment is for the continuous drying of paper, which is done on cylinder or "can" dryers. The temperature and humidity conditions are important to the consistency of the paper. Thermal drying is essential in the foodstuffs and agricultural fields. Spray drying and freeze drying are also widely used. In the ceramic industry, drying is a vital operation. Great care must be exercised because of the considerable shrinkage that occurs in drying. Thus control of humidity is important. Drying is also widely used in the textile industry. The need for product quality puts grave constraints on the dryer chosen and dryer operation [13]. Quality depends on the end use of the product. For many bulk chemicals, handling considerations determine moisture

content requirements. For foodstuffs, flavor retention, palatability, and rehydration properties are important. Timber must retain its strength and decorative properties after drying.

The choice of end moisture content is largely dictated by storage and stability requirements. The end moisture content determines the drying time and conditions required for drying. Overdrying should be avoided. Internal moisture gradients within particles and interparticle moisture content variation are important. Temperature restrictions may arise because of degradation, phase changes, discoloration and staining, flammability of dust, and other factors. Thermal sensitivity fixes the maximum temperature to which the substance can be exposed for the drying time. In spray and pneumatic dryers, the retention time of a few seconds permits drying heat-sensitive materials at higher temperatures. Many hygroscopic materials shrink on drying. The extent of shrinkage is linearly related to the moisture content change below the hygroscopic limit. Case hardening and tensile cracking pose problems. Details on how some of these problems are overcome by a compromise between energy efficiency and drying time are available [13].

The first subdivision is based on methods of heat transfer, namely, (a) conduction heating, (b) convection heating, (c) radiant heating, and (d) dielectric heating. Freeze drying is classified as a special case of conduction heating. The next subdivision is the type of drying vessel: tray, rotating drum, fluidized bed, pneumatic, or spray. Dryer classification based on physical form of the feed has also been done. These are indicated in Ref. [16].

Sloan [18] has identified some 20 types of dryers and classified them according to whether they are batch or continuous, conduct heat exchange through direct contact with gases or by heat exchange through vessel walls, and according to the motion of the equipment. Such a classification, though helpful in distinguishing and describing discrete systems, does not go far in relating the discrete systems to the process problems they are supposed to handle. McCormick [19] has tried to tackle the problem from the user's point of view. A total of 19 types of dryers were classified according to how well they handle different materials.

Schlünder [23] has given a classification of dryers that encompasses the physical state of the product as well as the dwell time of the product in the dryer. For very short drying times (<1 min), flash, spray, or drum dryers are used. For very long drying times (>1 h), only tunnel, truck, or conveyor dryers are appropriate. Most dryers operate in the intermediate range, for which a very wide assortment of dryers is available [21].

TABLE 1.1
Dryer Selection versus Feedstock Form

Nature of Feed	Liquids			Cakes		Free-Flowing Solids					Formed Solids
	Solution	Slurry	Pastes	Centrifuge	Filter	Powder	Granule	Fragile Crystal	Pellet	Fiber	
Convection Dryers											
Belt conveyer dryer							×	×	×	×	×
Flash dryer				×	×	×	×			×	
Fluid bed dryer	×	×		×	×	×	×		×		
Rotary dryer				×	×	×	×		×	×	
Spray dryer	×	×	×								
Tray dryer (batch)				×	×	×	×	×	×	×	×
Tray dryer (continuous)				×	×	×	×	×	×	×	
Conduction Dryers											
Drum dryer	×	×	×								
Steam jacket rotary dryer				×	×	×	×		×	×	
Steam tube rotary dryer				×	×	×	×		×	×	
Tray dryer (batch)				×	×	×	×	×	×	×	×
Tray dryer (continuous)				×	×	×	×	×	×	×	

Table 1.1 gives a summary of the type of dryer versus the type of feedstock, which may be a slurry, paste, filter cake, powder, granules, crystal, pellet, or fibrous or shaped material. Since thermal sensitivity as well as efficiency and dryer size depend to a major extent on the thermal conditions the product is exposed to within the dryer, Table 1.2 is presented to classify convection and conduction dryers on this basis. Such information is often helpful in narrowing down the choice of dryers.

TABLE 1.2
Solids' Exposures to Heat Conditions

Dryers	Typical Residence Time Within Dryer				
	0–10 (s)	10–30 (s)	5–10 (min)	10–60 (min)	1–6 (h)
<i>Convection</i>					
Belt conveyor dryer				×	
Flash dryer	×				
Fluid bed dryer				×	
Rotary dryer				×	
Spray dryer		×			
Tray dryer (batch)					×
Tray dryer (continuous)				×	
<i>Conduction</i>					
Drum dryer		×			
Steam jacket rotary dryer				×	
Steam tube rotary dryer				×	
Tray dryer (batch)					×
Tray dryer (continuous)				×	

Keey [3,25] has noted three principal factors that could be utilized in classifying dryers:

1. Manner in which heat is supplied to the material
2. Temperature and pressure of operation (high, medium, or low temperature; atmospheric or vacuum drying)
3. Manner in which the material is handled within the dryer

Further subclassification is of course possible but generally unnecessary. For example, a given dryer may be operated in batchwise or continuous mode.

1.5.1 HEATING METHODS

1.5.1.1 Convection

Convection is possibly the most common mode of drying particulate or sheet-form or pasty solids. Heat is supplied by heated air or gas flowing over the surface of the solid. Heat for evaporation is supplied by convection to the exposed surface of the material and the evaporated moisture carried away by the drying medium. Air (most common), inert gas (such as N₂ for drying solids wet with organic solvent), direct combustion gases, or superheated steam (or solvent vapor) can be used in convective drying systems.

Such dryers are also called direct dryers. In the initial constant rate drying period (drying in which surface moisture is removed), the solid surface takes on the wet bulb temperature corresponding to the air

temperature and humidity conditions at the same location. In the falling rate period the solids' temperature approaches the dry bulb temperature of the medium. These factors must be considered when drying heat-sensitive solids.

When drying with superheated vapors, the solids' temperature corresponds to the saturation temperature at the operating pressure, for example, 100°C for steam at 1 atm. For solids susceptible to oxidation or denaturation, for example, in the presence of oxygen, the heat sensitivity can be quite different in a steam environment. Product quality may differ as well. This can only be ascertained through laboratory tests.

Examples of convective (direct) dryers are air suspension dryers, such as fluid bed, flash, rotary, or spray dryers; air impingement dryers for paper or pulp; packed bed or through dryers; and conveyor-truck-tunnel dryers.

1.5.1.2 Conduction

Conduction or indirect dryers are more appropriate for thin products or for very wet solids. Heat for evaporation is supplied through heated surfaces (stationary or moving) placed within the dryer to support, convey, or confine the solids. The evaporated moisture is carried away by vacuum operation or by a stream of gas that is mainly a carrier of moisture. Vacuum operation is recommended for heat-sensitive solids. Because the enthalpy lost with the drying air in convective dryers is large, their thermal efficiency tends to be low. For conduction dryers the thermal efficiency is higher. Paddle dryers for drying of pastes, rotary dryers with internal steam tubes, and drum dryers for drying thin slurries are examples of indirect dryers.

A more efficient dryer can be designed for some operations that combines advantages of both direct and indirect heating, for example, a fluid bed dryer with immersed heating tubes or coils for drying of heat-sensitive polymer or resin pellets. Such a dryer can be only one third the size of a purely convective fluid bed dryer for the same duty.

It is noteworthy that sometimes one can operate the same apparatus in direct, indirect, or combined modes. For example, a vibrated fluid bed dryer can be purely convective (e.g., drying of tea), purely conductive (e.g., vacuum drying of pharmaceutical granules), or combined direct-indirect (e.g., drying of pulverized coal with immersed heating tubes). The drying medium could be steam for such products as coal.

1.5.1.3 Radiation

Various sources of electromagnetic radiation with wavelengths ranging from the solar spectrum to

microwave (0.2 m–0.2 μ m). Solar radiation barely penetrates beyond the skin of the material, which absorbs only a part of the incident radiation depending on its wavelength. Infrared radiation is often used in drying coatings, thin sheets, and films, for example (4–8 μ m band). Although most moist materials are poor conductors of 50–60 Hz current, the impedance falls dramatically at RF; such radiation can be used to heat the solid volumetrically, thus reducing internal resistance to heat transfer. Energy is absorbed selectively by the water molecules: as the product gets drier less energy is used. The capital as well as operating costs are high, and so these techniques are useful for drying high unit value products or for final correction of moisture profile wherein only small quantities of hard-to-get moisture are removed, as in moisture profiling of paper using RF heating. Combined mode drying with convection (e.g., infrared plus air jets or microwave with impingement for drying of sheet-form foodstuffs) is also commercially feasible.

1.5.2 TEMPERATURE AND PRESSURE OF OPERATION

Most dryers are operated at near atmospheric pressures. A slight positive pressure avoids in-leakage from outside, which may not be permissible in some cases. If no leakage is permitted to the outside, then a slight negative pressure is used.

Vacuum operation is expensive and is recommended only if the product must be dried at low temperatures or in the absence of oxygen or has flavors that are generated at medium- or high-temperature operation. High-temperature operation tends to be more efficient since lower gas flow rates and smaller equipment may be used for a given evaporation duty. Availability of low-temperature waste heat or energy from solar collectors may dictate the choice of a lower temperature operation. These dryers will then be large in size.

Freeze drying is a special case of drying under vacuum at a temperature below the triple point of water; here water (ice) sublimates directly into water vapor. Although the heat required for sublimation is severalfold lower than for evaporation, vacuum operation is expensive. Freeze drying of coffee, for example, costs two to three times more than spray drying. On the other hand, the product quality and flavor retention are better.

1.5.3 CONVEYING OF MATERIAL IN DRYER

Handling the material to be dried is of course one of the key considerations in dryer selection. This is best illustrated in [Table 1.3](#). In some cases the material may be treated or preformed to make it suitable for

TABLE 1.3
Capacity and Energy Consumption for Selected Dryers

Method	Typical Dryer	Typical Materials
Material not conveyed	Tray dryer	Wide range of pastes, granules
Material falls by gravity	Rotary dryer	Free-flowing granules
Material conveyed mechanically	Screw-conveyor, paddle	Wet sludges, pastes
Transported on trucks	Tunnel dryer	Wide range of materials
Sheet-form materials, supported on rolls	Cylinder dryers	Paper, textiles, pulp
Conveyed on bands	Band, conveyor dryer	Wide range of solids (pellets, grains)
Material suspended in air	Fluid bed, flash	Free-flowing granules
Slurries or solutions atomized in air	Spray dryer	Milk, coffee, etc.

Note: Most dryers may operate continuously, semicontinuously, or batchwise. Labor costs are high for tray and tunnel dryers.

handling in a particular dryer, for example, reslurrying of filter cake to make it pumpable for atomization and spray drying or pelletizing pasty materials. This, of course, costs extra and must be considered in an overall evaluation. The final product characteristics and quality requirements also govern the choice of dryer. In some cases a combination of two or more different types of dryers may be the optimal strategy if the product-handling properties change significantly as it dries or if its heat sensitivity changes during the process of drying.

It should be pointed out that many new dryers cannot conveniently fit the classification suggested earlier. For example, a pulsed combustion dryer for pasty solids or waste sludge, the Remaflam process for drying textiles by controlled combustion of solvent (alcohol) on the wet fabric itself, and vibrated bed drying of pastes cannot be placed under any one single category of dryers. Such a “coarse” classification is still of interest in that it allows one to “home in” on a limited number of possible dryers, which can then be evaluated in depth. One must look very carefully into some of the newer and novel dryers although they are not even mentioned in most textbooks or handbooks [22]. Many of them have the potential to supplant some of the age-old drying technologies, at least in some industrial applications.

Dittman [20] has proposed a structured classification of dryers according to two general classes and five subclasses. The general classes are adiabatic or nonadiabatic dryers. Adiabatic dryers are further subclassified according to whether drying gases pass through the material (for permeable solids or beds of solids) or across the surface. Nonadiabatic dryers are categorized according to the mode of heat supply, such as heat applied through a heat exchange surface or direct radiation, and according to the mode of

moisture carryover, for example, moisture removal by vacuum or by a carrier gas.

Many difficulties are encountered in the selection of dryers. These mainly arise because there is no standard set of systematized laboratory tests using standardized apparatus to provide key data on the drying characteristics of materials. The real mechanics of liquid removal from the solid is not really understood, nor is the operation of many dryers. A systematic comprehensive classification of existing dryers has yet to be agreed upon. There is also a lack of a reliable procedure for scaling up laboratory data and even pilot-plant data for some types of dryers.

In spite of the above-mentioned lacunae, dryers have still to be selected and some prior information is required to facilitate this job. This includes (a) flow sheet quantities, such as dry solid quantity, total liquid to be removed, and the source of the wet material; (b) batch or continuous feed physical characteristics, such as source of feed, presence of any previous dewatering stage, like filtration, mechanical pressing, or centrifuging, method of supplying material to the dryer, particle size distribution in the wet feed, physical characteristics and handleability, and abrasive properties of wet and dry materials; (c) feed chemical properties, such as toxicity, odor problems, whether the material can be dried with hot combustion gases containing carbon dioxide, sulfur dioxide, some nitrogen oxides, and traces of partially burnt hydrocarbon, fire and explosion hazards, temperature limitations, temperatures of relevant phase changes, and corrosive properties; (d) dry product specification and properties, such as moisture content, removal of solvent odor, particle size distribution, bulk density, maximum percentage of impurities, desired granular or crystalline form, flow properties, and temperature to which the dried product must be cooled before storage; and (e) drying data obtained from a pilot

plant or laboratory as well as previous experience of the drying performance of similar materials in a full-scale plant. Information on solvent recovery, product loss, and site conditions would be an added bonus.

The best method of selection involves using past experience. One of the preliminary ways of selecting dryers is based on the nature of the feed [16]. There is little difficulty in handling liquid feeds, and the choice of the equipment is normally limited to (a) spray dryer, (b) drum dryer, atmospheric or vacuum, and (c) agitated batch vacuum dryer. Other considerations that might influence the final choice are the need for small product losses and a clean plant, solvent recovery or the need to use an inert atmosphere in which an agitated vacuum dryer is preferred, and the temperature sensitivity of the material. The agitated vacuum dryer has a long residence time, the through circulation dryer, a moderate temperature and moderate residence time. The drum dryer can have a high mean temperature with a short contact time; the spray dryer has a short contact time with a wide range of operating temperatures. The above-mentioned selection is applicable to pumpable suspensions of fine solids, excluding pastes.

For the continuous drying of pastes and sludges, in which the solids are in a finely divided state, dust problems are a major consideration. However, the choice between batch and continuous operation is difficult. The batch dryers normally used are tray atmospheric or vacuum, agitated batch atmospheric or vacuum, and rotary atmospheric or vacuum. The vacuum operation is preferable in cases of solvent recovery, fire, or toxic hazards or when temperature limitations are necessary. Dryers used in continuous operation are (a) spray, where atomization itself poses a considerable problem; (b) fluidized bed, where dispersion of the feed in a deep bed is difficult; (c) continuous band circulation, suitable if dust-free product is required; (d) pneumatic, requires mixing of feed with dry product to facilitate dispersion of wet solid in the gas entering the dryer; and (e) continuous rotary, direct or indirect; here too blending of wet feed with dry product is necessary to facilitate handling. If the feed contains fine particles, the indirect mode of heat transfer is normally preferred. In case of free-blowing wet powders (particle size less than 300 μm) all the dryers used for pastes and sludges could be used with the inclusion of the vertical rotating shelf dryer. For granular crystalline solids with particle sizes greater than 300 μm , the direct rotary dryer is commonly used. With this type of dryer crystal breakage is a problem that can be overcome by proper flight design. For particles larger than 25 mesh, a through circulation dryer using a moving band or a vibrating screen could be used. Fibrous

solids hold a considerable amount of water but dry quite easily. These materials are often temperature sensitive due to their high specific surface, and care should be taken to keep the air temperature down. Through circulation tests at various temperatures should establish the need, if any, to avoid overheating. Apart from this, fibrous materials could be treated in a similar way to any of the solids mentioned above.

The final selection of the dryer will usually represent a compromise between the total cost, operating cost, quality of the product, safety consideration, and convenience of installation. It is always wise, in case of uncertainty, to run preliminary tests to ascertain both design and operating data and also the suitability of the dryer for the particular operation. On certain types of dryers, full-scale tests are the only way of establishing reliable design and operating data, but in some cases, such as through circulation dryers, there are reliable techniques for establishing reliable data from laboratory tests. Details on test techniques and as to when tests are needed are available in Refs. [13,16]. It should be noted that specialized drying techniques, such as vacuum freeze drying, microwave freeze drying, and dielectric or infrared heating or drying, have not been considered here. Details of some of these specialized techniques are given elsewhere in this handbook.

1.6 EFFECT OF ENERGY COSTS, SAFETY, AND ENVIRONMENTAL FACTORS ON DRYER SELECTION

Escalating energy costs and increasingly stringent legislation on pollution, working conditions, and safety have a direct bearing on the design as well as selection of industrial dryers. Lang [27] has discussed the effects of these factors on design of suspension dryers for particulates, such as spray, flash, and fluidized bed dryers. These factors must be considered during the phase of selection of dryers. In some cases a choice between competing drying systems exists; in others one must incorporate these factors at the design stage.

For a given dryer system (including preprocessing, such as mechanical dewatering, centrifugation, evaporation, and pressing, and postprocessing, such as product collection, cooling, agglomeration, granulation, and scrubbing), in general several energy-saving flow sheets may be devised, including gas recycle, closed cycle operation, self-inertization, multistage drying, and exhaust incineration. Areas of conflict may exist between legal requirements, hygienic operation, and energy efficiency. Lang gives the following possible scenarios of conflict:

1. Explosion vents could be a hygiene problem.
2. Dust in recycling streams fouls heat exchanger surfaces or causes difficulties in direct combustion systems.
3. Thermal expansion joints or fire-extinguishing equipment can cause product buildup and hence a fire hazard.
4. High product collection efficiency for particulate dryers means high pressure drop and increased fan noise.

Note that unnecessarily stringent product specifications can cause significant increase in dryer costs, both capital and operating.

In selecting energy-saving drying systems, it is important to note the following (mainly for particulate drying, but some factors are of general applicability):

1. When handling a thermally sensitive product, recycled exhaust must be totally free of product if the stream is to pass through or near a burner.
2. Recycling increases humidity level in drying, which may increase the equilibrium moisture content to unacceptable levels in some cases.
3. To avoid passing dust in recycled gas through air heaters, if fresh makeup air is heated and mixed with recycled gas, to obtain high mixture temperature (say, 400°C), the fresh gas must be heated to a temperature too high for simple materials of construction. Exotic metals or refractories are needed, which can cause product contamination or a source of ignition if it reaches high enough temperatures.
4. In multiple-stage drying, heat economy requires that the first-stage drying give a partially dried product, which is sometimes too sticky to handle.

A drying installation may cause air pollution by emission of dust and gases. Even plumes of clean water vapor are unacceptable in some areas. Particulates below the range 20–50 mg/nm³ of exhaust air are a common requirement. High-efficiency dust collection is essential. It is important to operate the dryer under conditions conducive to production of coarse product. On the other hand, larger products take longer to dry. Cyclones, bag filters, scrubbers, and electrostatic precipitators are commonly used for particle collection and gas cleaning on particulate material drying of materials in other forms, such as pulp sheets. For the removal of noxious gaseous pollutants, one may resort to absorption, adsorption, or incineration; the last operation is becoming more common.

Although rare, care must be taken in drying of airborne materials that can catch fire. Reduction in

oxygen content (by recycle) can suppress explosion hazard. Should explosion occur, suitable explosion vents must be included to avoid buildup of excessive pressure in the system. Elimination of sources of ignition is not acceptable as adequate assurance against fire or explosion hazard. When an explosion risk exists, buildup of product in the dryer or collector must be avoided. For example, venting doors on the roof of large-volume spray dryers must be flush with the interior surface to prevent any buildup of product. It is often cheaper to design a partially or fully inertized drying system than to design larger dryer chambers to withstand high internal pressures (5–6 psig).

Finally, local legislation about noise levels must be considered at the selection and design state. Depending on the stringency of noise requirements, the cost of acoustic treatment can be as much as 20% of the total system cost. For air suspension dryers, the fan is the main noise generator. Other sources, such as pumps, gearboxes, compressors, atomization equipment, burners, and mixers, also contribute to noise. Low fan noise requires a low-pressure drop in the system, which is in conflict with the high-pressure drop required for higher collection efficiency.

A simple direct-fired dryer with once through air-flow heating, especially with gas, can be achieved with little noise generation. On the other hand, a more efficient air recirculation unit often requires high-noise burners and ancillaries.

The aforementioned discussion is intended to provide the specifier of drying equipment some practical factors that should be considered at the stage of selection of the dryer system. Rarely, if ever, is it possible to select a dryer that meets all criteria. In most cases, however, one can modify the dryer system design or operation to meet all the essential specifications of the user.

Menon and Mujumdar [15] have indicated some energy-saving measures involved in conditioning of the feed, dryer design, and heat recovery from the exhaust stream, including the use of heat pumps. Some novel techniques, like displacement drying, steam drying, drying using superheated steam, RF drying, press drying, and combined impingement and through drying, have been proposed.

Although prior experience is a guide commonly used in specifying dryers, it is important to recognize that earlier dryers were often specified in times when energy costs were minimal and requirements of product quality and production rates were different. There is also a variation in energy costs from one geographic location to another and certainly from one country to another. It is strongly recommended that the process engineer make a selection of dryer from

TABLE 1.4
Capacity and Energy Consumption for Selected Dryers

Dryer Type	Typical Evaporation Capacity (kg H ₂ O/h m ² or kg H ₂ O/h m ³)	Typical Energy Consumption (kJ/kg of H ₂ O Evaporated)
Tunnel dryer	—	5,500–6,000
Band dryer	—	4,000–6,000
Impingement dryer	50/m ²	5,000–7,000
Rotary dryer	30–80/m ³	4,600–9,200
Fluid bed dryer		4,000–6,000
Flash dryer	5–100/m ³ (depends on particle size)	4,500–9,000
Spray dryer	1–30/m ³	4,500–11,500
Drum dryer (for pastes)	6–20/m ²	3,200–6,500

Note: Figures are only approximate and are based on current practice. Better results can often be obtained by optimizing operating conditions and using advanced technology to modify the earlier designs.

current conditions and geographic location while taking into account future expected trends. In many instances the most widely used dryers for a specific product have been found to be poor choices under prevailing conditions. Table 1.4 gives a summary of typical drying capacities and energy consumption in existing common industrial dryers.

1.7 DESIGN OF DRYERS

The engineer concerned with process design has to choose for a given dryer those conditions that enable the specified properties of the product to be obtained. The performance characteristics of alternative systems should also be assessed before the final choice of a dryer type. Almost always some small-scale tests are needed to determine the material's drying characteristics required to predict the way in which the raw material would behave in the actual unit. A flowchart illustrating the various steps needed to design a dryer is shown in Ref. [3].

For a preliminary estimate of dryer size, Toei [24] gives a simple method based on data obtained from operating industrial dryers. For convection dryers, the rate of heat transfer (kcal/h) is given by

$$q = (ha)(V)(t - t_m), \text{ for batch type}$$

and

$$q = (ha)(V)(t - t_m)_{l_m}, \text{ for continuous type}$$

For conduction dryers,

$$q = UA(t_k - t_m)$$

In this equation, t_m is the product temperature (K); t is the inlet temperature (K); $(t - t_m)_{l_m}$ is the logarithmic mean (K) of the temperature differences between the hot air and the product at the inlet and outlet, respectively; ha is the volumetric heat transfer coefficient (kcal/sec K m³); U is the overall heat transfer coefficient (kcal/s K m²); A is the heating area in contact with the product (m²); and t_k is the temperature of the heat source (K).

Table 1.4 is excerpted from Ref. [24]. Here the volume used to define ha includes void volumes, for example, above and below a fluid bed dryer. Thus it is the overall volume and is subject to significant variation. It is also dependent on the critical moisture of the solids. Note the units when using this table. An estimate must be made *a priori* about the drop in air temperature across the dryer and rise of product temperature in order to use Table 1.5 for a very rough sizing of a dryer.

Extensive work has been done in developing theories of drying, such as the theory of simultaneous transport, theories involving flow through porous media, and simplified models, like the wetted surface model and the receding-plane model. The characteristic drying rate curve has also found use in design. Details of these theories are available in Refs. [3,13,17,26].

In designing a dryer, basically one or more of the following sources of information are used: (a) information obtained from customers, (b) previous experience in the form of files on dryers sold and tended, or (c) pilot-plant tests and bench-scale tests. For design a computer is usually used; the input parameters are based on hard data and partly on experience. Pilot-plant tests ensure that the material can be processed in the desired manner. However, the scale-up procedures are by no means straightforward. The theory in the computer program is restricted to heat and mass balances. The design is based on drying times guessed by the designer, who may be guided by drying data obtained in bench-scale and pilot-plant tests. Design procedures for various dryers have been outlined in Refs. [7,16]. Factors affecting product quality should be borne in mind before deciding on a dryer. (For discussion of some new drying technologies, see Ref. [28].)

1.8 GUIDELINES FOR DRYER SELECTION

In view of the enormous choices of dryer types one could possibly deploy for most products, selection of

TABLE 1.5
Approximate Values of ha (kcal/h °C m³) for Various Dryer Types

Type	ha	$(t - t_m)_{lm}$ (°C)	Inlet Hot Air Temperature (°C)
<i>Convection</i>			
Rotary	100–200	Countercurrent: 80–150 Cocurrent: 100–180	200–600 300–600
Flash	2000–6000	Parallel flow only: 100–180	400–600
Fluid bed	2000–6000	50–150	100–600
Spray	20–80 (large five)	Counterflow: 80–90 Cocurrent: 70–170	200–300 200–450
Tunnel	200–300	Counterflow: 30–60 Cocurrent: 50–70	100–200 100–200
Jet flow	$h = 100$ –150	30–80	60–150
<i>Conduction</i>			
Drum	U (kcal/h °C m ²)	$t_k - t_m$ (°C)	
Agitated through rotary with steam tubes, etc.	100–200 60–130 (smaller for sticky solids)	50–80 50–100	

the best type is a challenging task that should not be taken lightly nor should it be left entirely to dryer vendors who typically specialize in only a few types of dryers. The user must take a proactive role and employ vendors' experience and bench-scale or pilot-scale facilities to obtain data, which can be assessed for a comparative evaluation of several options. A wrong dryer for a given application is still a poor dryer, regardless of how well it is designed. Note that minor changes in composition or physical properties of a given product can influence its drying characteristics, handling properties, etc., leading to a different product and in some cases severe blockages in the dryer itself. Tests should be carried out with the "real" feed material and not a "simulated" one where feasible.

Although here we will focus only on the selection of the dryer, it is very important to note that in practice one must select and specify a drying system which includes predrying stages (e.g., mechanical dewatering, evaporation, preconditioning of feed by solids backmixing, dilution or pelletization, and feeding) as well as the postdrying stages of exhaust gas cleaning, product collection, partial recirculation of exhausts, cooling of product, coating of product, agglomeration, etc. The optimal cost-effective choice of dryer will depend, in some cases significantly, on these stages. For example, a hard pasty feedstock can be diluted to a pumpable slurry, atomized and dried in a spray dryer to produce a powder, or it may be pelletized and dried in a fluid bed or in a through circulation dryer, or dried as is in a rotary or fluid bed unit. Also, in some cases, it may be necessary to examine the entire flow sheet to see if the drying problem can be simplified or even eliminated.

Typically, nonthermal dewatering is an order-of-magnitude less expensive than evaporation which, in turn, is manyfold energy efficient than thermal drying. Demands on product quality may not always permit one to select the least expensive option based solely on heat and mass transfer considerations, however. Often, product quality requirements have overriding influence on the selection process. For high-value products (e.g., pharmaceuticals and certain high-value food-stuffs) the selection of dryers depends mainly on the value of the dried products since cost of drying becomes a very small fraction of the sale price of the product. On the other hand for very low value products or even no value products (e.g., waste sludges) the choice of drying system depends entirely on the cost of drying, so the lowest cost system is selected for such applications.

As a minimum, the following quantitative information is necessary to arrive at a suitable dryer:

- Dryer throughput; mode of feedstock production (batch/continuous)
- Physical, chemical, and biochemical properties of the wet feed as well as desired product specifications; expected variability in feed characteristics
- Upstream and downstream processing operations
- Moisture content of the feed and product
- Drying kinetics; moist solid sorption isotherms
- Quality parameters (physical, chemical, biochemical)
- Safety aspects, e.g., fire hazard and explosion hazards, toxicity
- Value of the product
- Need for automatic control

- Toxicological properties of the product
- Turndown ratio, flexibility in capacity requirements
- Type and cost of fuel, cost of electricity
- Environmental regulations
- Space in plant

For high-value products like pharmaceuticals, certain foods and advanced materials, quality considerations override other considerations since the cost of drying is unimportant. Throughputs of such products are also relatively low in general. In some cases, the feed may be conditioned (e.g., size reduction, flaking, pelletizing, extrusion, backmixing with dry product) before drying, which affects the choice of dryers.

As a rule, in the interest of energy savings and reduction of dryer size, it is desirable to reduce the feed liquid content by less expensive operations such as filtration, centrifugation, and evaporation. It is also desirable to avoid overdrying, which increases the energy consumption as well as drying time.

Drying of food and biotechnological products require adherence to good manufacturing practice (GMP) and hygienic equipment design and operation. Such materials are subject to thermal as well as microbiological degradation during drying as well as in storage. If the feed rate is low (<100 kg/h), a batch-type dryer may be suited. Note that there is a limited choice of dryers that can operate in the batch mode.

In less than 1% of cases the liquid to be removed is a nonaqueous (organic) solvent or a mixture of water with a solvent. This is not uncommon in drying of pharmaceutical products, however. Special care is needed to recover the solvent and to avoid potential danger of fire and explosion. Table 1.6 presents a typical checklist of most dryer vendors use to select and quote an industrial dryer.

Drying kinetics play a significant role in the selection of dryers. Apart from simply deciding the residence time required, it limits the types of suitable dryers. Location of the moisture (whether near surface or distributed in the material), nature of moisture (free or strongly bound to solid), mechanisms of moisture transfer (rate-limiting step), physical size of product, conditions of drying medium (e.g., temperature, humidity, flow rate of hot air for a convective dryer), pressure in dryer (low for heat-sensitive products), etc., have a bearing on the type of suitable dryer as well as the operating conditions. Most often, not more than one dryer type will likely meet the specified selection criteria.

We will not focus on novel or special drying techniques here for lack of space. However, it is worth

TABLE 1.6
Typical Checklist for Selection of Industrial Dryers

Physical form of feed	<ul style="list-style-type: none"> • Granular, particulate, sludge, crystalline, liquid, pasty, suspension, solution, continuous sheets, planks, odd-shapes (small/large) • Sticky, lumpy
Average throughput	<ul style="list-style-type: none"> • kg/h (dry/wet); continuous • kg per batch (dry/wet)
<i>Expected variation in throughput (turndown ratio)</i>	
Fuel choice	<ul style="list-style-type: none"> • Oil • Gas • Electricity
<i>Pre- and postdrying operations (if any)</i>	
For particulate feed products	<ul style="list-style-type: none"> • Mean particle size • Size distribution • Particle density • Bulk density • Rehydration properties
Inlet–outlet moisture content	<ul style="list-style-type: none"> • Dry basis • Wet basis
<i>Chemical/biochemical/microbiological activity</i>	
Heat sensitivity	<ul style="list-style-type: none"> • Melting point • Glass transition temperature
<i>Sorption isotherms (equilibrium moisture content)</i>	
Drying time	<ul style="list-style-type: none"> • Drying curves • Effect of process variables
Special requirements	<ul style="list-style-type: none"> • Material of construction • Corrosion • Toxicity • Nonaqueous solution • Flammability limits • Fire hazard • Color/texture/aroma requirements (if any)
Footprint of drying system	<ul style="list-style-type: none"> • Space availability for dryer and ancillaries

mentioning that many of the new techniques use superheated steam as the drying medium or are simply intelligent combinations of traditional drying techniques, e.g., combination of heat transfer modes, multistaging of different dryer types. Superheated steam as the convective drying medium offers several advantages, e.g., higher drying rates under certain conditions, better quality for certain products, lower net energy consumption if the excess steam produced in the dryer is used elsewhere in the process, elimination of fire and explosion hazard. Vacuum steam drying of timber, for example, can reduce drying

TABLE 1.7
Conventional vs Innovative Drying Techniques

Feed Type	Dryer Type	New Techniques
Liquid suspension	<ul style="list-style-type: none"> • Drum • Spray 	<ul style="list-style-type: none"> • Fluid/spouted beds of inert particles • Spray/fluid bed combination • Vacuum belt dryer • Pulse combustion dryers
Paste/sludge	<ul style="list-style-type: none"> • Spray • Drum • Paddle 	<ul style="list-style-type: none"> • Spouted bed of inerts • Fluid bed (with solid backmixing) • Superheated steam dryers
Particles	<ul style="list-style-type: none"> • Rotary • Flash • Fluidized bed (hot air or combustion gas) 	<ul style="list-style-type: none"> • Superheated steam FBD • Vibrated bed • Ring dryer • Pulsated fluid bed • Jet-zone dryer • Yamato rotary dryer
Continuous sheets (coated paper, paper, textiles)	<ul style="list-style-type: none"> • Multicylinder contact dryers • Impingement (air) 	<ul style="list-style-type: none"> • Combined impingement/radiation dryers • Combined impingement and through dryers (textiles, low basis weight paper) • Impingement and MW or RF

times by a factor of up to 4, at the same time enhancing wood quality and reducing net fuel and electricity consumption by up to 70%. The overall economics are also highly favorable.

As the product quality requirements become increasingly stringent and as the environmental legislation becomes more and more demanding it is often found that we need to switch from one drying technology to the other. The rising cost of energy as well as the differences in the cost of fossil fuels versus electrical energy can also affect the choice of a dryer. Since up to 70% of the life cycle cost of a convective dryer is due to energy it is important to choose an energy-efficient dryer where possible even at a higher initial cost. Note that energy costs will continue to rise in the future so this will become increasingly important. Fortunately, improved efficiency also translates into better environmental implications in terms of reduced emissions of the greenhouse gas (CO₂) as well as NO_x resulting from combustion.

New dryers are developed continuously as a result of industrial demands. Over 250 U.S. patents are granted each year related to dryers (equipment) and drying (process); in the European Community about 80 patents are issued annually on dryers. Kudra and Mujumdar [44] have discussed a wide assortment of novel drying technologies, which are beyond the scope of this chapter. Suffice it to note that many of

the new technologies (e.g., superheated steam, pulse combustion—newer gas-particle contactors as dryers) will eventually replace conventional dryers in the next decade or two. New technologies are inherently more risky and more difficult to scale-up. Hence there is natural reluctance to their adoption. Readers are encouraged to review the new developments in order to make sure their selection is the most appropriate one for the application at hand. Some conventional and more recent drying techniques are listed in Table 1.7.

As is evident from the brief discussion the process of dryer selection is rather complex as typically several different choices may exist and the final choice rests on numerous criteria. Indeed, several researchers have developed expert systems for computer-aided selection of dryers. It is important to note that selection of the drying system (e.g., pre- and postdrying operations as well as the dryer) is the correct goal of the exercise. Finally, even if the dryer is selected correctly it is just as important to operate it under optimal conditions to obtain desired product quality and production rate at minimum total cost.

1.9 CONCLUSIONS

Recent studies have resulted in significant advances in the understanding of the thermodynamics of drying hygroscopic materials, kinetics of drying, evaporation

of multicomponent mixtures from porous bodies, behavior of particulate motion in various dryers, and so on. In general the empirical knowledge gained in the past two decades has been of considerable value in the design of industrial dryers, including modeling of dryers and control. On the other hand the understanding at the microscopic level of the drying mechanisms remain at a rudimentary level in the sense that modeling “drying” remains a complex and challenging task. Numerous textbooks [29–33] have appeared in recent years that focus on one or more aspects of drying. The interested reader is referred to the series *Advances in Drying* [26,34] and the journal *Drying Technology* [35] for more recent developments in the field of drying. In addition, the series *Drying of Solids* [36–38] contains valuable recent information on drying technology. The technical catalogs published by various manufacturers of drying and ancillary equipment are also very valuable. The proceedings of the biennial international drying symposia (IDS), published in the *Drying* series provide a useful guide to current R&D in drying [39–41].

It should be stressed that one must think in terms of the drying system and not just the dryer when examining an industrial dehydration problem. The preprocessing steps (feeding, dewatering, etc.) of the feed as well as the postprocessing steps (e.g., cooling, granulation, blending, etc.) and cleaning of the dryer emissions are often just as important as the dryer itself. In view of the increasingly stringent environmental regulations coming in force around the world it is not unusual for the dryer itself to cost only a small fraction of the total drying system cost. This edition of the handbook contains details concerning feeders as well as treatment of dryer emissions.

The selection of dryers and drying systems is another important area of practical significance. The cost of a poorly selected drying system is often underestimated since the user must pay for it over the entire life span of the system. If past experience is used as the sole guide, we automatically eliminate the potential benefits of specifying some of the newer dryers marketed recently around the world. Indeed, to cover the broad spectrum of new and special drying technologies that have appeared in the marketplace in the past decade, a new chapter (see Chapter 36) is devoted exclusively to this subject. The reader is urged to go through that chapter in conjunction with any specific drying application to become familiar with some of the newly developed technologies that may afford some advantages over the conventional drying technologies. Use of superheated steam rather than

hot air as the drying medium for direct dryers has attracted considerable attention in recent years. The topic is covered in some detail in a separate chapter (see Chapter 35) from the applications viewpoint.

Scale-up of dryers is perhaps the central issue of most significance in the design of dryers. Information on this subject is rather limited and widely scattered. Genskow [42] provides perhaps the only compilation of papers dealing with the scale-up of several dryer types. For design, analysis or scale-up of dryers, Houska et al. [43] have presented the general equations and methodology as applied to a number of industrial dryers. Unfortunately, the extreme diversity of products and dryer types precludes development of a single design package.

It is difficult to generate rules for both classification and selection of dryers because exceptions occur rather frequently. Often, minor changes in feed or product characteristics result in different dryer types as the appropriate choices. It is not uncommon to find that different dryer types are used to dry apparently the same material. The choice is dependent on production throughput, flexibility requirements, cost of fuel as well as on the subjective judgment of the individual who specified the equipment.

We have not considered novel dryers in this chapter. Kudra and Mujumdar [44] have discussed in detail most of the nonconventional and novel drying technologies. Most of them have yet to mature; a few have been commercialized successfully for certain products. It is useful to be aware of such advances so that the user can make intelligent decisions about dryer selection. Since dryer life is typically 25–40 years that effect of a poor “prescription” can have a long-term impact on the economic health of the plant. It is typically not a desirable option to depend exclusively on prior experience, reports in the literature or vendors’ recommendations. Each drying problem deserves its own independent evaluation and solution.

This chapter has not considered various mathematical models for drying kinetics. The reader is referred to Turner and Mujumdar [45] as well as publications in the journal *Drying Technology* for numerous approaches to mathematical modeling and numerical techniques applicable in drying technology. Mathematical models are increasingly used for design, optimization, and control of industrial dryers.

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NOMENCLATURE

A	heat transfer area, m^2
C_{LW}	heat capacity of liquid moisture, $\text{J}/(\text{kg K})$
C_{P}	heat capacity at constant pressure, $\text{J}/(\text{kg K})$
C_{PG}	heat capacity of dry gas, $\text{J}/(\text{kg K})$
C_{PW}	heat capacity of moisture vapor, $\text{J}/(\text{kg K})$
C_{PY}	humid heat, $\text{J}/(\text{kg K})$
D	molecular diffusivity, m^2/s
E	efficiency of the dryer (dimensionless)
f	relative drying rate (dimensionless)
G_{G}	dry gas flow rate, kg/s
G_{V}	evaporation rate, kg/s
ha	volumetric heat transfer coefficient, $\text{kcal}/(\text{m}^3 \text{K})$
h_{C}	convective heat transfer coefficient, $\text{W}/(\text{m}^2 \text{K})$
h_{R}	radiative heat transfer coefficient, $\text{W}/(\text{m}^2 \text{K})$
H	enthalpy, J/kg
H_{GG}	dry gas enthalpy, J/kg
H_{GW}	moisture vapor enthalpy, J/kg
I_{G}	humid enthalpy, J/kg
I_{GS}	enthalpy of dry gas, J/kg
I_{LW}	enthalpy of added moisture, J/kg
K	mass transfer coefficient $\text{kg}/(\text{m}^2 \text{s})$
m_{G}	mass of dry air, kg
m_{W}	moisture of mass vapor, kg
M_{G}	molar mass of dry gas, kg/mol
M_{W}	molar mass of moisture, kg/mol
N	rate of drying per unit surface area, $\text{kg}/(\text{m}^2 \text{s})$
N_{G}	molar gas flow per unit area, $\text{mol}/(\text{m}^2 \text{s})$
P_{G}	partial pressure of dry gas, Pa
P_{W}	partial pressure of moisture vapor, Pa
P_{W}^0	vapor pressure, Pa
P	total pressure, Pa
q	heat flux, W/m^2
Q	heat quality, J
R	gas constant, $\text{J}/(\text{mol K})$
T	temperature, K
T_{av}	average temperature, K
T_{C}	critical temperature, K
T_{D}	dew point temperature, K
T_{G}	gas temperature (dry bulb), K
T_{GS}	adiabatic saturation temperature, K
T_0	initial or reference temperature, K
T_{W}	wet bulb temperature, K
u	internal energy, J/kg
U	overall heat transfer coefficient, $\text{kcal}/(\text{s m}^2)$
V_{H}	humid volume, m^3/kg
V_{W}	specific molar volume, m^3/mol
V_{L}	specific molar volume of liquid moisture, m^3/mol
W	moisture content (wet basis) (dimensionless)
X	moisture content (dry basis) (dimensionless)
X^*	equilibrium moisture content (dimensionless)
X_{cr}	critical moisture content (dimensionless)
X_0	initial moisture content (dimensionless)

Y	humidity (mass ratio vapor/dry gas) (dimensionless)
Y_{GS}	humidity at adiabatic saturation temperature (dimensionless)
Y_{S}	saturation humidity (dimensionless)
Y_{W}	wet bulb humidity (dimensionless)

GREEK SYMBOLS

ΔH_{GM}	residual gas-mixing enthalpy, J/kg
ΔH_{R}	molar latent heat of reference substance, J/mol
ΔH_{VD}	latent heat of vaporization at T_{D} , J/kg
ΔH_{V0}	latent heat of vaporization at T_0 , J/kg
ΔH_{VS}	latent heat of vaporization at T_{S} , J/kg
ΔH_{VW}	latent heat of vaporization at T_{W} , J/kg
ΔH_{W}	molar latent heat of vaporization, J/mol
θ	temperature difference, K
θ^*	temperature difference on saturation, K
ρ	gas density, kg/m^3
ϕ	humidity-potential coefficient (dimensionless)
ϕ	characteristic moisture content (dimensionless)
ϕ_{E}	Ackermann correction, (dimensionless)
ψ	relative humidity (dimensionless)

DIMENSIONLESS GROUPS

Lu	Luikov number
Nu	Nusselt number
Pr	Prandtl number
Sc	Schmidt number

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2 Experimental Techniques in Drying

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2.1 INTRODUCTION

The calculation of drying processes requires a knowledge of a number of characteristics of drying techniques, such as the characteristics of the material, the coefficients of conductivity and transfer, and the characteristics of shrinkage. In most cases these characteristics cannot be calculated by analysis, and it is emphasized in the description of mathematical

models of the physical process that the so-called global conductivity and transfer coefficients, which reflect the total effect on the partial processes, must frequently be interpreted as experimental characteristics. Consequently, these characteristics can be determined only by adequate experiments. With experimental data it is possible to apply analytical or numerical solutions of simultaneous heat and mass transfer to practical calculations.

The general aim of drying experiments is as follows [1]:

1. Choice of adequate drying equipment
2. Establishing the data required for planning
3. Investigation of the efficiency and capacity of existing drying equipment
4. Investigation of the effect of operational conditions on the shape and quality of the product
5. Study of the mechanism of drying

When the aim of experiments is the choice of adequate drying equipment and the determination of the data required for planning, the effect of the different variables must be examined. It is practical to carry out such a series of experiments in small- or pilot-plant equipment, which operates from both thermal- and materials-handling aspects as does actual equipment, and by means of which the effect of the parameters to be examined can be studied over a wide range of conditions.

Experiments are frequently carried out in equipment of plant scale, for example, in order to establish data for certain materials, required for planning, to select the type of adequate drying equipment, or to test the efficiency of existing equipment or its suitability for drying other materials. In general, the efficiency and, further, the heat and mass balances, can be determined from data obtained with equipment of plant size.

Owing to the diversity of the aims of experimental investigations and the large number of drying characteristics, the techniques of experimental determination are extremely diverse and in many cases very specific. Therefore, without attempting to be complete, only the methods generally applied are presented here; some special cases are mentioned with reference to the sources of the literature.

2.2 DETERMINATION OF MOISTURE CONTENT

2.2.1 DETERMINATION OF THE MOISTURE CONTENT OF SOLID MATERIALS

Although determination of the moisture content of wet materials appears simple, the results obtained are often not sufficiently accurate, namely, many materials may suffer not only moisture loss but also chemical changes (oxidation, decomposition, destructive distillation, and others) on heating, and this may change the material as well. At the same time the adsorbed water must be distinguished from the so-called water of crystallization, which is frequently a very complex problem.

On selecting the techniques of moisture determination one must take into account the desired accuracy, the case of the procedure, the length of the investigation, and the complexity of the required instruments and equipment. Possible methods of measuring the distribution of moisture content during drying are presented separately.

2.2.1.1 Direct Methods

The direct methods consist essentially of determination of the moisture content of a sample by drying carried out in a drying oven with or without blow through of air, or by drying in a vacuum chamber or in a vacuum desiccator. The sample material must be prepared in every case in the following way. The material is disintegrated into pieces of 1–2 mm³, and a sample of known mass (4–5 g) is placed into a previously dried and weighed glass container, which is put into the drying chamber and dried at 102–105°C. The measurement of mass is carried out at ambient temperature, previously allowing the sample to be cooled in a desiccator. The drying process may be considered complete when the difference between the values obtained for the moisture content of the material by two consecutive measurements does not exceed $\pm 0.05\%$. The literature indicates that this process is faster when drying is carried out at 130–150°C. However, our investigations proved that results obtained in this way may deviate by 0.5–1.0%. Thus, the quick method appears to be suitable only for an approximate determination of the moisture content of a material.

The drying period may be significantly shortened by blowing air through the drying chamber, provided this air has been previously heated to 102–105°C, generally by electricity [2].

Drying of foods and other materials sensitive to heat is carried out at 60°C in a vacuum desiccator, in order to prevent their decomposition. In this case drying requires several days.

2.2.1.2 Indirect Methods

Under industrial conditions the moisture present in material must be determined by faster methods, such as by electrical methods of which three main varieties have become widespread: moisture determination based on the change of the ohmic dc resistance, a measurement of the electrostatic capacitance (dielectric constant of the material), and a measurement of the loss in an ac field. Other quick methods are the chemical methods developed mainly for the most frequently occurring case, when the moisture is water, such as the Karl–Fischer analysis based on

the chemical reaction of iodine in the presence of water [3], the distillation method, in which moisture is determined by distillation with toluene, and the extraction method, which is carried out with absolute ethanol.

For the determination of factors required for the description of the moisture transport mechanism of wet materials, the distribution of the moisture content of the dried material and even its change during the drying process must be known. These measurements are usually carried out under laboratory conditions. Two main varieties of these measurements are widespread:

1. Mechanical disintegration of the material for the rapid determination of the moisture content of the individual elements and, on compressing the specimen, its further drying [4,5]
2. Special adaptation of one of the above-mentioned electrical methods (see [Section 2.2.1.2](#)) [6]

2.2.2 DETERMINATION OF MOISTURE CONTENT OF GASES

The following methods have become widespread for the determination of the moisture content of gases: determination of the absolute value of moisture content by gravimetry or barometry (direct method), measurement of wet gas particles in air by determination of the dry and wet bulb temperatures and the dew point temperature (indirect method), and measurement of a property of the wet gas that depends on the moisture content, such as the absorptivity of the wet gas to electromagnetic waves.

Most frequently applied is the indirect method—determining the dry and wet bulb temperatures or measuring the dew point temperature, which requires a slightly more expensive instrument. However, on using these methods, the determination of the moisture content of the wet gas cannot be carried out in every case with sufficient accuracy. When high accuracy is required, one of the absolute methods must be applied. (In case of air of 50°C temperature and 50% relative moisture content, a change of 1°C in the temperature of the wet bulb results in a 9% change of the absolute moisture content, or a change of 1°C in the dew point temperature results in a 7% change of the absolute moisture content, respectively.)

The absolute determination of the moisture content of gases can be carried out by the absorption method (allowing gas to flow through a silica-gel bed or allowing it to be absorbed by methanol and then determining the water content by Karl–Fischer titration, for example). A detailed description of these and other special methods are given in Refs. [3,7].

2.3 EXPERIMENTAL DETERMINATION OF THE SORPTION EQUILIBRIUM CHARACTERISTICS OF MATERIALS

2.3.1 INTERPRETATION OF THE EQUILIBRIUM MOISTURE CONTENT OF MATERIALS

A significant part of materials contains a multitude of capillaries, micropores and macropores, and cells and micelles of various dimensions and shapes. In these materials the potential sites of moisture are determined by structural buildup.

The moisture-binding properties of materials are affected by their interior structural buildup. Thus, the equilibrium characteristics determined by measurement may be applied only to structural materials strictly identical with the material used for the measurement.

Although the materials may approximate the equilibrium moisture content either by moisture uptake (adsorption) or by drying (desorption), the value of equilibrium vapor pressure generally depends, for capillary-porous materials, upon the direction of the approximation of the equilibrium moisture content. This phenomenon is known as *sorption hysteresis*. For this reason it is necessary to distinguish adsorption equilibrium moisture from desorption equilibrium moisture.

The equilibrium moisture content is developed as a result of an interaction between the material and the environment: $\bar{X}^* = \bar{X}^*(p_v, T)$. Changes in the moisture content of the material are due to conditions (p_v, T) prevailing on the surface of the material. After a sufficiently long time—with steady-state limit conditions—an internal moisture diffusion balance takes place until the equilibrium moisture content is attained. In an equilibrium state a steady internal moisture distribution exists. Although theoretically an infinite time is needed for its formation, a practically acceptable accurate approximation may be attained by a number of procedures and methods with a finite time.

2.3.2 INTERPRETATION OF THE EQUILIBRIUM VAPOR PRESSURE

The vapor pressure at which a material of a given \bar{X} moisture content is at a given temperature T is a sorption equilibrium (i.e., it does not lose and does not take up any moisture) is the equilibrium vapor pressure $p_v^* = p_v^*(T)_{\bar{X}}$.

Generally, in drying, a knowledge of the equilibrium vapor pressure at a constant temperature is needed. Thus, since when T is constant p_v^* is constant, the equilibrium relative vapor content was

applied in drying as a characteristic of the vapor pressure:

$$\psi = \frac{p_v}{p_{ov}^*} \quad (2.1)$$

2.3.3 CHARACTERISTIC FUNCTIONS OF SORPTION EQUILIBRIUM

Since the function $\bar{X}^* = \bar{X}^*(T, p_v)$ expresses sorption equilibrium, sorption isotherms $\bar{X}^* = \bar{X}^*(p_v)_T$, sorption isobars $\bar{X}^* = \bar{X}^*(T)_{p_v}$, and sorption isosteres $p_v = p_v(T)_{\bar{X}^*}$ are applied to drying. Since sorption isotherms are most frequently applied, the method of their determination is dealt with here. Sorption isotherms are determined from point to point:

$$\bar{X}_1^*(\Psi_1)_T, \bar{X}_2^*(\Psi_2)_T \dots \bar{X}_n^*(\Psi_n)_T$$

Each pair of values determining a “point” is in general the result of a measurement.

The elements of this measurement are as follows:

1. Presentation of the pair of values to be measured on the condition that T is constant
2. Measurement of the value of p_v and ψ during the measurement or at the end of the measurement
3. At the end of the measurement the determination of equilibrium moisture content \bar{X}^* of the sample of material

Widespread methods of measurement in practice differ from each other decisively in the presentation of p_v and ψ with respect to \bar{X}^* . Techniques of determining the moisture content of wet gases are described in detail in [Section 2.2.2](#) and those of the materials are described in [Section 2.2.1](#). In general, two main methods of developing the pair of values to be measured are applied.

2.3.3.1 Measuring Techniques Carried Out at a Constant Vapor Pressure

During measurement, p_v and ψ are kept at a constant value until the moisture \bar{X} of the material sample attains the constant equilibrium value of \bar{X}^* . Practically, the sample is dried to equilibrium by means of a gas held at a constant state.

2.3.3.2 Measuring Techniques Based on Developing an Equilibrium Vapor Pressure

Equilibrium vapor pressure is developed by measurement of the material sample itself, and measurement

is continued until the values of p_v and ψ become constant in the measuring space. In preparing a drying (desorption) isotherm, the source of moisture serves as the sample.

2.3.4 MEASURING TECHNIQUES AT CONSTANT VAPOR PRESSURE

The gravimetric, the bithermal, and the method based on the interpolation of weight changes are the best known, but of these the gravimetric method is the most widespread. This method is described here in detail; references are provided for the others [8–10,29]. Constant vapor pressure is maintained by means of an acid or salt solution of a specified concentration. It is known that the moisture content of wet air, in equilibrium with an acid or salt solution of known concentration, is constant at a given temperature. Partial pressures of water vapor developed with the use of solutions of sulfuric acid can be determined, as shown in Figure 2.1. The relative vapor pressure above an aqueous solution of sulfuric acid as a function of concentration and temperature can be calculated by the correlation [2]

$$\log \frac{p_v}{p_{ov}^*} = \left(a_1 - \frac{a_2}{T} \right) + \log p_{ov}^* \quad (2.2)$$

where values for constants a_1 and a_2 are given in Table 2.1. The term p_{ov}^* in correlation Equation 2.2

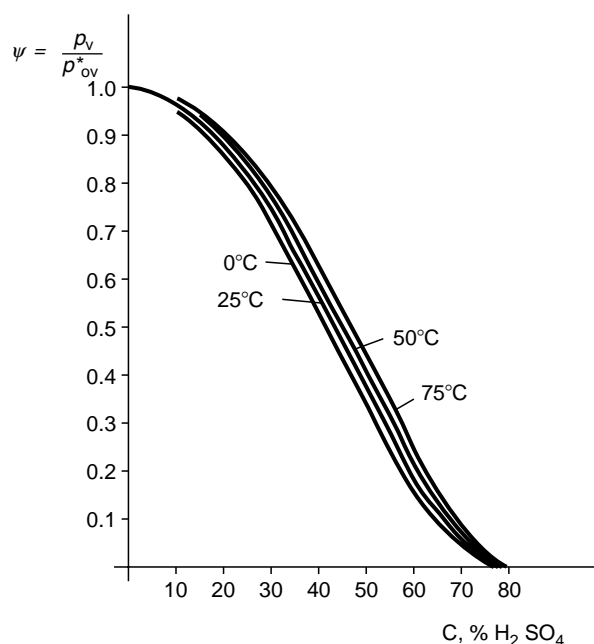


FIGURE 2.1 Partial pressures of water vapor developed with the use of solutions of sulfuric acid.

TABLE 2.1
Values of a_1 and a_2 against H_2SO_4 Concentration C

C (%)	a_1	a_2	C (%)	a_1	a_2
10	8.925	2259	60	8.841	2457
20	8.922	2268	70	9.032	2688
30	8.864	2271	80	9.293	3040
40	8.84	2299	90	9.265	3390
50	8.832	2357	95	9.79	3888

is the saturation vapor pressure of water referred to the given temperature (Table 2.2).

The relative equilibrium vapor pressures developed above saturated salt solutions are summarized in Table 2.3 [2,11,12].

With the gravimetric method a closed measuring space must exist in the thermostat so that a constant vapor pressure is formed and maintained in the manner already described. The equilibrium moisture content of the sample placed in the equilibrium airspace serves as a source for one point of the isotherm.

The static and the dynamic methods have become widespread; their names indicate a stagnant volume of air in the measuring space and airflow, respectively. The static measuring arrangement is shown in Figure 2.2, and that of a dynamic system is shown in Figure 2.3.

A variant of the dynamic method may be, for example, an air conditioner of high accuracy that maintains the constant state of a sample placed in the airflow, which is kept at given T and ψ values [13]. Development of the equilibrium moisture content of the material is attained when the weight of the sample becomes constant. This may be checked by gravimetric measurements.

The sorption equilibrium of the system material-air-solution in principle is attained only after an infinite time. However, the constant weight is observed after a finite time.

The measurement period of the static method is, particularly with high ψ values, very long (several hundred hours). Thus this method cannot be applied to perishable materials even when the thickness δ of the material sample is small and its specific surface area is high. The mass of the sample cannot be significantly decreased because of the hazard of increasing the relative error of measurement originating in the inaccuracy of the weight measurement.

With change in the moisture content of the sample during investigation the concentration of the solution is also altered, as are p_v and ψ . This change is negligible when the mass of the solution is higher by at least two orders of magnitude than the mass of the sample. But even in this case it is advisable to repeatedly

TABLE 2.2
Saturation Vapor Pressure of Water Referred to the Given Temperature

Temperature (T)		Pressure ($P \times 10^{-4}$) (Pa)	Temperature (T)		Pressure ($P \times 10^{-4}$) (Pa)
$^{\circ}\text{C}$	K		$^{\circ}\text{C}$	K	
0	273.15	0.061076	51	324.15	1.296047
1	274.15	0.065656	52	325.15	1.361163
3	276.15	0.075747	53	326.15	1.429221
4	277.15	0.081287	54	327.15	1.500123
5	278.15	0.008891	55	328.15	1.573967
6	279.15	0.093477	56	329.15	1.650950
7	280.15	0.100126	57	330.15	1.731168
8	281.15	0.107206	58	331.15	1.814623
9	282.15	0.114728	59	332.15	1.901509
10	283.15	0.122711	60	333.15	1.991731
11	284.15	0.131174	61	334.15	2.085874
12	285.15	0.140157	62	335.15	2.183941
13	286.15	0.149669	63	336.16	2.284949
14	287.15	0.159741	64	337.15	2.390861
15	288.15	0.170410	65	338.15	2.500696
16	289.15	0.181698	66	339.15	2.614453
17	290.15	0.193642	67	340.15	2.733113
18	291.15	0.206234	68	341.15	2.855696
19	292.15	0.219571	69	342.15	2.984164
20	293.15	0.233692	70	343.15	3.116553
21	294.15	0.248599	71	344.15	3.252846
22	295.15	0.264289	72	345.15	3.396043
23	296.15	0.280764	73	346.15	3.543134
24	297.15	0.298220	74	347.15	3.696126
25	298.15	0.316657	75	348.15	3.854994
26	299.15	0.335976	76	349.15	4.018765
27	300.15	0.356374	77	350.15	4.189401
28	301.15	0.377850	78	351.15	4.364940
29	302.15	0.400406	79	352.15	4.547344
30	303.15	0.424138	80	353.15	4.735631
31	304.15	0.449145	81	354.15	4.930784
32	305.15	0.375328	82	355.15	5.132801
33	306.15	0.502885	83	356.15	5.341682
34	307.15	0.531815	84	357.15	5.557429
35	308.15	0.562215	85	358.15	5.780040
36	309.15	0.593989	86	359.15	6.010496
37	310.15	0.627429	87	360.15	6.248797
38	311.15	0.662439	88	361.15	6.494944
39	312.15	0.669116	89	362.15	6.748937
40	313.15	0.737460	90	363.15	7.010774
41	314.15	0.777765	91	364.15	7.280457
42	315.15	0.819836	92	365.15	7.560927
43	316.15	0.863868	93	366.15	7.849243
44	317.15	0.909959	94	367.15	8.146384
45	318.15	0.958208	95	368.15	8.452352
46	319.15	1.008516	96	369.15	8.769106
47	320.15	1.061178	97	370.15	9.094687
48	321.15	1.116193	98	371.15	9.430075
49	322.15	1.173562	99	372.15	9.778211
50	323.15	1.233480	100	373.15	10.13223

TABLE 2.3
Relative Equilibrium Vapor Pressures Developed above Saturated Salt Solutions

Salt	T (°C)	$\Psi = \frac{P_v}{P_{ov}^*}$	Salt	T (°C)	$\Psi = \frac{P_v}{P_{ov}^*}$
BaCl ₂ ·2H ₂ O	24.5	0.88	KI	100	0.562
CaCl ₂ ·6H ₂ O	2	0.398	K ₂ S	100	0.562
	10	0.38	KNO ₂	20	0.45
	18.5	0.35	LiCl·H ₂ O	20	0.15
	20.0	0.323		73.06	0.112
	24.5	0.31		88.96	0.0994
CaHPO ₄ ·H ₂ O	22.8	0.952	LiSO ₄	24.7	0.865
	30.1	0.94	MgCl ₂	41.7	0.33
	39.0	0.954		50.8	0.311
Ca(NO ₃) ₂ ·4H ₂ O	18.5	0.56		62.3	0.301
	24.5	0.51		79.5	0.287
CaSO ₄ ·5H ₂ O	20	0.98	Mg(C ₂ H ₃ O ₂) ₂ ·4H ₂ O	20	0.65
CdBR ₂	24	0.896	Mg(NO ₃) ₂ ·6H ₂ O	18.5	0.56
	31.6	0.857		24.5	0.52
	41.5	0.835		33.9	0.505
CdSO ₄	23.9	0.906		42.6	0.472
	31.4	0.865		54.8	0.428
CuCl ₂ ·2H ₂ O	28.9	0.67		76.3	0.3275
	36.1	0.669	MgSO ₄	30	0.942
CuSO ₄	22.6	0.974		40.8	0.865
	29.5	0.97		55.1	0.843
	38.7	0.964		95.4	0.776
	91.74	0.888	NH ₄ Cl	20.0	0.792
CH ₄ N ₂ O	26.6	0.77		25.0	0.793
	34.8	0.719		30.0	0.795
	46.1	0.658	NH ₄ Cl + KNO ₃	20	0.726
C ₄ H ₆ O ₆	24.25	0.877		25	0.716
	32.43	0.818		30	0.686
	43.27	0.760	NH ₄ H ₂ PO ₄	20	0.931
CrO ₃	20	0.35		25	0.93
H ₂ C ₂ O ₄ ·2H ₂ O	20	0.76		30	0.929
H ₃ PO ₅ ·5H ₂ O	24.5	0.09		40	0.9048
KCl	24.5	0.866	(NH ₄) ₂ SO ₄	25	0.811
	32.3	0.825		30	0.811
	41.8	0.833		108.2	0.75
	55.4	0.832	NH ₄ Br	95.6	0.772
	95.5	0.773	NH ₄ NO ₃	28	0.706
KBr	20	0.84		42	0.488
	10	0.692		52.9	0.469
KC ₂ H ₃ O ₂	168.0	0.13		76	0.332
	20	0.20			0.229
K ₂ CO ₃ ·2H ₂ O	18.5	0.44	NaBr	100	
	24.5	0.43	NaBr·2H ₂ O	20	0.58
KCNS	20	0.47	NaBrO ₃	20	0.92
K ₂ CrO ₄	20	0.88	NaC ₂ H ₃ O ₂ ·H ₂ O	20	0.76
KF	100	0.229	NaCl	26.82	0.758
K ₂ HPO ₄	20	0.92		34.2	0.743
KHSO ₄	20	0.86		96.8	0.738
			NaCl + KClO ₃	16.4	0.366
NaCl + KNO ₃	16.4	0.326	Na ₄ P ₂ O ₇	22.6	0.973
NaCl + KNO ₃ + NaNO ₃	16.4	0.305	Na ₂ SO ₃ ·7H ₂ O	20	0.95
				23.5	0.9215
NaC ₂ H ₃ O ₂ ·3H ₂ O	20	0.76		30.9	0.894

TABLE 2.3 (continued)
Relative Equilibrium Vapor Pressures Developed above Saturated Salt Solutions

Salt	T (°C)	$\Psi = \frac{p_v}{p_{ov}^*}$	Salt	T (°C)	$\Psi = \frac{p_v}{p_{ov}^*}$
Na ₂ CO ₃ ·10H ₂ O	18.5	0.92		40.8	0.867
	24.5	0.87		55.4	0.83
NaClO ₃	20	0.75		94.9	0.785
	100	0.54	Na ₂ S ₂ O ₃ ·5H ₂ O	20	0.93
Na ₂ Cr ₂ O ₇ ·2H ₂ O	20	0.52	Na ₂ SO ₄ ·10H ₂ O	20	0.93
NaF	100	0.966		31.3	0.8746
Na ₂ HPO ₄ ·2H ₂ O	20	0.95	Pb(NO ₃) ₂	20	0.98
NaHSO ₄ ·2H ₂ O	20	0.52		103.5	0.884
NaI	100	0.504	TlCl	100.1	0.997
NaNO ₂	20	0.66	Tl ₂ SO ₄	104.7	0.848
NaNO ₃	27.4	0.7295	ZuCl ₂ · $\frac{1}{2}$ H ₂ O	20	0.10
	35.1	0.708	Zu(NO ₃) ₂ ·6H ₂ O	20	0.42
	59.0	0.702	ZuSO ₄	5	0.947
	102.0	0.654	ZuSO ₄ ·7H ₂ O	20	0.90

determine the concentration of the solution when weight constancy is achieved.

With the dynamic method of measurement the period of measurement is significantly shorter because over the course of evaporation the material transfer resistance decreases to a greater extent. According to the suggestion of Likov [2] the optimum rinsing rate of the measuring space is $0.2\text{--}0.3 \times 10^{-4} \text{ (m}^3/\text{s)}$.

The gravimetric method is simple and yields reliable results. There are also fully automated measuring systems [14] provided with computerized evaluation possibilities [15].

2.3.5 MEASURING TECHNIQUES BASED ON DEVELOPING AN EQUILIBRIUM VAPOR PRESSURE

The sample of a wet material placed in a closed measuring space adjusts, and after a sufficiently long

period, the vapor pressure of the space reaches equilibrium with the moist solid. If the change of moisture content of the solid is to be negligible, the moisture in the sample must be higher by at least three orders of magnitude than the amount of water required for saturation of the space. The saturation of the measuring space practically takes place at a constant moist solid vapor pressure, and the time required to attain equilibrium can be fairly long, with knowledge of the air mass present in the measuring space, the evaporation surface of the material sample, and the average evaporation coefficient [8]. For this measurement determination of the partial pressure and of the relative vapor content of the space are of fundamental importance.

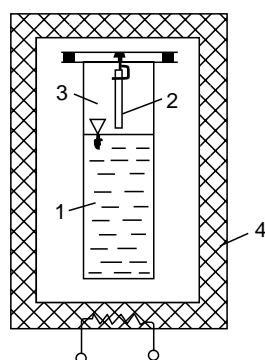


FIGURE 2.2 The measuring arrangement of a static method: (1) sulfuric acid or salt solutions, (2) sample, (3) measuring place, (4) thermostat.

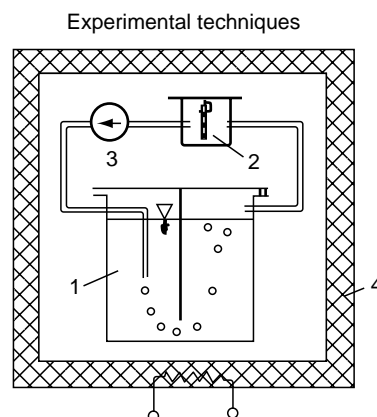


FIGURE 2.3 Schematic diagram of gravimetric dynamic system: (1) bubbling recipient with sulfuric acid or salt solution, (2) sample, (3) air pump, (4) thermostat.

Vapor pressure can be measured by manometer on freezing out the water vapor. After insertion of the material sample, the measuring space is evacuated at constant temperature. The pressure value can be read on an accurate micromanometer. The measuring space is removed from the thermostat, water vapor is frozen out, and the manometer is read again. The difference between the two values read on the manometer is equal to the equilibrium vapor pressure of the material.

If the medium in the measuring space is allowed to flow, the measurement will be lowered since the coefficient of evaporation increases. However, care must be taken during measurement to maintain the mass flow of the flowing air at a constant value.

The essence of the isotonic method [16] is that solutions of acids or salts of various concentration are placed in the closed space, where the mass of these solutions is negligible in comparison with the mass of the material sample, together with the material sample itself. These solutions act as sources of absorbents of moisture, respectively, from the material sample. At a given temperature over the course of measurement their composition approaches equilibrium concentration of the solution, corresponding to the equilibrium pressure.

In a method utilizing the deliquescence of crystalline salts, the salt crystals deliquesce in a space with a vapor pressure that exceeds the vapor pressure of their saturated solution. In this way it is possible to draw from their state conclusions about whether the vapor pressure of the measuring space is actually below or above the limit value. The dry bulb temperature and the dew point temperature of air determine the state of air present in the measuring space and also the partial pressures. The method based on this consists essentially of measuring the dew point of the air. The measurement of the equilibrium vapor tension of the air can be carried out by means of hygroscopic cell.

The state of the air is unequivocally determined by the dry and wet bulb temperatures of the air. Methods based on the measurement of the wet bulb thermometer apply this principle. Figure 2.4 presents the principle of the measuring equipment of a method of this type.

2.4 TECHNIQUES AND EQUIPMENT OF THE INVESTIGATION OF DRYING KINETICS

In drying, the heat transfer coefficient h and the mass transfer coefficient K between the drying gas and the wet material, the heat diffusivity κ_h and moisture (mass) diffusivity κ_m coefficients, and the thermal conductivity λ_h and moisture conductivity λ_m

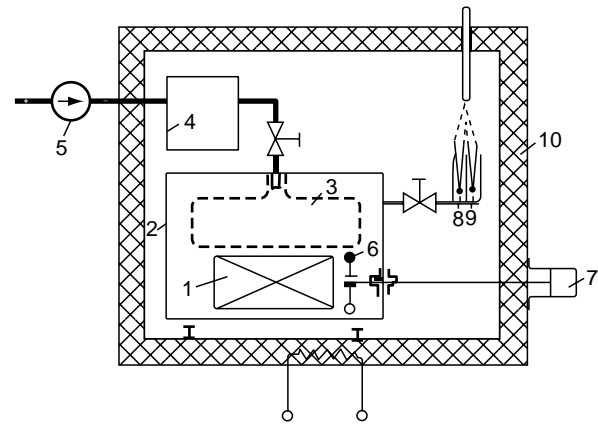


FIGURE 2.4 Schematic diagram of the arrangement based on measuring the wet bulb temperature: (1) sample, (2) measuring chamber, (3) foil balloon, (4) dashpot, (5) air pump, (6) paddle wheel with magnetic clutch, (7) motor, (8) thermometer, (9) wet bulb thermometer, (10) thermostat.

coefficients are the parameters of interest. Although several methods are employed for the determination of these parameters, here only the techniques and equipments that allow their determination via a drying experiment will be described.

From the simpler drying experiments a fair amount of data may be obtained on the basis of which—in fractional drying—conclusions can be drawn concerning the time required for the prescribed change in moisture content and the transport characteristics listed above.

When the heat and mass diffusivity properties of the drying material are also needed, a technique of investigation and measurement is applied such that the distribution of temperature and moisture content can also be measured during drying, along with the thickness of the dried material. When, however, during investigation, the above-mentioned distributions are not measured and only the integral-average moisture content of the dried material and eventually the temperature of the drying surface (or the average temperature) are measured, the model used to analyze the data is the so-called “lumped parameter” type as opposed to the “distributed parameter” model applied when local values are measured.

2.4.1 DESCRIPTION OF THE MEASURING EQUIPMENT

Experimental measuring equipment developed for drying by convection is shown in Figure 2.5. The equipment is in fact a controlled climate air duct. The wet material is placed on a balance in the measuring space. The temperature and moisture content of

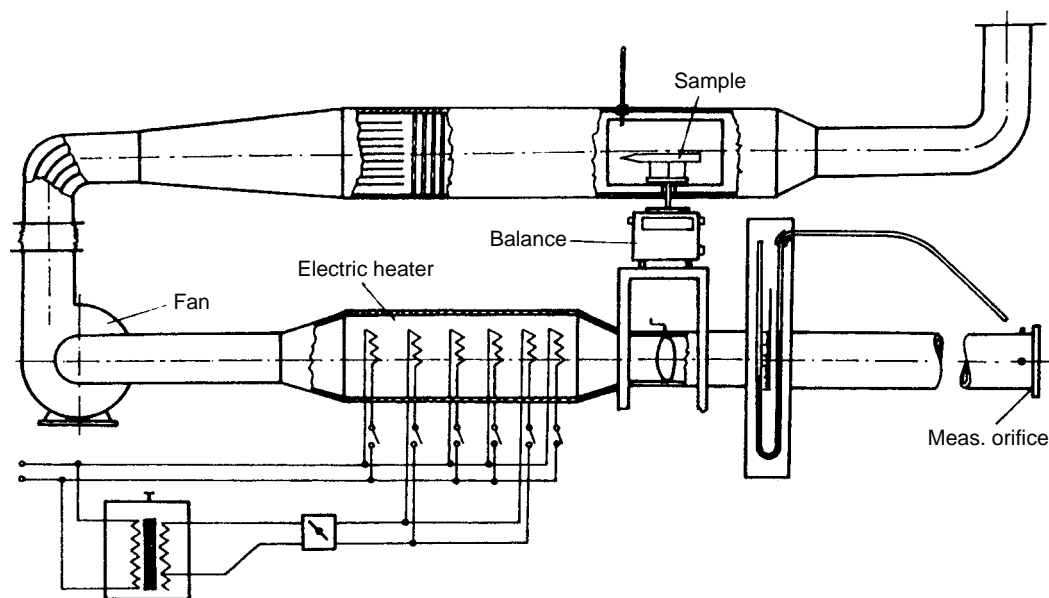


FIGURE 2.5 Drying apparatus.

the drying air is measured by wet and dry bulb thermometers. Other online humidity measuring devices may be used for convenience.

The flow rate can be determined by means of an orifice meter. Prior to beginning the measurement, the fan is started and the desired airflow adjusted by a throttle valve; the temperature and moisture content of the air are controlled by means of an electric heater and a steam valve. In the next section the development of an adequate value of flow rate may be ensured by means of screens. The time-varying mass of the sample to be dried is measured continuously and recorded.

2.4.2 DRYING EXPERIMENTS (LUMPED APPROACH)

To determine the drying rate, the mass of a sample placed in the airflow (constant temperature, humidity, and velocity) must be measured as a function of time. In order to obtain results that can be applied for scale-up, the following aspects must be taken into account: the sample must not be too small and the conditions of drying must if possible be identical to the conditions anticipated in the industrial unit [1,17]:

1. The sample must be placed in a similar way in the laboratory unit.
2. The ratios of the drying surface to the nondrying surface must be identical.
3. The conditions of heat transfer by radiation must be similar.

4. The temperature and the moisture content of air (drying gas) and its velocity and direction must be identical with respect to the sample.

It is practical to carry out several experiments with material samples of various thicknesses and to determine in every case the dry matter content of the sample in the way described in Section 2.2.

When the temperature, moisture content, and velocity of the drying air are constant, drying takes place under constant drying conditions. This way of drying is ensured by the equipment sketched in Figure 2.5.

Since the wet mass is

$$m_{sw} = m_w + m_s \quad (2.3)$$

and the moisture content is

$$X = \frac{m_w}{m_s} \quad (2.4)$$

thus

$$\frac{m_{sw}}{m_s} = 1 + X$$

That is,

$$X = \frac{m_{sw} - m_s}{m_s} \quad (2.5)$$

These correlations indicate that, with a knowledge of the drying wet mass as a function of time [$m_{sw} = m_{sw}(\tau)$] and of the bone-dry mass of the sample, it is

possible to plot the moisture content of the sample as a function of time (Figure 2.6).

This curve can be directly applied for the determination of time of drying greater masses to a prescribed lower moisture content, provided the drying is carried out under identical conditions. However, better information is obtained when, on the basis of Figure 2.6, the drying rate is plotted against the moisture content of the material. The drying rate is defined by

$$N_W = -\frac{1}{A_s} \frac{dm_{SW}}{d\tau} = \frac{m_S}{A_s} \frac{d\bar{X}}{d\tau} \quad (2.6)$$

In Equation 2.6, A_s denotes the contact surface of the drying gas and the dried material, and a cross section perpendicular to the airflow in through circulation drying.

According to Equation 2.6, with the knowledge of the weight loss curve we can plot Figure 2.7, which represents the drying rate. If the drying rate is known under one set of constant drying conditions, it can be extrapolated within limits to other conditions [17]. Figure 2.8 shows the variation of drying surface temperature with time whereas Figure 2.9 summarizes the characteristic parameters of the drying process.

The determination of transport coefficients by a drying experiment is achieved as follows. During constant rate drying, the temperature of the drying surface is, in a purely convective drying, the so-called wet bulb temperature. With this value, the steady-state heat flux is

$$j_q = h(T_G - T_W) \quad (2.7)$$

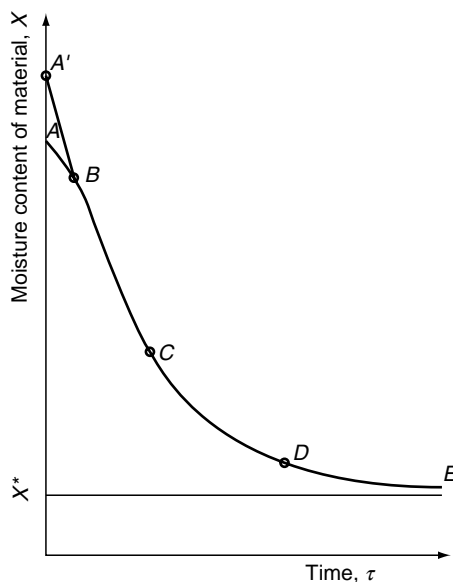


FIGURE 2.6 Batch, convection drying curve.

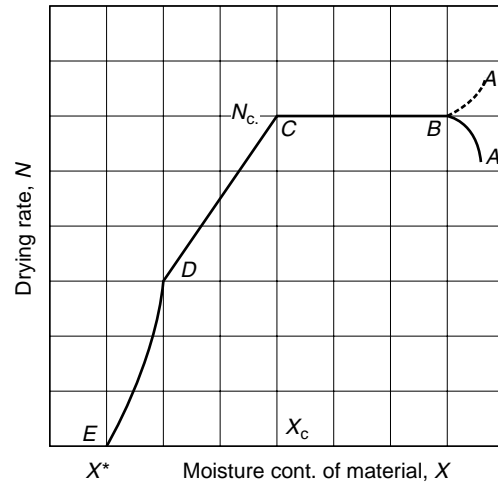


FIGURE 2.7 Drying rate curve.

where j_q can also be calculated using the constant drying rate

$$j_q = N_{wc} \Delta H \quad (2.8)$$

Thus,

$$h = N_{wc} \left(\frac{\Delta H}{T_G - T_W} \right) \quad (2.9)$$

In terms of the mass transfer coefficient,

$$N_{wc} = K(Y_W - Y_G) \quad (2.10)$$

where Y_W is the moisture content of the saturated air at temperature T_W , the coefficient of evaporation (mass transfer) is therefore

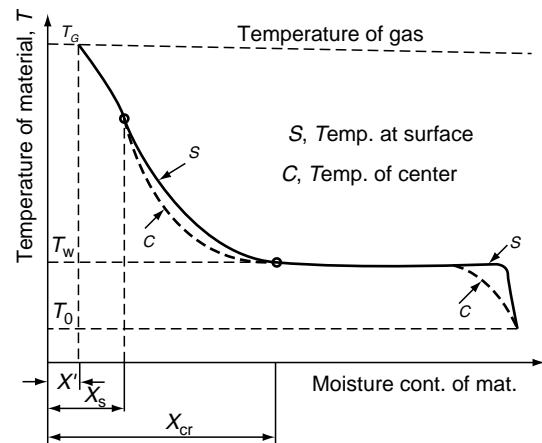


FIGURE 2.8 Temperature curves of a wet material at the surface and the middle of the sample.

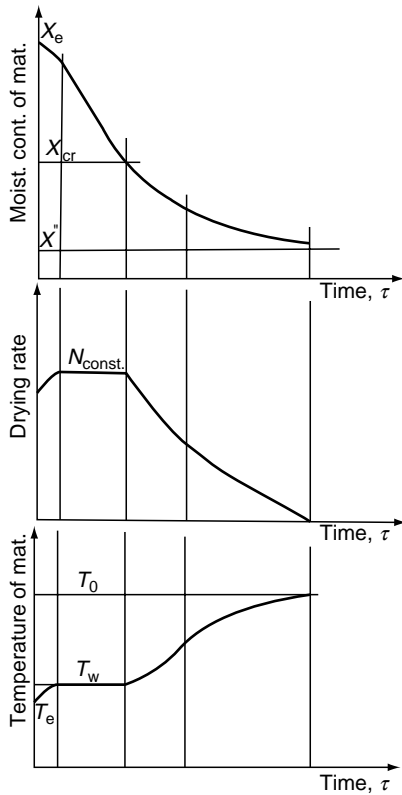


FIGURE 2.9 Characteristic functions of a convective drying.

$$K = \frac{N_{wc}}{Y_w - Y_g} \quad (2.11)$$

The heat and mass transfer coefficients on the gas side, characterizing the steady state, may be considered constant during the entire drying process, under constant drying conditions. During the falling rate period, the process is significantly affected by the internal heat and moisture diffusivities. Therefore, determination of these coefficients is also necessary. It may be noted here that the transport coefficients on the gas side also change during this period owing to the continuous decrease in vapor diffusivity of the material in the surface layer [18].

2.4.3 TECHNIQUES OF INVESTIGATION WITH DISTRIBUTED PARAMETERS

In the initial constant rate drying period, a knowledge of the heat diffusivity and heat conductivity coefficients of the wet material is necessary because they are the controlling factors for heat transport within the material. In this section of drying it is presumed that the pores of the wet material are saturated by moisture. Consequently, the above-mentioned characteristics of a heterogeneous material consisting of a solid skeleton and, within this, a capillary system (filled

with moisture) of diversified sizes and shapes must be determined.

2.4.3.1 Determination of the Thermal Conductivity and Diffusivity of Wet Materials

The method for drying a plane slab is described, but a similar method may also be developed for bodies of other geometric shapes [2,8,30].

During drying of wet capillary-porous slabs, when only surface moisture is removed and the value of the so-called phase-change criterion ε is zero—that is, no evaporation takes place within the material itself—the development of the temperature of the wet material at the surface and in the plane of symmetry can be determined analytically [8,19].

Using an apparatus of the type shown in Figure 2.5, the surface temperature T_s and the slab midplane temperature T_z are monitored in addition to the change in weight of the sample. In the initial transient period, surface temperature variation is plotted as a function of the corresponding temperatures at the midplane (Figure 2.10). Figure 2.11 shows a plot of the logarithm of the absolute difference between T_s and T_z versus time. Except at very short times, the curve is linear as expected from the analytical solution of transient one-dimensional diffusion of mass or heat.

The surface temperature is expressed by the function [20]

$$\frac{T_s - T_w}{T_0 - T_w} = \sum_{k=1}^{k=\infty} 2 \left(\frac{\sin \delta_k \cos \delta_k}{\delta_k + \sin \delta_k \cos \delta_k} \right) \exp(-\delta_k^2 Fo) \quad (2.12)$$

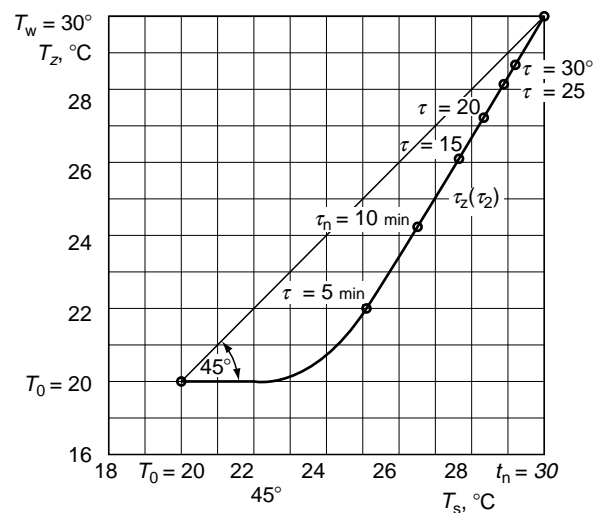


FIGURE 2.10 Temperature on the surface versus the temperature on the symmetry plane.

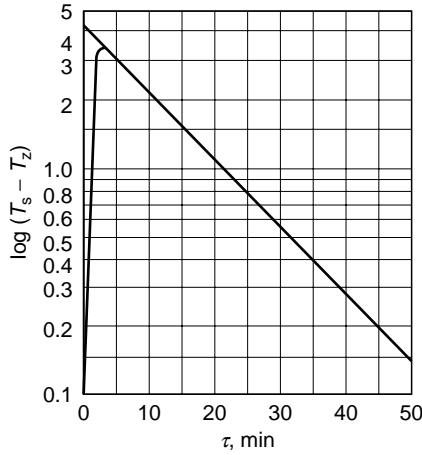


FIGURE 2.11 Plot of $\log(T_s - T_z)$ versus time.

and the temperature of the center of the slab by the function

$$\frac{T_z - T_W}{T_0 - T_W} = \sum_{k=1}^{k=\infty} 2 \left(\frac{\sin \delta_k}{\delta_k + \sin \delta_k \cos \delta_k} \right) \exp(-\delta_k^2 Fo) \quad (2.13)$$

where the eigenvalues δ_k are roots of the transcendent equation

$$\text{ctg } \delta_k = \frac{k}{Bi_{hm}} \quad (2.14)$$

In the above equations the Biot and Fourier numbers are

$$Bi_{hm} = \frac{h}{\lambda} Z \left(1 + \frac{h_m}{h} \right) \quad (2.15)$$

and

$$Fo = \frac{\kappa_r}{Z^2} \quad (2.16)$$

The coefficient h_m in Equation 2.15 is given by

$$h_m = \Delta HKS^* \quad (2.17)$$

where

$$S^* = \frac{Y_W - Y_G}{T_W - T_d} \cong \frac{Y_s - Y_G}{T_s - T_d} = \text{constant} \quad (2.18)$$

is the equation of the equilibrium curve, considered a straight line. It appears from Equation 2.12 and Equation 2.13 that after certain time (Fourier number)

only the first term of the series is predominant. Neglecting further terms we obtain

$$\frac{T_z - T_W}{T_s - T_W} = \frac{dT_z}{dT_s} = \frac{1}{\cos \delta_1} \quad (2.19)$$

and

$$\frac{d[\ln(T_s - T_z)]}{d\tau} = -\frac{\delta_1^2 \kappa_h}{Z^2} \quad (2.20)$$

In Equation 2.19, δ_1 is an eigenvalue, with which the thermal diffusivity κ_h can be determined from Equation 2.20.

The function $T_z(T_s)$ shown in Figure 2.10 becomes nonlinear after a certain time, when the superficial moisture disappears. However, on lengthening the straight section, the point of intersection of unit steepness (45°), starting from the origin, indicates the limiting temperature T_W that the temperature of the material approaches, over the course of drying of the surface moisture.

Knowing κ , the thermal conductivity coefficient λ of the wet material can be calculated from Equation 2.21 when c_{sw} , the specific heat of the wet material, and $\bar{\rho}_{Os}$, the density of the dry material, are known:

$$\kappa = \frac{\lambda}{\bar{\rho}_{Os} c_{sw}} \quad (2.21)$$

In an example of the determination of the thermal diffusivity and thermal conductivity, the wet material is a suspension of water and powdered chalk. Its characteristics are given as $T_0 = 14.3^\circ\text{C}$, $Z = 40 \text{ mm} = 4 \times 10^{-2} \text{ m}$, $c_{sw} = 2.51 \text{ kJ/kg } ^\circ\text{C}$, and $\rho_{Os} = 1500 \text{ kg/m}^3$. The drying gas is air. Its characteristics are $T_G = 52.5^\circ\text{C}$, $T_d = 10.8^\circ\text{C}$, $P = 1 \text{ bar}$, and $c_{PG} = 1.05 \text{ kJ/kg } ^\circ\text{C}$. Values of temperatures observed during drying are given in Figure 2.12.

T_W extrapolated from Figure 2.12 is $T_W = 25.2^\circ\text{C}$. According to Equation 2.19,

$$\frac{dT_z}{dT_s} = \frac{1}{\cos \delta_1} = \frac{T_{z81'} - T_{z24'}}{T_{s81'} - T_{s24'}} = \frac{20.73 - 16.13}{24.18 - 23.17} = 4.55$$

where

$$\cos \delta_1 = 0.22$$

that is, the first eigenvalue is $\delta_1 = 1.35$.

Figure 2.13 shows the logarithm of the difference between the temperatures at the surface and in the

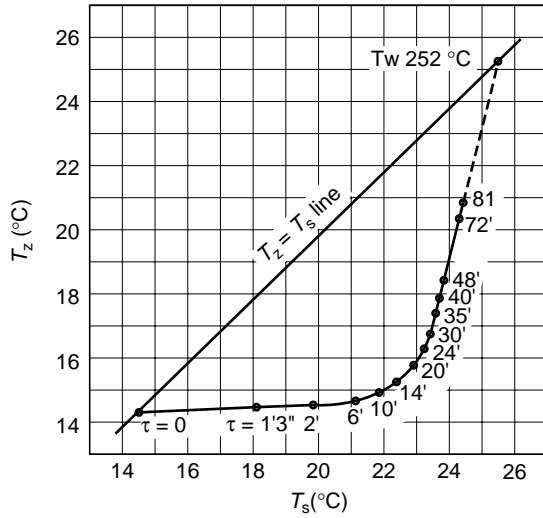


FIGURE 2.12 Experimental $T_z(T_s)$ curve for suspension of water and powdered chalk.

center plane plotted against time. Using the data of Figure 2.13, it follows according to Equation 2.20 that

$$\begin{aligned} -\frac{d[\ln(T_s - T_z)]}{d\tau} &= \frac{\delta_1^2}{Z^2} \\ &= \frac{\ln(T_{s24'} - T_{z24'}) - \ln(T_{s81'} - T_{z81'})}{81 - 24} \\ &= 0.7551/h \end{aligned}$$

Thus the heat diffusivity of the given wet material is

$$\begin{aligned} &= \frac{\lambda}{c_{sw}\bar{\rho}_{Os}} = 0.755 \frac{Z^2}{\delta_1^2} = 0.755 \frac{16 \times 10^{-4}}{1.35^2} \\ &= 0.655 \times 10^{-3} \text{ m}^2/\text{h} \end{aligned}$$

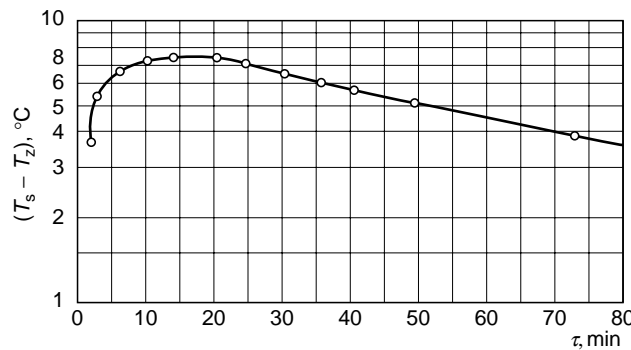


FIGURE 2.13 Experimental $\log(T_s - T_z) - \tau$ functions for suspension of water and powdered chalk.

and the thermal conductivity is

$$\begin{aligned} \lambda &= c_{sw}\bar{\rho}_{Os} = (0.655 \times 10^{-3})(2.51)(1500) \\ &= 2.47 \text{ kJ/m h } ^\circ\text{C} \end{aligned}$$

2.4.3.2 Determination of the Mass Diffusivity and Moisture Conductivity Coefficient of Wet Materials

The mass diffusivity coefficient can be determined from the drying curve already discussed. It has been shown for drying of a plane slab [20], that in the falling rate period of drying as the equilibrium moisture content is approached, the time rate of change to the weight of the slab as a function of its weight is a linear function, given by

$$\frac{-(dm/dr)}{m - m^*} = \nu_1^2 \frac{m}{Z^2} = \nu_1^2 \frac{\lambda_m}{c_m \bar{\rho}_{Os} Z^2} \quad (2.22)$$

The eigenvalue ν_1 in Equation 2.22 is defined by

$$\text{ctg } \nu_1 = \frac{\nu_1}{Bi_m} = \frac{\nu_1 \lambda_m}{S^{**} K Z c_m} \quad (2.23)$$

where S^{**} in Equation 2.23 is given by

$$S^{**} = \frac{P_{Vs} - P_{VW}}{X_s - X_w} \quad (2.24)$$

therefore the slope of the sorption isotherm (assuming that the isotherms are not strongly dependent on temperature).

On the basis of Equation 2.22 and Equation 2.23, we obtain

$$\nu_1 \text{ ctg } \nu_1 = \frac{(-dm/dr)\bar{\rho}_{Os}Z}{(m - m^*)KS^{**}} \quad (2.25)$$

The right side of Equation 2.25 contains known data when the weight of the wet material is measured as a function of time and when we have already determined, according to the previous section, the mass transfer coefficient K from data obtained in the constant rate drying period. The first eigenvalue can be determined numerically (by trial and error or by iteration) from the transcendental equation. (The value of the ν ranges between 0 and π , Ref. [2].) The mass diffusivity coefficient κ_m can be calculated directly from Equation 2.22 if ν_1 is known. When the moisture capacity coefficient c_m is determined separately [2,19], then the moisture conductivity coefficient λ_m can also be determined from Equation 2.23.

2.4.3.3 Determination of the Thermal Conductivity and Effective Diffusivity Coefficient of the Dry Material

In the course of drying of macroporous and capillary-porous materials—in the falling rate period of drying—the moisture from the pores of the material is removed, and an evaporation front penetrates from the surface of the material into the interior. At a given time the primarily wet material can be divided into two parts separated by an evaporation front: a part already dried and a part that is still wet. The moisture evaporated at the plane of phase change must pass through the pores of the already dried layer in order to leave the material. At the same time the heat required for drying has to pass through the already dried layer to carry out the evaporation. This model of drying is known as the *penetrating* or *receding front model* and is described in the literature [2,4,21,22].

Besides drying with a penetrating front, other drying phenomena also occur in which, in the course of the drying, development of a layer inhibits the drying process (in drying two-layer materials, such as salami varieties, seeds coated with an earth layer, and so on). The mathematical model of the drying process of such two-layer materials of constant thickness is described in detail in the literature [23,24]. For a description of drying by convection of double-layer materials, in the drying section with a penetrating front, and also of the macroporous and capillary-porous bodies—besides the material characteristics and transport coefficients—knowledge of the thermal diffusivity of the dry layer and the effective diffusivity coefficient value is also necessary. When the mean free path of the molecules removed is shorter than the characteristic dimension of the capillary, the vapor diffusion process through the pores of the dry layer is the so-called molecular diffusion process. However, the length of the diffusion path and the cross section available for vapor flow may change the direction of the diffusion. For this case, Krischer and Kröll [25] defined an effective diffusivity coefficient

$$D_a = \frac{D_{VG}}{\mu_D} \quad (2.26)$$

where D_{VG} is the conventional diffusivity of the vapor in free air and μ_D is the coefficient of resistance due to diffusion. Determination of the effective diffusivity coefficient by diffusion experiments is widely known in the literature [2,19]. Since these experimental measurements are relatively complex, we describe here the determination of this coefficient by means of a simple series of drying experiments. For drying double layers the rate of drying can be defined as [22]

$$j_m = K_G(p_{v\xi} - p_{VG}) \quad (2.27)$$

where

$$K_G = \frac{1}{1/k_G M_V + \xi(RT/D_a M_V)} \quad (2.28)$$

is the overall mass transfer coefficient, p_{VG} the partial pressure of vapor of the bulk gas phase, and $p_{v\xi}$ the vapor pressure at the evaporation plane (see Figure 2.14).

It can be shown that a steady state exists in which the rate of drying is

$$j_{me} = \frac{1}{1/k_G M_V + \xi(RT/D_a M_V)} (p_{v\xi e} - p_{VG}) \quad (2.29)$$

where $p_{v\xi e}$ is the equilibrium partial vapor pressure corresponding to the equilibrium temperature of the site evaporation front. The equilibrium temperature is given by

$$T_{\xi e} = \frac{T_d + T_G(K_H/\Delta HSK_G)}{1 + (K_H/\Delta HSK_G)} \quad (2.30)$$

In Equation 2.30,

$$S = \frac{p_{v\xi e} - p_{VG}}{T_{\xi e} - T_d} \quad (2.31)$$

and the overall heat transfer coefficient is

$$K_H = \frac{1}{1/h + \xi/\lambda_1} \quad (2.32)$$

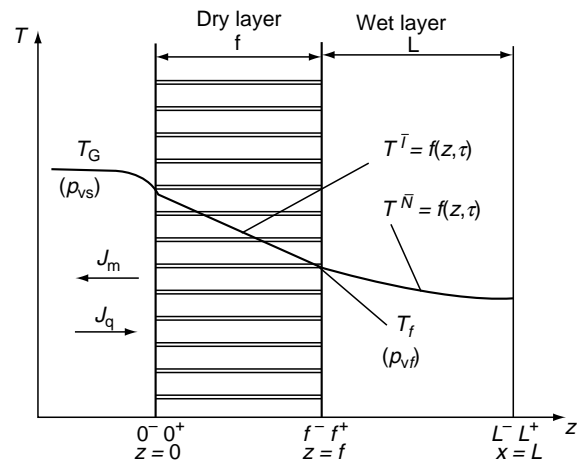


FIGURE 2.14 The double-layer model investigated.

TABLE 2.4
Data Measured on the Stabilized State

No. Experiments	$\frac{\xi}{Z}$	ξ (mm)	$j_{me}10^5$ (kg/m ² s)	$T_{\xi e}$ (°C)	T_d (°C)	T_G (°C)
1	0.2	4	8.14	40.0	10.0	48.0
2	0.3	6	6.44	40.9	6.5	48.6
3	0.5	10	4.53	41.9	11.5	48.0
4	0.7	14	3.25	42.0	12.0	47.2
5	0.8	16	3.11	42.2	11.5	47.7

On rearranging Equation 2.29, we obtain

$$\frac{P_{V\xi e} - P_{VG}}{j_{me}} = \frac{1}{k_G M_V} + \xi \frac{RT}{D_a M_V} \quad (2.33)$$

Equation 2.33 can be used as the basis for determining D_a if the mass transfer coefficient k_G is known. Other necessary parameters are measurable from the weight loss curve.

For the determination of the thermal conductivity of the dry layer, a steady state yields

$$j_{qe} = j_{me} \Delta H \quad (2.34)$$

or

$$K_H(T_G - T_{\xi e}) = j_{me} \Delta H \quad (2.35)$$

Substituting K_H given by Equation 2.32 into Equation 2.35 and rearranging we obtain

$$\frac{T_G - T_{\xi e}}{\Delta H j_{me}} = \frac{1}{h} + \xi \frac{1}{\lambda_1} \quad (2.36)$$

This equation can be used to determine λ_1 if all other parameter values are known.

In an experimental apparatus utilizing constant (and known) drying conditions, change in weight of

the two-layer slab as well as the temperature of the evaporation front must be measured as functions of time. Appropriate plots of Equation 2.33 and Equation 2.36 using this data yield the effective and thermal conductivity of the dry layer. An example of the determination of the thermal conductivity and effective diffusivity coefficients of dry layers follows.

A two-layer slab consisting of a sand layer and a layer of gypsum is used in this illustration. Five different thicknesses of the sand layer were employed. The flow rate of the drying air was the same in all experiments. The measured data at steady state are summarized in Table 2.4. Table 2.5 gives the calculated results using Equation 2.33 and Equation 2.36. Plots corresponding to Equation 2.33 and Equation 2.36 are presented in Figure 2.15 and Figure 2.16. The transfer coefficients are determined from the slopes of the linear regions of the curves. The results are

$$\begin{aligned} k_G &= 5.56 \times 10^{-4} \text{ kmol/(m}^2 \text{ bar s)} \\ D_a &= 9.1 \times 10^{-6} \text{ m}^2/\text{s} \\ h &= 0.031 \text{ kJ/(m}^2 \text{ s K)} \\ \lambda_1 &= 0.4 \text{ J/(m s K)} \end{aligned}$$

2.5 DRYING OF FIXED AND MOVING BEDS

In through circulation fixed bed dryers and in moving or suspended bed dryers (spouted beds, fluidized beds, rotary dryers, and so on), the actual interfacial area participating in heat and mass transfer is unknown. For such cases it is appropriate to define volumetric transfer coefficients for heat and mass transfer, h_V and K_V , respectively, as follows:

$$h_V = h \frac{dA_s}{dV} \approx h \frac{A_s}{V} \quad (2.37)$$

and

$$K_V = K \frac{dA_s}{dV} \approx K \frac{A_s}{V} \quad (2.38)$$

TABLE 2.5
Calculated Results (from Table 2.4)

No. Experiments	ξ (mm)	$p_{V\xi e}$ (bar)	p_{VG} (bar)	$\frac{P_{V\xi e} - P_{VG}}{j_{me}}$ (bar·m ² ·s/kg)	$\frac{T_G - T_{\xi e}}{\Delta H j_{me}}$ (°C·m ² ·s/kJ)
1	4	0.07374	0.01251	752.21	40.91
2	6	0.07889	0.009871	1071.72	47.2
3	10	0.08317	0.01383	1519.2	55.8
4	14	0.0836	0.01429	2131.2	66.75
5	16	0.08449	0.01383	2271.6	73.72

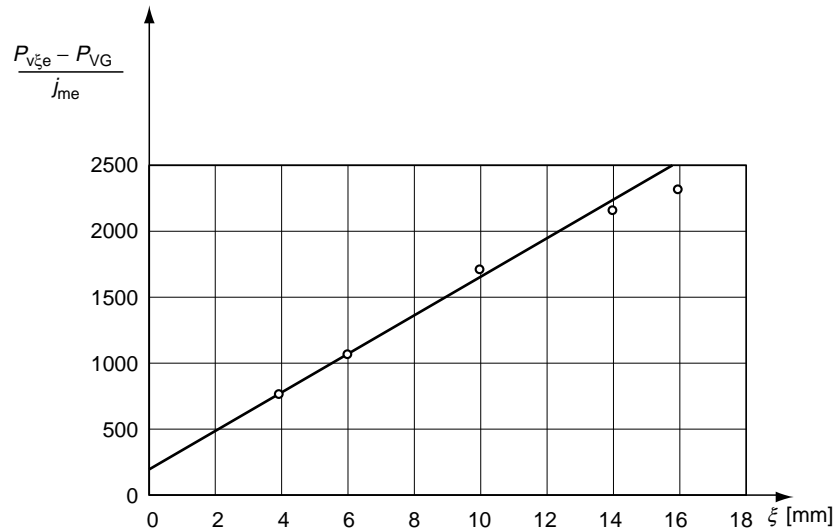


FIGURE 2.15 Determination of the mass transfer coefficient and the effective diffusivity.

Here dA_s and A_s are the differential and total interfacial areas, respectively, for heat and mass transfer whereas dV and V are the corresponding differential and total bed volumes, respectively.

2.5.1 DETERMINATION OF THE VOLUMETRIC HEAT TRANSFER COEFFICIENT

The method described here is for a plant-size, direct rotary dryer although the technique is applicable to other dryers of granular media as well. In a plant scale, effects of various parameters cannot be studied in depth as one could in a laboratory or pilot-plant size equipment. Hence, such data are not generally applicable.

Figure 2.17 is a schematic of the dryer employed for this investigation. It consists of a cylindrical shell (1 in Figure 2.17) covered with a heating jacket (2

and fitted with an agitator (3) within the shell. Heat for drying can be supplied directly by hot air or indirectly by circulating a heating fluid through the jacket. In direct drying the drying air is heated.

The stirrer carries out agitation and disintegration of the dried grains driving rings (4 in Figure 2.17) and plates (5) are attached in a staggered fashion on the agitator in order to increase the turbulence of the drying medium and the residence time of the material to be dried. The scraper-stirrer plates (8), which stir the material and feed it into the flow of the drying gas, are fastened to these rings and plates, respectively. The material to be dried is fed through the hopper (6) to the feeder (7). Exit air leaves through the dust separator (9).

The parameters measured include flow rate of the solids feed, mass flow rate of air, solids temperature at entry and exit, and the humidity of the drying air

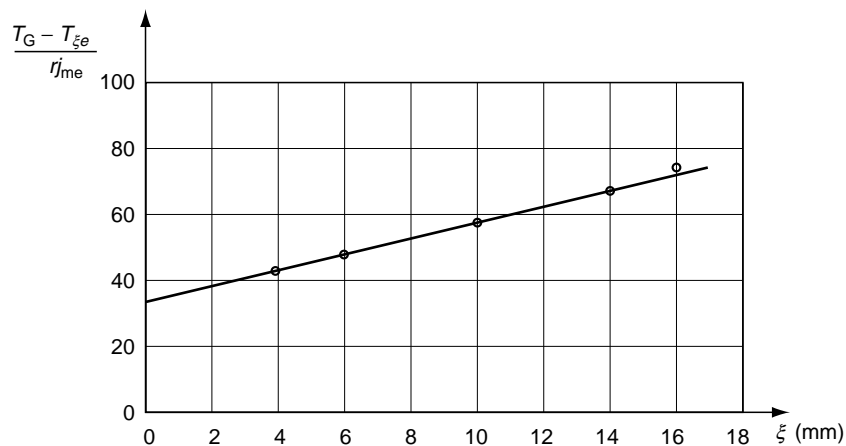


FIGURE 2.16 Determination of the heat transfer coefficient and the thermal conductivity.

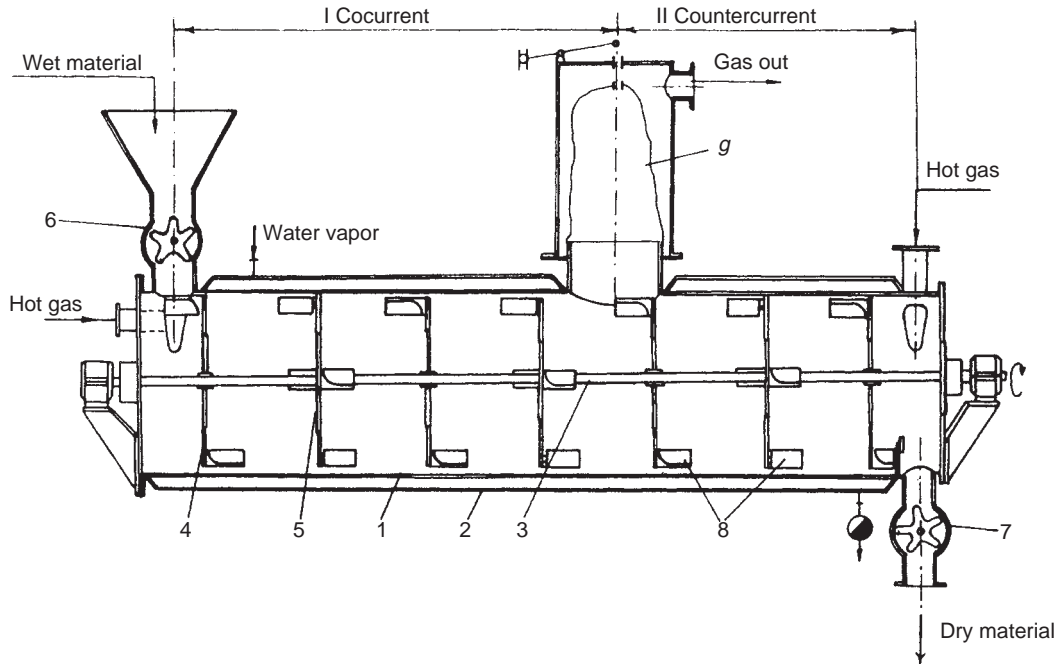


FIGURE 2.17 The contact-fluid dryer (see text for discussion).

entering and leaving the dryer. The volumetric heat transfer coefficient can then be calculated from the definition

$$Q_H = hA_s \Delta T_m \quad (2.39)$$

where the average temperature difference for cocurrent flow is the logarithmic average value

$$\begin{aligned} \Delta T_m &= \Delta T_{\log} \\ &= \frac{(T_{G \text{ in}} - T_{SW \text{ in}}) - (T_{G \text{ out}} - T_{SW \text{ out}})}{\ln \frac{T_{G \text{ in}} - T_{SW \text{ in}}}{T_{G \text{ out}} - T_{SW \text{ out}}}} \end{aligned} \quad (2.40)$$

Dividing both sides of Equation 2.39 by the empty volume of the equipment V results in

$$\frac{Q_H}{V} = h \frac{A_s}{V} \Delta T_m \quad (2.41)$$

that is,

$$\frac{Q_H}{V} = h_V \Delta T_m \quad (2.42)$$

The transfer rate is given by

$$\begin{aligned} Q_H &= G_T c_{pG} (T_{G \text{ in}} - T_{G \text{ out}}) \\ &= L_{SW} c_{SW} (T_{SW \text{ out}} - T_{SW \text{ in}}) \end{aligned} \quad (2.43)$$

On the basis of Equation 2.42 and Equation 2.43, the volumetric heat transfer coefficient can also be determined as

$$\begin{aligned} h_V &= \frac{G_T c_{pG} (T_{G \text{ in}} - T_{G \text{ out}})}{V \Delta T_{\log}} \\ &= \frac{L_{SW} c_{SW} (T_{SW \text{ out}} - T_{SW \text{ in}})}{V \Delta T_{\log}} \end{aligned} \quad (2.44)$$

In these experiments already dried granular material must be used to produce a process of pure sensible heat transfer.

For a direct-heated rotary dryer, the volumetric heat transfer coefficient can be estimated from the following empirical correlation

$$h_V = 2950 n \phi^{0.173} G_0^{0.9} \quad (2.45)$$

where ϕ is the fractional holdup (volume of granular material as a fraction of the total volume of the dryer), n is the rpm of the agitator, and G_0 is the mass flow rate of the drying air.

2.5.2 DETERMINATION OF THE HEAT AND MASS TRANSFER COEFFICIENTS IN THROUGH CIRCULATION DRYING

In general one needs only the volumetric heat and mass transfer coefficients for design purposes.

In some special cases (e.g., drying alfalfa plants), the material to be dried may consist of a mixture of materials with distinctly different characteristics (the stems and the leaves in the case of alfalfa). The stems have a well-defined smaller surface area whereas the leaves have large areas which may, however, be covered, therefore not all areas are available for heat and mass transfer. The drying rates are quite dissimilar for the two components, therefore care must be exercised in drying whole alfalfa plants. Refs. [26,27] give the necessary details for the determination of the volumetric as well as real heat and mass transfer coefficients. These references give a methodology for evaluation of the interfacial area for heat and mass transfer on the basis of certain assumptions.

The volumetric mass and heat transfer coefficients can be determined simply from the definitions (Equation 2.46 and Equation 2.47) for an adiabatic operation. (See Ref. [28] for details.)

$$K_V = \frac{G_T}{AH} \frac{Y_{G \text{ out}} - Y_{G \text{ in}}}{\Delta Y_{\log}} \quad (2.46)$$

and

$$h_V = \frac{G_T c_{wG}}{AH} \frac{T_{G \text{ in}} - T_{G \text{ out}}}{\Delta T_{\log}} \quad (2.47)$$

where

$$\Delta Y_{\log} = \frac{(Y_W - Y_{G \text{ in}}) - (Y_W - Y_{G \text{ out}})}{\ln \frac{Y_W - Y_{G \text{ in}}}{Y_W - Y_{G \text{ out}}}} \quad (2.48)$$

and

$$\Delta T_{\log} = \frac{(T_{G \text{ in}} - T_W) - (T_{G \text{ out}} - T_W)}{\ln \frac{T_{G \text{ in}} - T_W}{T_{G \text{ out}} - T_W}} \quad (2.49)$$

Further parameters can be seen in Figure 2.18 and Figure 2.19. When the saturation is nonadiabatic and

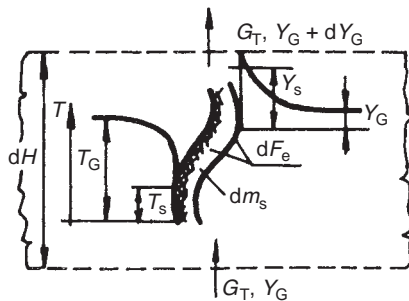


FIGURE 2.18 Elementary part of alfalfa bed.

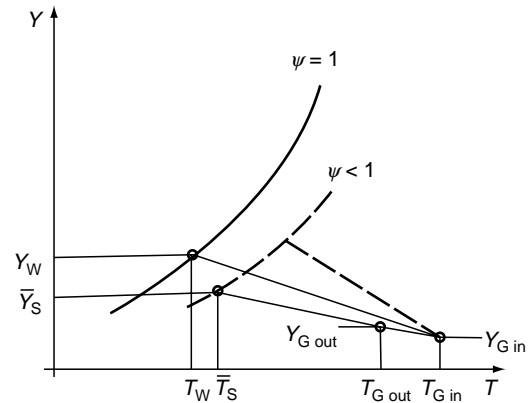


FIGURE 2.19 Saturation of air during circulation through bed.

the bed height is sufficiently small, it is evident according to Ref. [28] that

$$\frac{1}{T_s - T_G} \cong \frac{1}{\bar{T}_s - \bar{T}_G} \quad (2.50)$$

and

$$\left(\frac{1}{Y_s - Y_G} \right) \cong \frac{1}{\bar{Y}_s - \bar{Y}_G} \quad (2.51)$$

The volumetric transport coefficients are in this case

$$K_V(\tau) = \frac{G_T}{A\Delta H} \frac{\Delta Y_G}{\bar{Y}_s - \bar{Y}_G} \quad (2.52)$$

and

$$h_V(\tau) = \frac{G_T c_{wG}}{A\Delta H} \frac{\Delta T_G}{\bar{T}_s - \bar{T}_G} \quad (2.53)$$

In the above correlations,

$$\Delta Y_G = Y_{G \text{ out}} - Y_{G \text{ in}} \quad (2.54)$$

and

$$\Delta T_G = T_{G \text{ out}} - T_{G \text{ in}} \quad (2.55)$$

\bar{T}_s can be calculated by iteration from the correlation

$$\frac{\Delta Y_G}{\Delta T_G} = \frac{\frac{M_V}{M_G} \frac{p_{Ov}^*}{P} \Psi(\bar{T}_s, \bar{X}) - \bar{Y}_G}{\bar{T}_s - \bar{T}_G} \quad (2.56)$$

and with \bar{T}_s known, \bar{Y}_s can also be determined:

$$\bar{Y}_s = \frac{M_V}{M_G} \frac{p_{Ov}^*(\bar{T}_s)}{P} \Psi(\bar{T}_s, \bar{X}) \quad (2.57)$$

2.6 CONCLUSION

Simple experimental methods are discussed in this chapter for determination of the key equilibrium and heat and mass transfer parameters need in engineering calculations for common dryers. The coverage is by no means all-inclusive. The reader is referred to the literature cited for details.

NOMENCLATURE

A	surface, m^2
Bi	Biot number
c_m	moisture capacity, kg/kg
c_{SW}	specific heat capacity of wet material, $J/(kg\ K)$
D_a	effective diffusivity coefficient, m^2/s
D_{VG}	molecular diffusivity coefficient between vapor and gas, m^2/s
$ Fo$	Fourier number
G_T	total gas flow rate, kg/s
G_0	total gas flow, $kg/(m^2\ s)$
h	heat transfer coefficient, $W/(m^2\ K)$
h_m	modified material transfer coefficient, Equation 2.17, $W/(m^2\ K)$
h_v	volumetric heat transfer coefficient, $W/(m^3\ K)$
ΔH	sum of heat of evaporation plus heat of moisture, J/kg
j_m	mass flow density, $kg/(m^2\ s)$
j_q	heat flow density, W/m^2
k_G	mass transfer coefficient, $kmol/(m^2\ s\ Pa)$
K	mass transfer coefficient, $kg/(m^2\ s\ Pa)$
K_G	overall mass transfer coefficient, $kg/(m^2\ s\ Pa)$
K_H	overall heat transfer coefficient, $W/(m^2\ K)$
K_V	volumetric mass transfer coefficient, $kg/(m^2\ s)$
L_{SW}	mass flow of wet material, kg/s
m	mass, kg
n	revolutions per minute, $1/s$
N_W	drying rate, $kg/(m^2\ s)$
N_{We}	constant drying rate, $kg/(m^2\ s)$
p	partial pressure, Pa
p_o	partial vapor pressure of pure component, Pa
Q	heat rate, W
R	gas constant, $J/(kg\ K)$
S	slope of equilibrium curve defined by Equation 2.31, Pa/K
S^*	slope of equilibrium curve, per K
S^{**}	slope of equilibrium curve defined by Equation 2.24, Pa
T	temperature, K
T_m	average temperature, K
X	moisture content of material, dry basis
Y	moisture content of air, dry basis
Z	half thickness of the plane, m
V	volume, m^3

SUBSCRIPTS

d	belonging to the dew point
e	equilibrium value
G	gas
h	heat transfer
in	entering
log	logarithmic mean value
m	mass transfer
0	initial value
out	leaving
s	surface
S	solid
V	vapor
W	wet, wet bulb
Z	in the symmetry plane of a plate
ξ	at the border of the dry and wet layer

SUPERSCRIPTS

$-$	average value
$*$	equilibrium

GREEK SYMBOLS

δ	thickness, m
δ_k	eigenvalue, defined by Equation 2.14
ϕ	fractional holdup of solids
ε	phase-change criterion
k_h	heat diffusivity, m^2/s
k_m	moisture diffusivity coefficient, m^2/s
λ_h	thermal conductivity, $W/(m\ K)$
λ_m	moisture conductivity, $kg/(ms\ K)$
ν	eigenvalue, defined by Equation 2.23
ψ	relative moisture content
μ_D	diffusion-resistance coefficient
ρ_{Os}	bulk density of solids, kg/m^3
τ	time, s, h
ξ	thickness of dry zone, m
Δ	symbol of difference

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3 Basic Process Calculations and Simulations in Drying

Zdzisław Pakowski and Arun S. Mujumdar

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3.1 INTRODUCTION

Since the publication of the first and second editions of this handbook, we have been witnessing a revolution in methods of engineering calculations. Computer tools have become easily available and have replaced the old graphical methods. An entirely new discipline of computer-aided process design (CAPD) has emerged. Today even simple problems are solved using dedicated computer software. The same is not necessarily true for drying calculations; dedicated software for this process is still scarce. However, general computing tools including Excel, Mathcad, MATLAB, and Mathematica are easily available in any engineering company. Bearing this in mind, we have decided to present here a more computer-oriented calculation methodology and simulation methods than to rely on old graphical and shortcut methods. This does not mean that the computer will relieve one from thinking. In this respect, the old simple methods and rules of thumb are still valid and provide a simple common-sense tool for verifying computer-generated results.

3.2 OBJECTIVES

Before going into details of process calculations we need to determine when such calculations are necessary in industrial practice. The following typical cases can be distinguished:

- Design—(a) selection of a suitable dryer type and size for a given product to optimize the capital and operating costs within the range of limits imposed—this case is often termed *process synthesis* in CAPD; (b) specification of all process parameters and dimensioning of a selected dryer type so the set of design parameters or assumptions is fulfilled—this is the common *design problem*.
- Simulation—for a given dryer, calculation of dryer performance including all inputs and outputs, internal distributions, and their time dependence.
- Optimization—in design and simulation an optimum for the specified set of parameters is sought. The objective function can be formu-

lated in terms of economic, quality, or other factors, and restrictions may be imposed on ranges of parameters allowed.

- Process control—for a given dryer and a specified vector of input and control parameters the output parameters at a given instance are sought. This is a special case when not only the accuracy of the obtained results but the required computation time is equally important. Although drying is not always a rapid process, in general for real-time control, calculations need to provide an answer almost instantly. This usually requires a dedicated set of computational tools like neural network models.

In all of the above methods we need *a model* of the process as the core of our computational problem. A model is a set of equations connecting all process parameters and a set of constraints in the form of inequalities describing adequately the behavior of the system. When all process parameters are determined with a probability equal to 1 we have a *deterministic* model, otherwise the model is a *stochastic* one.

In the following sections we show how to construct a suitable model of the process and how to solve it for a given case. We will show only deterministic models of convective drying. Models beyond this range are important but relatively less frequent in practice.

In our analysis we will consider each phase as a continuum unless stated otherwise. In fact, elaborate models exist describing aerodynamics of flow of gas and granular solid mixture where phases are considered noncontinuous (e.g., bubbling bed model of fluid bed, two-phase model for pneumatic conveying, etc.).

3.3 BASIC CLASSES OF MODELS AND GENERIC DRYER TYPES

Two classes of processes are encountered in practice: steady state and unsteady state (batch). The difference can easily be seen in the form of general balance equation of a given entity for a specific volume of space (e.g., the dryer or a single phase contained in it):

$$\text{Inputs} - \text{outputs} = \text{accumulation} \quad (3.1)$$

For instance, for mass flow of moisture in a solid phase being dried (in kg/s) this equation reads:

$$W_S X_1 - W_S X_2 - w_D A = m_S \frac{dX}{d\tau} \quad (3.2)$$

In steady-state processes, as in all continuously operated dryers, the accumulation term vanishes and the balance equation assumes the form of an algebraic equation. When the process is of batch type or when a continuous process is being started up or shut down, the accumulation term is nonzero and the balance equation becomes an ordinary differential equation (ODE) with respect to time.

In writing Equation 3.1, we have assumed that only the input and output parameters count. Indeed, when the volume under consideration is perfectly mixed, all phases inside this volume will have the same property as that at the output. This is the principle of a lumped parameter model (LPM).

If a property varies continuously along the flow direction (in one dimension for simplicity), the balance equation can only be written for a differential space element. Here Equation 3.2 will now read

$$W_S X - W_S \left(X + \frac{\partial X}{\partial l} dl \right) - w_D dA = dm_S \frac{\partial X}{\partial \tau} \quad (3.3)$$

or, after substituting $dA = a_v S dl$ and $dm_S = (1 - \varepsilon) \rho_S S dl$, we obtain

$$-W_S \frac{\partial X}{\partial l} - w_D a_v S = (1 - \varepsilon) \rho_S S \frac{\partial X}{\partial \tau} \quad (3.4)$$

As we can see for this case, which we call a distributed parameter model (DPM), in steady state (in the one-dimensional case) the model becomes an ODE with respect to space coordinate, and in unsteady state it

becomes a partial differential equation (PDE). This has a far-reaching influence on methods of solving the model. A corresponding equation will have to be written for yet another phase (gaseous), and the equations will be coupled by the drying rate expression.

Before starting with constructing and solving a specific dryer model it is recommended to classify the methods, so typical cases can easily be identified. We will classify typical cases when a solid is contacted with a heat carrier. Three factors will be considered:

1. Operation type—we will consider either batch or continuous process with respect to given phase.
2. Flow geometry type—we will consider only parallel flow, cocurrent, countercurrent, and cross-flow cases.
3. Flow type—we will consider two limiting cases, either plug flow or perfectly mixed flow.

These three assumptions for two phases present result in 16 generic cases as shown in Figure 3.1. Before constructing a model it is desirable to identify the class to which it belongs so that writing appropriate model equations is facilitated.

Dryers of type 1 do not exist in industry; therefore, dryers of type 2 are usually called *batch dryers* as is done in this text. An additional term—*semicontinuous*—will be used for dryers described in Section 3.9. Their principle of operation is different from any of the types shown in Figure 3.1.

3.4 GENERAL RULES FOR A DRYER MODEL FORMULATION

When trying to derive a model of a dryer we first have to identify a volume of space that will represent a dryer.

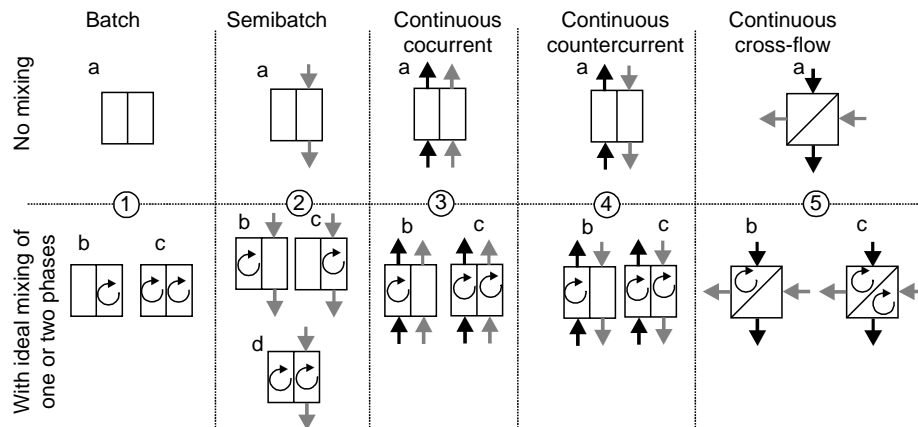


FIGURE 3.1 Generic types of dryers.

If a dryer or a whole system is composed of many such volumes, a separate submodel will have to be built for each volume and the models connected together by streams exchanged between them. Each stream entering the volume must be identified with parameters. Basically for systems under constant pressure it is enough to describe each stream by the name of the component (humid gas, wet solid, condensate, etc.), its flowrate, moisture content, and temperature. All heat and other energy fluxes must also be identified.

The following five parts of a deterministic model can usually be distinguished:

1. Balance equations—they represent Nature's laws of conservation and can be written in the form of Equation 3.1 (e.g., for mass and energy).
2. Constitutive equations (also called kinetic equations)—they connect fluxes in the system to respective driving forces.
3. Equilibrium relationships—necessary if a phase boundary exists somewhere in the system.
4. Property equations—some properties can be considered constant but, for example, saturated water vapor pressure is strongly dependent on temperature even in a narrow temperature range.
5. Geometric relationships—they are usually necessary to convert flowrates present in balance equations to fluxes present in constitutive equations. Basically they include flow cross-section, specific area of phase contact, etc.

Typical formulation of basic model equations will be summarized later.

3.4.1 MASS AND ENERGY BALANCES

Input–output balance equations for a typical case of convective drying and LPM assume the following form:

3.4.1.1 Mass Balances

Solid phase:

$$W_S X_1 - W_S X_2 - w_{Dm} A = m_S \frac{dX}{d\tau} \quad (3.5)$$

Gas phase:

$$W_B Y_1 - W_B Y_2 + w_{Dm} A = m_B \frac{dY}{d\tau} \quad (3.6)$$

3.4.1.2 Energy balances

Solid phase:

$$W_S i_{m1} - W_S i_{m2} + (\Sigma q_m - w_{Dm} h_A) A = m_S \frac{di_m}{d\tau} \quad (3.7)$$

Gas phase:

$$W_B i_{g1} - W_B i_{g2} - (\Sigma q_m - w_{Dm} h_A) A = m_B \frac{di_g}{d\tau} \quad (3.8)$$

In the above equations Σq_m and w_{Dm} are a sum of mean interfacial heat fluxes and a drying rate, respectively. Accumulation in the gas phase can almost always be neglected even in a batch process as small compared to accumulation in the solid phase. In a continuous process the accumulation in solid phase will also be neglected.

In the case of DPMs for a given phase the balance equation for property Γ reads:

$$\begin{aligned} \text{div}[\Gamma \cdot u] - \text{div}[D \cdot \text{grad } \Gamma] \pm \beta a_V \Delta \Gamma \pm G \\ - \frac{\partial \Gamma}{\partial \tau} = 0 \end{aligned} \quad (3.9)$$

where the LHS terms are, respectively (from the left): convective term, diffusion (or axial dispersion) term, interfacial term, source or sink (production or destruction) term, and accumulation term.

This equation can now be written for a single phase for the case of mass and energy transfer in the following way:

$$\text{div}[\rho X \cdot u] - \text{div}[D \cdot \text{grad}(\rho X)] \pm k_X a_V \Delta X - \frac{\partial \rho X}{\partial \tau} = 0 \quad (3.10)$$

$$\begin{aligned} \text{div}[\rho c_m T \cdot u] - \text{div}\left[\frac{\lambda}{\rho c_m} \cdot \text{grad}(\rho c_m T)\right] \\ \pm \alpha a_V \Delta T + q_{ex} - \frac{\partial \rho c_m T}{\partial \tau} = 0 \end{aligned} \quad (3.11)$$

Note that density here is related to the whole volume of the phase: e.g., for solid phase composed of granular material it will be equal to $\rho_m(1 - \varepsilon)$. Moreover, the interfacial term is expressed here as $k_X a_V \Delta X$ for consistency, although it is expressed as $k_Y a_V \Delta Y$ elsewhere (see Equation 3.27).

Now, consider a one-dimensional parallel flow of two phases either in co- or countercurrent flow, exchanging mass and heat with each other. Neglecting diffusional (or dispersion) terms, in steady state the balance equations become

$$W_S \frac{dX}{dl} = -w_D a_V S \quad (3.12)$$

$$\pm W_B \frac{dY}{dl} = w_D a_V S \quad (3.13)$$

$$W_S \frac{di_m}{dl} = (q - w_D h_{Av}) a_V S \quad (3.14)$$

$$\pm W_B \frac{di_g}{dl} = -(q - w_D h_{Av}) a_V S \quad (3.15)$$

where the LHSs of Equation 3.13 and Equation 3.15 carry the positive sign for cocurrent and the negative sign for countercurrent operation. Both heat and mass fluxes, q and w_D , are calculated from the constitutive equations as explained in the following section. Having in mind that

$$\frac{di_g}{dl} = (c_B + c_A Y) \frac{dt_g}{dl} + (c_A t_g + \Delta h_{v0}) \frac{dY}{dl} \quad (3.16)$$

and that enthalpy of steam emanating from the solid is

$$h_{Av} = c_A t_m + \Delta h_{v0} \quad (3.17)$$

we can now rewrite (Equation 3.12 through Equation 3.15) in a more convenient working version

$$\frac{dX}{dl} = -\frac{S}{W_S} w_D a_V \quad (3.18)$$

$$\frac{dt_m}{dl} = \frac{S}{W_S} \frac{a_V}{c_S + c_{Al} X} [q + w_D ((c_{Al} - c_A) t_m - \Delta h_{v0})] \quad (3.19)$$

$$\frac{dY}{dl} = \frac{1}{\chi} \frac{S}{W_B} w_D a_V \quad (3.20)$$

$$\frac{dt_g}{dl} = -\frac{1}{\chi} \frac{S}{W_B} \frac{a_V}{c_B + c_A Y} [q + w_D c_A (t_g - t_m)] \quad (3.21)$$

where χ is 1 for cocurrent and -1 for countercurrent operation.

For a monolithic solid phase convective and interfacial terms disappear and in unsteady state, for the one-dimensional case, the equations become

$$D_{\text{eff}} \frac{\partial^2 X}{\partial x^2} = \frac{\partial X}{\partial \tau} \quad (3.22)$$

$$\lambda \frac{\partial^2 t_m}{\partial x^2} = c_p \rho_m \frac{\partial t}{\partial \tau} \quad (3.23)$$

These equations are named Fick's law and Fourier's law, respectively, and can be solved with suitable boundary and initial conditions. Literature on solving

these equations is abundant, and for diffusion a classic work is that of Crank (1975). It is worth mentioning that, in view of irreversible thermodynamics, mass flux is also due to thermodiffusion and barodiffusion. Formulation of Equation 3.22 and Equation 3.23 containing terms of thermodiffusion was favored by Luikov (1966).

3.4.2 CONSTITUTIVE EQUATIONS

They are necessary to estimate either the local non-convective fluxes caused by conduction of heat or diffusion of moisture or the interfacial fluxes exchanged either between two phases or through system boundaries (e.g., heat losses through a wall). The first are usually expressed as

$$q = -\lambda \frac{dt}{dl} \quad (3.24)$$

$$j = -\rho D_{\text{eff}} \frac{dX}{dl} \quad (3.25)$$

and they are already incorporated in the balance equations (3.22 and 3.23). The interfacial flux equations assume the following form:

$$q = \alpha (t_g - t_m) \quad (3.26)$$

$$w_D = k_Y \phi (Y^* - Y) \quad (3.27)$$

where ϕ is

$$\phi = \frac{M_A/M_B}{Y^* - Y} \ln \left(1 + \frac{Y^* - Y}{M_A/M_B + Y} \right) \quad (3.28)$$

While the convective heat flux expression is straightforward, the expression for drying rate needs explanation. The drying rate can be calculated from this formula, when drying is controlled by gas-side resistance. The driving force is then the difference between absolute humidity at equilibrium with solid surface and that of bulk gas. When solid surface is saturated with moisture, the expression for Y^* is identical to Equation 3.48; when solid surface contains bound moisture, Y^* will result from Equation 3.46 and a sorption isotherm. This is in essence the so-called equilibrium method of drying rate calculation.

When the drying rate is controlled by diffusion in the solid phase (i.e., in the falling drying rate period), the conditions at solid surface are difficult to find, unless we are solving the DPM (Fick's law or equivalent) for the solid itself. Therefore, if the solid itself has lumped parameters, its drying rate must be represented by an empirical expression. Two forms are commonly used.

3.4.2.1 Characteristic Drying Curve

In this approach the measured drying rate is represented as a function of the actual moisture content (normalized) and the drying rate in the constant drying rate period:

$$w_D = w_{Df}f(\Phi) \quad (3.29)$$

The f function can be represented in various forms to fit the behavior of typical solids. The form proposed by Langrish et al. (1991) is particularly useful. They split the falling rate periods into two segments (as it often occurs in practice) separated by Φ_B . The equations are:

$$\begin{aligned} f &= \Phi_B^{a-c} & \text{for } \Phi \leq \Phi_B \\ f &= \Phi^a & \text{for } \Phi > \Phi_B \end{aligned} \quad (3.30)$$

Figure 3.2 shows the form of a possible drying rate curve using Equation 3.30.

Other such equations also exist in the literature (e.g., Halström and Wimmerstedt, 1983; Nijdam and Keey, 2000).

3.4.2.2 Kinetic Equation (e.g., Thin-Layer Equations)

In agricultural sciences it is common to present drying kinetics in the form of the following equation:

$$\Phi = f(\tau, \text{process parameters}) \quad (3.31)$$

The function f is often established theoretically, for example, when using the drying model formulated by Lewis (1921)

$$\frac{dX}{d\tau} = k(X - X^*) \quad (3.32)$$

After integration one obtains

$$\Phi = \exp(-k\tau) \quad (3.33)$$

A similar equation can be obtained by solving Fick's equation in spherical geometry:

$$\Phi = \frac{6}{\pi^2} \sum_{n=1}^{\infty} \frac{1}{n^2} \exp\left(-n^2 \pi^2 \frac{D_{\text{eff}}}{R^2} \tau\right) \quad (3.34)$$

By truncating the RHS side one obtains

$$\Phi = \frac{6}{\pi^2} \exp\left(-\pi^2 \frac{D_{\text{eff}}}{R^2} \tau\right) = a \exp(-k\tau) \quad (3.35)$$

This equation was empirically modified by Page (1949), and is now known as the Page equation:

$$\Phi = \exp(-k\tau^n) \quad (3.36)$$

A collection of such equations for popular agricultural products is contained in Jayas et al. (1991). Other process parameters such as air velocity, temperature, and humidity are often incorporated into these equations.

The volumetric drying rate, which is necessary in balance equations, can be derived from the TLE in the following way:

$$\begin{aligned} w_D a_V &= -\frac{m_S}{A} a_V \frac{d\Phi}{d\tau} (X_c - X^*) \\ &= -\frac{m_S}{V} \frac{d\Phi}{d\tau} (X_c - X^*) \end{aligned} \quad (3.37)$$

while

$$m_S = V(1 - \varepsilon)\rho_S \quad (3.38)$$

and

$$w_D a_V = -(1 - \varepsilon)\rho_S (X_c - X^*) \frac{d\Phi}{d\tau} \quad (3.39)$$

The drying rate ratio of CDC is then calculated as

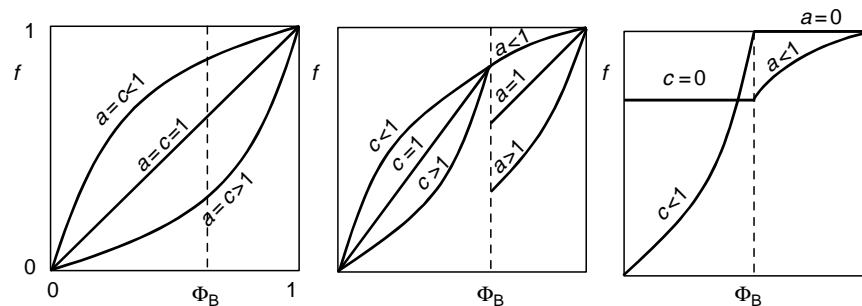


FIGURE 3.2 The influence of parameters a and c of Equation 3.30 on CDC shape.

$$f = -\frac{(1-\varepsilon)\rho_s(X_c - X^*)}{k_Y\phi(Y^* - Y)a_V} \frac{d\Phi}{d\tau} \quad (3.40)$$

To be able to calculate the volumetric drying rate from TLE, one needs to know the voidage ε and specific contact area a_V in the dryer.

When dried solids are monolithic or grain size is overly large, the above lumped parameter approximations of drying rate would be unacceptable, in which case a DPM represents the entire solid phase. Such models are shown in [Section 3.7](#).

3.4.3 AUXILIARY RELATIONSHIPS

3.4.3.1 Humid Gas Properties and Psychrometric Calculations

The ability to perform psychrometric calculations forms a basis on which all drying models are built. One principal problem is how to determine the solid temperature in the constant drying rate conditions.

In psychrometric calculations we consider thermodynamics of three phases: inert gas phase, moisture vapor phase, and moisture liquid phase. Two gaseous phases form a solution (mixture) called humid gas. To determine the degree of complexity of our approach we will make the following assumptions:

- Inert gas component is insoluble in the liquid phase
- Gaseous phase behavior is close to ideal gas; this limits our total pressure range to less than 2 bar

- Liquid phase is incompressible
- Components of both phases do not chemically react with themselves

Before writing the psychrometric relationships we will first present the necessary approximating equations to describe physical properties of system components.

Dependence of saturated vapor pressure on temperature (e.g., Antoine equation):

$$\ln p_s = A - \frac{B}{C + t} \quad (3.41)$$

Dependence of latent heat of vaporization on temperature (e.g., Watson equation):

$$\Delta h_v = H(t - t_{\text{ref}})^n \quad (3.42)$$

Dependence of specific heat on temperature for vapor phase—polynomial form:

$$c_A = c_{A0} + c_{A1}t + c_{A2}t^2 + c_{A3}t^3 \quad (3.43)$$

Dependence of specific heat on temperature for liquid phase—polynomial form:

$$c_{Al} = c_{Al0} + c_{Al1}t + c_{Al2}t^2 + c_{Al3}t^3 \quad (3.44)$$

Table 3.1 contains coefficients of the above listed property equations for selected liquids and [Table 3.2](#) for gases. These data can be found in specialized books (e.g., Reid et al., 1987; Yaws, 1999) and computerized data banks for other liquids and gases.

TABLE 3.1
Coefficients of Approximating Equations for Properties of Selected Liquids

Property		Water	Ethanol	Isopropanol	Toluene
Molar mass, kg/kmol	M_A	18.01	46.069	60.096	92.141
Saturated vapor pressure, kPa	A	16.376953	16.664044	18.428032	13.998714
	B	3878.8223	3667.7049	4628.9558	3096.52
	C	229.861	226.1864	252.636	219.48
Heat of vaporization, kJ/kg	H	352.58	110.17	104.358	47.409
	t_{ref}	374.14	243.1	235.14	318.8
	n	0.33052	0.4	0.371331	0.38
Specific heat of vapor, kJ/(kg K)	c_{A0}	1.883	0.02174	0.04636	−0.4244
	$c_{A1} \times 10^3$	−0.16737	5.662	5.95837	6.2933
	$c_{A2} \times 10^6$	0.84386	−3.4616	−3.54923	−3.9623
	$c_{A3} \times 10^9$	−0.26966	0.8613	−16.3354	0.93604
Specific heat of liquid, kJ/(kg K)	c_{Al0}	2.822232	−1.4661	5.58272	−0.61169
	$c_{Al1} \times 10^2$	1.182771	4.0052	−4.6261	1.9192
	$c_{Al2} \times 10^4$	−0.350477	−1.5863	1.701	−0.56354
	$c_{Al3} \times 10^8$	3.60107	22.873	−16.3354	5.9661

TABLE 3.2
Coefficients of Approximating Equations for Properties of Selected Gases

Property		Air	Nitrogen	CO ₂
Molar mass, kg/kmol	M_B	28.9645	28.013	44.010
Specific heat of gas, kJ/(kg K)	c_{B0}	1.02287	1.0566764	0.48898
	$c_{B1} \times 10^3$	-0.5512	-0.197286	1.46505
	$c_{B2} \times 10^6$	0.181871	0.49471	-0.94562
	$c_{B3} \times 10^9$	-0.05122	-0.18832	0.23022

3.4.3.2 Relations between Absolute Humidity, Relative Humidity, Temperature, and Enthalpy of Humid Gas

With the above assumptions and property equations we can use Equation 3.45 through Equation 3.47 for calculating these basic relationships (note that moisture is described as component A and inert gas as component B).

Definition of relative humidity φ (we will use here φ defined as decimal fraction instead of RH given in percentage points):

$$\varphi(t) = p/p_s(t) \quad (3.45)$$

Relation between absolute and relative humidities:

$$Y = \frac{M_A}{M_B} \frac{\varphi p_s(t)}{P_0 - \varphi p_s(t)} \quad (3.46)$$

Definition of enthalpy of humid gas (per unit mass of dry gas):

$$i_g = (c_A Y + c_B)t + \Delta h_{v0} Y \quad (3.47)$$

Equation 3.46 and Equation 3.47 are sufficient to find any two missing humid gas parameters from Y , φ , t , i_g , if the other two are given. These calculations were traditionally done graphically using a psychrometric chart, but they are easy to perform numerically. When solving these equations one must remember that resulting Y for a given t must be lower than that at saturation, otherwise the point will represent a fog (supersaturated condition), not humid gas.

3.4.3.3 Calculations Involving Dew-Point Temperature, Adiabatic-Saturation Temperature, and Wet-Bulb Temperature

Dew-point temperature (DPT) is the temperature reached by humid gas when it is cooled until it

becomes saturated (i.e., $\varphi = 1$). From Equation 3.46 we obtain

$$Y_s = \frac{M_A}{M_B} \frac{p_s(t)}{P_0 - p_s(t)} \quad (3.48)$$

To find DPT when Y is known this equation must be solved numerically. On the other hand, the inverse problem is trivial and requires substituting DPT into Equation 3.48.

Adiabatic-saturation temperature (AST) is the temperature reached when adiabatically contacting limited amounts of gas and liquid until equilibrium. The suitable equation is

$$\frac{i_g - i_{gs,AST}}{Y - Y_{s,AST}} = c_{Al} t_{AS} \quad (3.49)$$

Wet-bulb temperature (WBT) is the one reached by a small amount of liquid exposed to an infinite amount of humid gas in steady state. The following are the governing equations.

- For water–air system, approximately

$$\frac{t - t_{WB}}{Y - Y_{s,WBT}} = - \frac{\Delta h_{v,WBT}}{c_H} \quad (3.50)$$

where

$$c_H(t) = c_A(t)Y + c_B(t) \quad (3.51)$$

Incidentally, this equation is equivalent to Equation 3.49 (see Treybal, 1980) for air and water vapor system.

- For other systems with higher Lewis numbers the deviation of WBT from AST is noticeable and can reach several degrees Celsius, thus causing serious errors in drying rate estimation. For such systems the following equation is recommended (Keey, 1978):

$$\frac{t - t_{WB}}{Y - Y_{s,WBT}} = -\frac{\Delta h_{v,WBT}}{c_H} Le^{-2/3} \phi \quad (3.52)$$

Typically in the wet-bulb calculations the following two situations are common:

- One searches for humidity of gas of which both dry- and wet-bulb temperatures are known: it is enough to substitute relationships for Y_s , Δh_v , and c_H into Equation 3.52 and solve it for Y .
- One searches for WBT once dry-bulb temperature and humidity are known: the same substitutions are necessary but now one solves the resulting equation for WBT.

The Lewis number

$$Le = \frac{\lambda_g}{c_p \rho_g D_{AB}} \quad (3.53)$$

is defined usually for conditions midway of the convective boundary layer. Recent investigations (Berg et al., 2002) indicate that Equation 3.52 needs corrections to become applicable to systems of high WBT approaching boiling point of liquid. However, for common engineering applications it is usually sufficiently accurate.

Over a narrow temperature range, e.g., for water–air system between 0 and 100°C, to simplify calculations one can take constant specific heats equal to $c_A = 1.91$ and $c_B = 1.02$ kJ/(kg K). In all calculations involving enthalpy balances specific heats are averaged between the reference and actual temperature.

3.4.3.4 Construction of Psychrometric Charts

Construction of psychrometric charts by computer methods is common. Three types of charts are most popular: Grosvenor chart, Grosvenor (1907) (or the psychrometric chart), Mollier chart, Mollier (1923) (or enthalpy-humidity chart), and Salin chart (or deformed enthalpy-humidity chart); these are shown schematically in Figure 3.3.

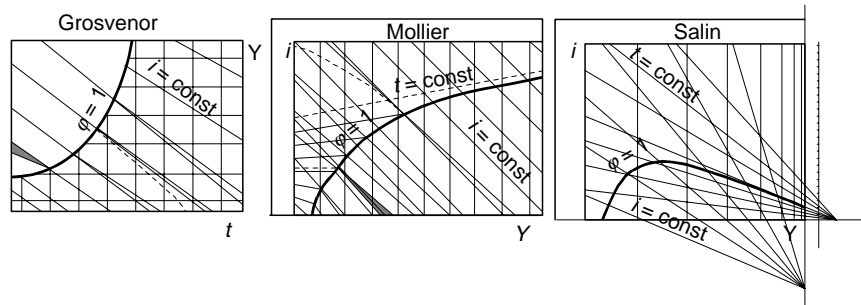


FIGURE 3.3 Schematics of the Grosvenor, Mollier, and Salin charts.

Since the Grosvenor chart is plotted in undistorted Cartesian coordinates, plotting procedures are simple. Plotting methods are presented and charts of high accuracy produced as explained in Shallcross (1994). Procedures for the Mollier chart plotting are explained in Pakowski (1986) and Pakowski and Mujumdar (1987), and those for the Salin chart in Soininen (1986).

It is worth stressing that computer-generated psychrometric charts are used mainly as illustration material for presenting computed results or experimental data. They are now seldom used for graphical calculation of dryers.

3.4.3.5 Wet Solid Properties

Humid gas properties have been described together with humid gas psychrometry. The pertinent data for wet solid are presented below.

Sorption isotherms of the wet solid are, from the point of view of model structure, equilibrium relationships, and are a property of the solid–liquid–gas system. For the most common air–water system, sorption isotherms are, however, traditionally considered as a solid property. Two forms of sorption isotherm equations exist—explicit and implicit:

$$\phi^* = f(t, X) \quad (3.54)$$

$$X^* = f(t, a_w) \quad (3.55)$$

where a_w is the water activity and is practically equivalent to ϕ . The implicit equation, favored by food and agricultural sciences, is of little use in dryer calculations unless it can be converted to the explicit form. In numerous cases it can be done analytically. For example, the GAB equation

$$X^* = \frac{a\phi}{(1 - b\phi)(1 + c\phi)} \quad (3.56)$$

can be solved analytically for ϕ , and when the wrong root is rejected, the only solution is

$$\varphi^* = \frac{-\left(\frac{a}{X} + b - c\right) + \sqrt{\left(\frac{a}{X} + b - c\right)^2 + 4bc}}{2bc} \quad (3.57)$$

Numerous sorption isotherm equations (of approximately 80 available) cannot be analytically converted to the explicit form. In this case they have to be solved numerically for φ^* each time Y^* is computed, i.e., at every drying rate calculation. This slows down computations considerably.

Sorptional capacity varies with temperature, and the thermal effect associated with this phenomenon is isosteric heat of sorption, which can be numerically calculated using the Clausius–Clapeyron equation

$$\Delta h_s = -\frac{R}{M_A} \left[\frac{d \ln \varphi}{d(1/T)} \right]_{X = \text{const}} \quad (3.58)$$

If the sorption isotherm is temperature-independent the heat of sorption is zero; therefore a number of sorption isotherm equations used in agricultural sciences are useless from the point of view of dryer calculations unless drying is isothermal. It is noteworthy that in the model equations derived in this section the heat of sorption is neglected, but it can easily be added by introducing Equation 3.59 for the solid enthalpy in energy balances of the solid phase.

Wet solid enthalpy (per unit mass of dry solid) can now be defined as

$$i_m = (c_s + c_{Al}X)t_m - \Delta h_s X \quad (3.59)$$

The specific heat of dry solid c_s is usually presented as a polynomial dependence of temperature.

Diffusivity of moisture in the solid phase due to various governing mechanisms will be here termed as an effective diffusivity. It is often presented in the Arrhenius form of dependence on temperature

$$D_{\text{eff}} = D_0 \exp\left(-\frac{E_a}{RT}\right) \quad (3.60)$$

However, it also depends on moisture content. Various forms of dependence of D_{eff} on t and X are available (e.g., Marinos-Kouris and Maroulis, 1995).

3.4.4 PROPERTY DATABASES

As in all process calculations, reliable property data are essential (but not a guarantee) for obtaining sound results. For drying, three separate databases are necessary: for liquids (moisture), for gases, and for solids. Data for gases and liquids are widespread and are easily available in printed form (e.g., Yaws, 1999) or in electronic version. Relatively good property

prediction methods exist (Reid et al., 1987). However, when it comes to solids, we are almost always confronted with a problem of availability of property data. Only a few source books exist with data for various products (Nikitina, 1968; Ginzburg and Savina, 1982; Iglesias and Chirife, 1984). Some data are available in this handbook also. However, numerous data are spread over technical literature and require a thorough search. Finally, since solids are not identical even if they represent the same product, it is always recommended to measure all the required properties and fit them with necessary empirical equations.

The following solid property data are necessary for an advanced dryer design:

- Specific heat of bone-dry solid
- Sorption isotherm
- Diffusivity of water in solid phase
- Shrinkage data
- Particle size distribution for granular solids

3.5 GENERAL REMARKS ON SOLVING MODELS

Whenever an attempt to solve a model is made, it is necessary to calculate the degrees of freedom of the model. It is defined as

$$N_D = N_V - N_E \quad (3.61)$$

where N_V is the number of variables and N_E the number of independent equations. It applies also to models that consist of algebraic, differential, integral, or other forms of equations. Typically the number of variables far exceeds the number of available equations. In this case several selected variables must be made constants; these selected variables are then called *process variables*. The model can be solved only when its degrees of freedom are zero. It must be borne in mind that not all vectors of process variables are valid or allow for a successful solution of the model.

To solve models one needs appropriate tools. They are either specialized for the specific dryer design or may have a form of universal mathematical tools. In the second case, certain experience in handling these tools is necessary.

3.6 BASIC MODELS OF DRYERS IN STEADY STATE

3.6.1 INPUT–OUTPUT MODELS

Input–output models are suitable for the case when both phases are perfectly mixed (cases 3c, 4c, and 5c

in Figure 3.1), which almost never happens. On the other hand, this model is very often used to represent a case of unmixed flows when there is lack of a DPM. Input–output modeling consists basically of balancing all inputs and outputs of a dryer and is often performed to identify, for example, heat losses to the surroundings, calculate performance, and for dryer audits in general.

For a steady-state dryer balancing can be made for the whole dryer only, so the system of Equation 3.5 through Equation 3.8 now consists of only two equations

$$W_S(X_1 - X_2) = W_B(Y_2 - Y_1) \quad (3.62)$$

$$W_S(i_{m2} - i_{m1}) = W_B(i_{g1} - i_{g2}) + q_c - q_l + \Delta q_t + q_m \quad (3.63)$$

where subscripts on heat fluxes indicate: c, indirect heat input; l, heat losses; t, net heat carried in by transport devices; and m, mechanical energy input. Let us assume that all q , W_S , W_B , X_1 , i_{m1} , Y_1 , i_{g1} are known as in a typical design case. The remaining variables are X_2 , Y_2 , i_{m2} , and i_{g2} . Since we have two equations, the system has two degrees of freedom and cannot be solved unless two other variables are set as process parameters. In design we can assume X_2 since it is a design specification, but then one extra parameter must be assumed. This of course cannot be done rationally, unless we are sure that the process runs in constant drying rate period—then i_{m2} can be calculated from WBT. Otherwise, we must look for other equations, which could be the following:

$$W_S(X_1 - X_2) = V a_Y k_Y \Delta Y_m \quad (3.64)$$

$$W_S(i_{m2} - i_{m1}) = V a_Y (\alpha \Delta t_m + \Sigma q - k_Y \Delta Y_m h_A) \quad (3.65)$$

Provided that we know all kinetic data, a_Y , k_Y , and α , these two equations carry only one new variable V since temperatures can be derived from suitable enthalpies. Provided that we know how to calculate the averaged driving forces, the model now can be solved and exit stream parameters and volume of the dryer calculated. The success, however, depends on how well we can estimate the averaged driving forces.

3.6.2 DISTRIBUTED PARAMETER MODELS

3.6.2.1 Cocurrent Flow

For cocurrent operation (case 3a in Figure 3.1) both the case design and simulation are simple. The four balance equations (3.18 through 3.21) supplemented by a suitable drying rate and heat flux equations are solved starting at inlet end of the dryer, where all boundary conditions (i.e., all parameters of incoming streams) are defined. This situation is shown in Figure 3.4.

In the case of design the calculations are terminated when the design parameter, usually final moisture content, is reached. Distance at this point is the required dryer length. In the case simulation the calculations are terminated once the dryer length is reached.

Parameters of both gas and solid phase (represented by gas in equilibrium with the solid surface) can be plotted in a psychrometric chart as process paths. These phase diagrams (no timescale is available there) show schematically how the process goes on.

To illustrate the case the model composed of Equation 3.18 through Equation 3.21, Equation 3.26, and Equation 3.27 is solved for a set of typical conditions and the results are shown in Figure 3.5.

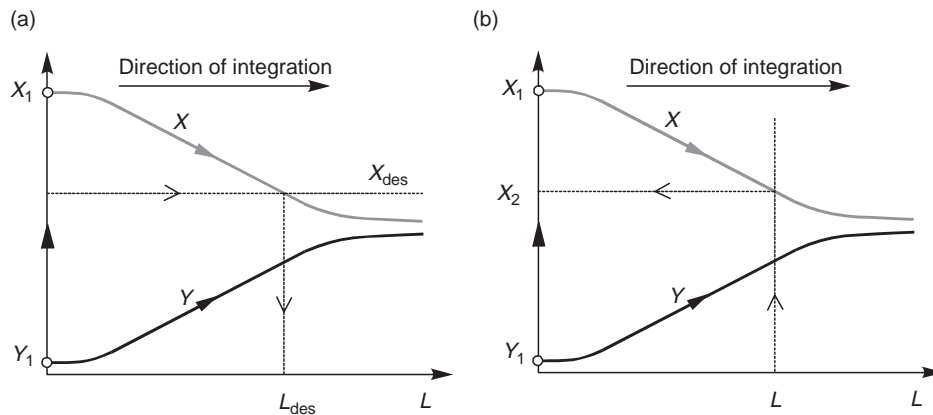


FIGURE 3.4 Schematic of design and simulation in cocurrent case: (a) design; (b) simulation. X_{des} is the design value of final moisture content.

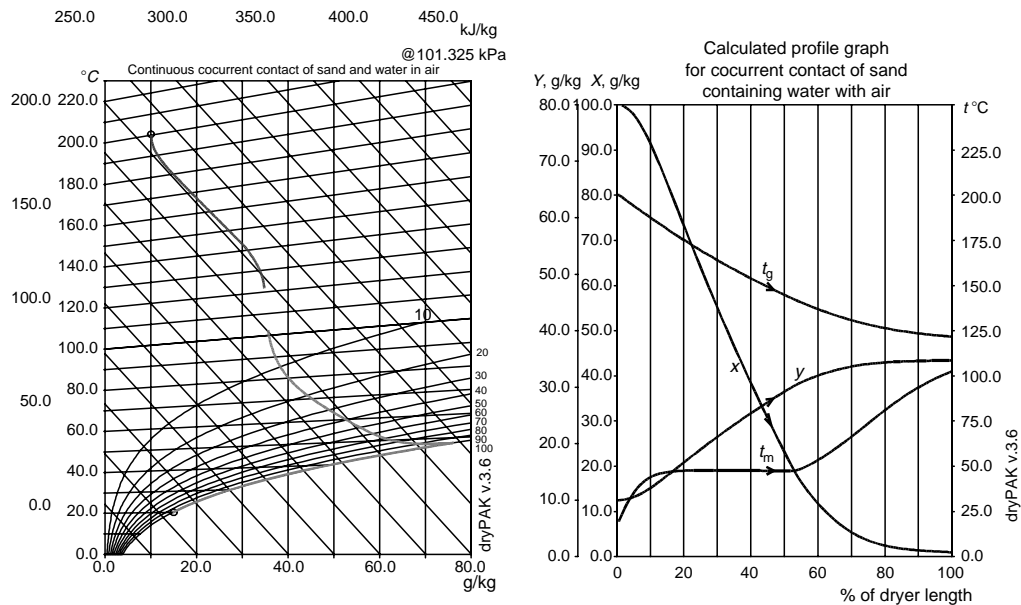


FIGURE 3.5 Process paths and longitudinal distribution of parameters for cocurrent drying of sand in air.

3.6.2.2 Countercurrent Flow

The situation in countercurrent case (case 4a in Figure 3.1) design and simulation is shown in Figure 3.6. In both cases we see that boundary conditions are defined at opposite ends of the integration domain. It leads to the *split boundary value* problem.

In design this problem can be avoided by using the design parameters for the solid specified at the exit end. Then, by writing input–output balances over the whole dryer, inlet parameters of gas can easily be found (unless local heat losses or other distributed parameter phenomena need also be considered). However, in simulation the split boundary value

problem exists and must be solved by a suitable numerical method, e.g., the *shooting method*. Basically the method consists of assuming certain parameters for the exiting gas stream and performing integration starting at the solid inlet end. If the gas parameters at the other end converges to the known inlet gas parameters, the assumption is satisfactory; otherwise, a new assumption is made. The process is repeated under control of a suitable convergence control method, e.g., Wegstein. Figure 3.7 contains a sample countercurrent case calculation for the same material as that used in Figure 3.5.

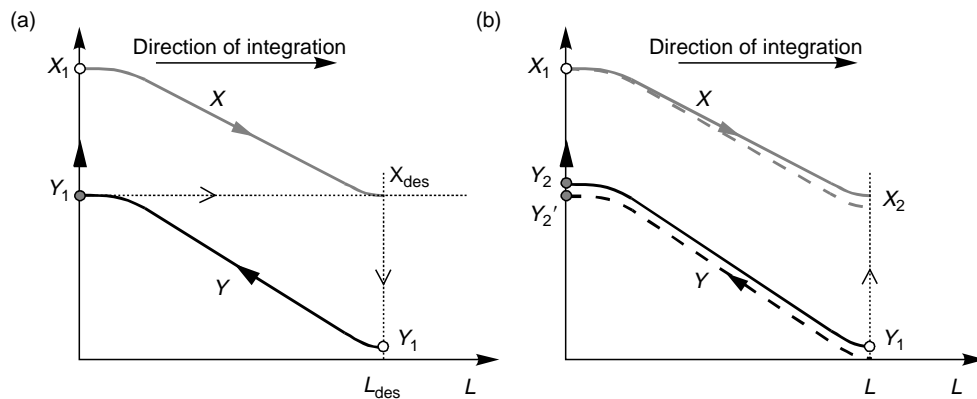


FIGURE 3.6 Schematic of design and simulation in cocurrent case: (a) design—split boundary value problem is avoided by calculating Y_1 from the overall mass balance; (b) simulation—split boundary value problem cannot be avoided, broken line shows an unsuccessful iteration, solid line shows a successful iteration—with Y_2 assumed the Y profile converged to Y_1 .

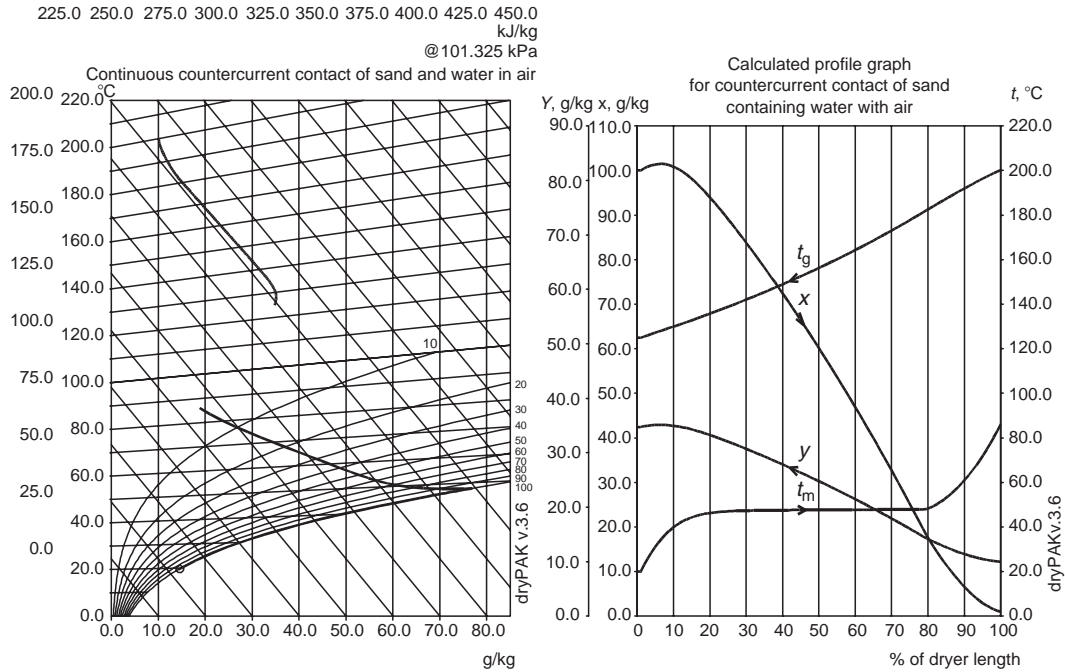


FIGURE 3.7 Process paths and longitudinal distribution of parameters for countercurrent drying of sand in air.

3.6.2.3 Cross-Flow

3.6.2.3.1 Solid Phase is One-Dimensional

This is a simple case corresponding to case 5b of Figure 3.1. By assuming that the solid phase is perfectly mixed in the direction of gas flow, the solid phase becomes one-dimensional. This situation occurs with a continuous plug-flow fluid bed dryer. Schematic of an element of the dryer length with finite thickness Δl is shown in Figure 3.8.

The balance equations for the solid phase can be derived from Equation 3.12 and Equation 3.14 of the parallel flow:

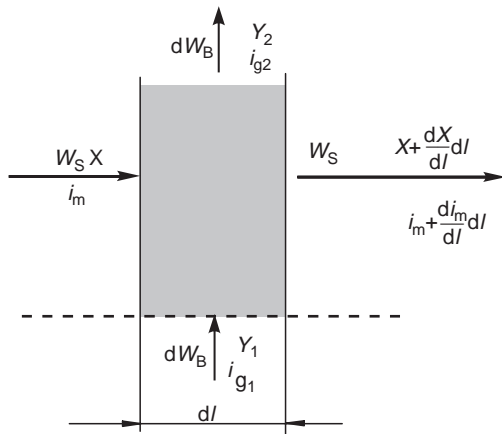


FIGURE 3.8 Element of a cross-flow dryer.

$$\frac{W_S}{S} \frac{dX}{dl} = -w_D a_V \quad (3.66)$$

$$\frac{W_S}{S} \frac{di_m}{dl} = (q - w_D h_{Av}) a_V \quad (3.67)$$

The analogous equations for the gas phase are:
mass balance

$$\frac{1}{S} \frac{dW_B(Y_2 - Y_1)}{dl} = w_D a_V \quad (3.68)$$

energy balance

$$\frac{1}{S} \frac{dW_B(i_{g2} - i_{g1})}{dl} = -(q - w_D h_{Av}) a_V \quad (3.69)$$

In the case of an equilibrium method of calculation of the drying rate the kinetic equations are:

$$w_D = k_Y \Delta Y_m \quad (3.70)$$

$$q = \alpha \Delta t_m \quad (3.71)$$

In other models (CDC and TLE) the drying rate will be modified as shown in Section 3.4.2.

Since the heat and mass coefficients can be defined on the basis of either the inlet driving force or the mean logarithmic driving force, ΔY_m and Δt_m are calculated respectively as

$$\Delta Y_m = (Y^* - Y_1) \quad (3.72)$$

or

$$\Delta Y_m = \frac{Y_2 - Y_1}{\ln\left(\frac{Y^* - Y_1}{Y^* - Y_2}\right)} \quad (3.73)$$

$$\Delta t_m = (t_m - t_{g1}) \quad (3.74)$$

or

$$\Delta t_m = \frac{t_{g2} - t_{g1}}{\ln\left(\frac{t_m - t_{g2}}{t_m - t_{g1}}\right)} \quad (3.75)$$

To solve Equation 3.68 and Equation 3.69 one needs to assume a uniform distribution of gas over the whole length of the dryer, and therefore

$$\frac{dW_B}{dl} = \frac{W_B}{L} \quad (3.76)$$

When the algebraic Equation 3.68 and Equation 3.69 are solved to obtain the exiting gas parameters Y_2 and i_{g2} , one can plug the LHS of these equations into Equation 3.66 and Equation 3.67 to obtain

$$\frac{dX}{dl} = \frac{1}{W_S} \frac{W_B}{L} (Y_2 - Y_1) \quad (3.77)$$

$$\frac{di_m}{dl} = \frac{1}{W_S} \frac{W_B}{L} (i_{g2} - i_{g1}) \quad (3.78)$$

The following equations can easily be integrated starting at the solids inlet. In Figure 3.9 sample process parameter profiles along the dryer are shown.

Cross-flow drying in a plug-flow, continuous fluid bed is a case when axial dispersion of flow is often considered. Let us briefly present a method of solving

this case. First, the governing balance equations for the solid phase will have the following form derived from Equation 3.10 and Equation 3.11

$$u_m \frac{dX}{dl} = E_m \frac{d^2 X}{dl^2} - \frac{a_V w_D}{\rho_S (1 - \varepsilon)} \quad (3.79)$$

$$u_m \frac{di_m}{dl} = E_h \frac{d^2 i_m}{dl^2} + \frac{a_V (q - w_D h_{AV})}{\rho_S (1 - \varepsilon)} \quad (3.80)$$

or

$$u_m \frac{dt_m}{dl} = E_h \frac{d^2 t_m}{dl^2} + \frac{a_V}{\rho_S (1 - \varepsilon)} \frac{1}{c_S + c_{Al} X} \times [q + ((c_{Al} - c_A)t_m - \Delta h_{v0})w_D] \quad (3.81)$$

where

$$u_m = \frac{W_S}{\rho_S (1 - \varepsilon)} \quad (3.82)$$

These equations are supplemented by equations for w_D and q according to Equation 3.70 and Equation 3.71. It is a common assumption that $E_m = E_h$, because in fluid beds they result from longitudinal mixing by rising bubbles. Boundary conditions (BCs) assume the following form:

At $l = 0$

$$X = X_0 \quad \text{and} \quad i_m = i_{m0} \quad (3.83)$$

At $l = L$

$$\frac{dX}{dl} = 0 \quad \text{and} \quad \frac{di_m}{dl} = 0 \quad (3.84)$$

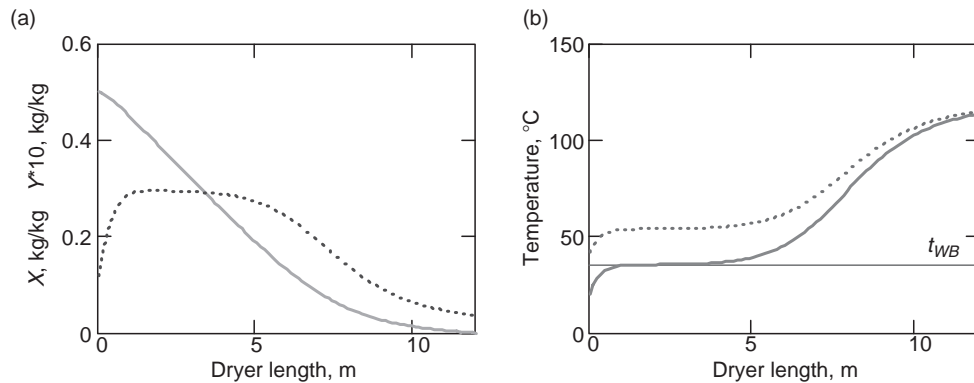


FIGURE 3.9 Longitudinal parameter distribution for a cross-flow dryer with one-dimensional solid flow. Drying of a moderately hygroscopic solid: (a) material moisture content (solid line) and local exit air humidity (broken line); (b) material temperature (solid line) and local exit air temperature (broken line). t_{WB} is wetbulb temperature of the incoming air.

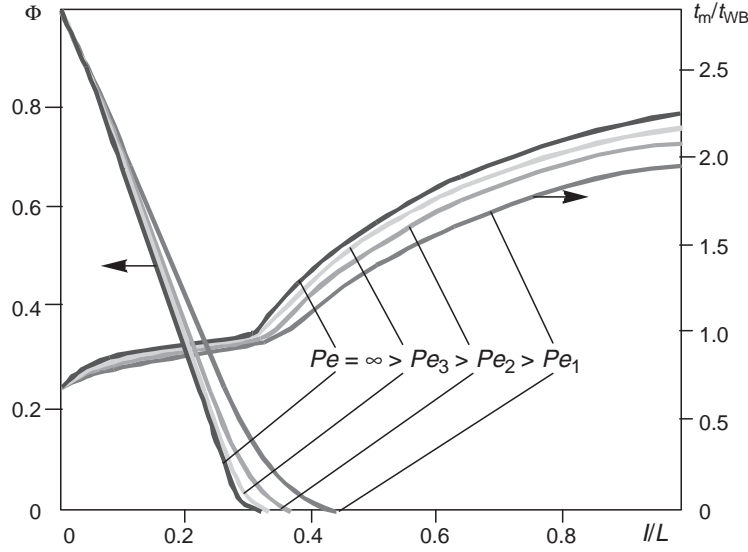


FIGURE 3.10 Sample profiles of material moisture content and temperature for various Pe numbers.

The second BC is due to Danckwerts and has been used for chemical reactor models. This leads, of course, to a split boundary value problem, which needs to be solved by an appropriate numerical technique. The resulting longitudinal profiles of solid moisture content and temperature in a dryer for various Peclet numbers ($Pe = u_m L/E$) are presented in Figure 3.10.

As one can see, only at low Pe numbers, profiles differ significantly. When $Pe > 0.5$, the flow may be considered a plug-flow.

3.6.2.3.2 Solid Phase is Two-Dimensional

This case happens when solid phase is not mixed but moves as a block. This situation happens in certain dryers for wet grains. The model must be derived for differential bed element as shown in Figure 3.11.

The model equations are now:

$$\frac{dX}{dl} = -\frac{w_D a_V}{\sigma_H} \quad (3.85)$$

$$\frac{dY}{dh} = \frac{w_D a_V}{\sigma_L} \quad (3.86)$$

$$\frac{dt_m}{dl} = \frac{a_V}{\sigma_H} \frac{1}{c_S + c_{A1} X} \times [q - ((c_A - c_{A1})t_m + \Delta h_{v0})w_D] \quad (3.87)$$

$$\frac{dt_g}{dh} = -\frac{a_V}{\sigma_L} \frac{1}{c_B + c_A Y} [q + c_A(t_g - t_m)w_D] \quad (3.88)$$

The symbols σ_H and σ_L are flow densities per 1 m for solid and gas mass flowrates, respectively, and are defined as follows:

$$\sigma_H = \frac{dW_S}{dh} = \frac{W_S}{H} \quad (3.89)$$

$$\sigma_L = \frac{dW_B}{dl} = \frac{W_B}{L} \quad (3.90)$$

The third term in these formulations applies when distribution of flow is uniform, otherwise an adequate distribution function must be used. An exemplary model solution is shown in Figure 3.12. The solution only presents the heat transfer case (cooling of granular solid with air), so mass transfer equations are neglected.

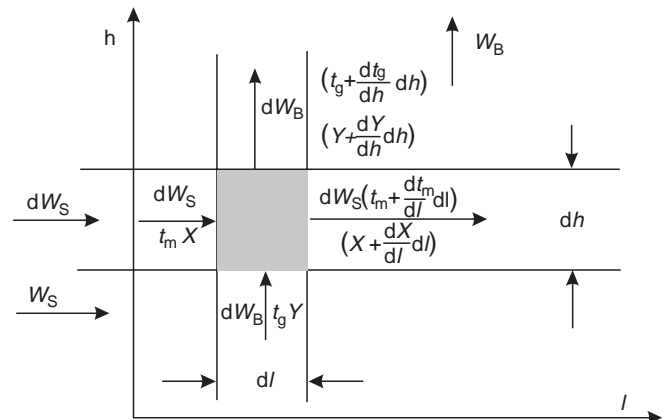


FIGURE 3.11 Schematic of a two-dimensional cross-flow dryer.

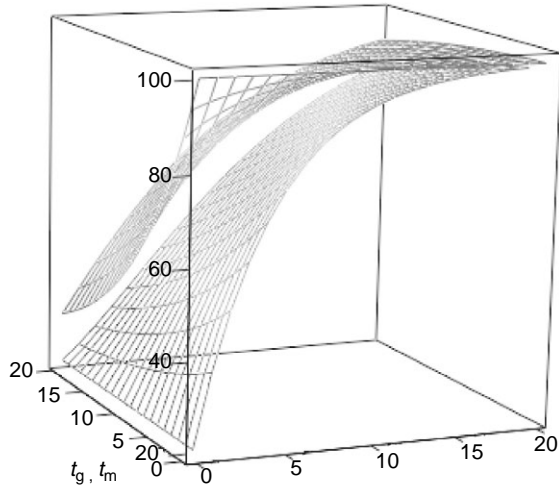


FIGURE 3.12 Solution of a two-dimensional cross-flow dryer model for cooling of granular solid with hot air. Solid flow enters through the front face of the cube, gas flows from left to right. Upper surface, solid temperature; lower surface, gas temperature.

3.7 DISTRIBUTED PARAMETER MODELS FOR THE SOLID

This case occurs when dried solids are monolithic or have large grain size so that LPM for the drying rate would be an unacceptable approximation. To answer the question as to whether this case applies one has to calculate the Biot number for mass transfer. It is recommended to calculate it from Equation 3.100 since various definitions are found in the literature. When $Bi < 1$, the case is externally controlled and no DPM for the solid is required.

3.7.1 ONE-DIMENSIONAL MODELS

3.7.1.1 NONSHRINKING SOLIDS

Assuming that moisture diffusion takes place in one direction only, i.e., in the direction normal to surface for plate and in radial direction for cylinder and sphere, and that no other way of moisture transport exists but diffusion, the following second Fick's law may be derived

$$\frac{\partial X}{\partial \tau} = \frac{1}{r^n} \frac{\partial}{\partial r} \left[r^n D_{\text{eff}}(t_m, X) \frac{\partial X}{\partial r} \right] \quad (3.91)$$

where $n = 0$ for plate, 1 for cylinder, 2 for sphere, and r is current distance (radius) measured from the solid center. This parameter reaches a maximum value of R , i.e., plate is $2R$ thick if dried at both sides.

Initially we assume that moisture content is uniformly distributed and the initial solid moisture content is X_0 . To solve Equation 3.91 one requires a set of BCs. For high Bi numbers ($Bi > 100$) BC is called BC of the first kind and assumes the following form at the solid surface:

At $r = R$

$$X = X^*(\tau, Y) \quad (3.92)$$

For moderate Bi numbers ($1 < Bi < 100$) it is known as BC of the third kind and assumes the following form:

At $r = R$

$$-D_{\text{eff}} \rho_m \left(\frac{\partial X}{\partial r} \right)_i = k_Y [Y^*(X, t)_i - Y] \quad (3.93)$$

where subscript i denotes the solid–gas interface. BC of the second kind as known from calculus (constant flux at the surface)

At $r = R$

$$w_{Di} = \text{const} \quad (3.94)$$

has little practical interest and can be incorporated in BC of the third kind. Quite often (here as well), therefore, BC of the third kind is named BC of the second kind. Additionally, at the symmetry plane we have

At $r = 0$

$$\frac{\partial X}{\partial r} = 0 \quad (3.95)$$

When solving the Fick's equation with constant diffusivity it is recommended to convert it to a dimensionless form. The following dimensionless variables are introduced for this purpose:

$$\Phi = \frac{X - X^*}{X_c - X^*}, \quad Fo = \frac{D_{\text{eff}} \tau}{R^2}, \quad \zeta = \frac{r}{R} \quad (3.96)$$

In the nondimensional form Fick's equation becomes

$$\frac{\partial \Phi}{\partial Fo} = \frac{1}{\zeta^n} \frac{\partial}{\partial \zeta} \left[\zeta^n \frac{D_{\text{eff}}}{D_{\text{eff}0}} \frac{\partial \Phi}{\partial \zeta} \right] \quad (3.97)$$

and the BCs assume the following form:

BC I BC II

$$\text{at } \zeta = 1, \quad \Phi = 0 \left(\frac{\partial \Phi}{\partial \zeta} \right)_i + Bi_D^* \Phi = 0 \quad (3.98)$$

$$\text{at } \zeta = 0, \quad \frac{\partial \Phi}{\partial \zeta} = 0 \quad \frac{\partial \Phi}{\partial r} = 0 \quad (3.99)$$

where

$$Bi_D^* = m_{XY} \frac{k_Y \phi R}{D_{\text{eff}} \rho_m} \quad (3.100)$$

is the modified Biot number in which m_{XY} is a local slope of equilibrium curve given by the following expression:

$$m_{XY} = \frac{Y^*(X, t_m)_i - Y}{X - X^*} \quad (3.101)$$

The diffusional Biot number modified by the m_{XY} factor should be used for classification of the cases instead of $Bi_D = k_Y R / (D_{\text{eff}} \rho_m)$ encountered in several texts. Note that due to dependence of D_{eff} on X Biot number can vary during the course of drying, thus changing classification of the problem.

Since drying usually proceeds with varying external conditions and variable diffusivity, analytical solutions will be of little interest. Instead we suggest using a general-purpose tool for solving parabolic (Equation 3.97) and elliptic PDE in one-dimensional geometry like the *pdepe* solver of MATLAB. The result for $Bi_D^* = 5$ obtained with this tool is shown in Figure 3.13. The results were obtained for isothermal conditions. When conditions are nonisothermal, a question arises as to whether it is necessary to simultaneously solve Equation 3.22 and Equation 3.23. Since Biot numbers for mass transfer far exceed those for heat transfer, usually the problem of heat transfer is purely external,

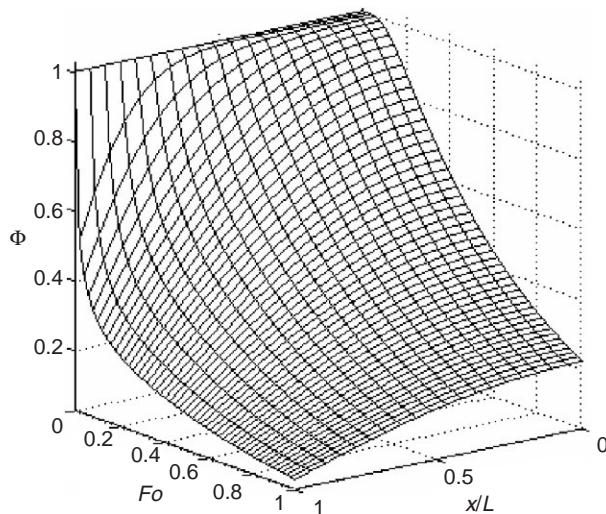


FIGURE 3.13 Solution of the DPM isothermal drying model of one-dimensional plate by *pdepe* solver of MATLAB. Finite difference discretization by uniform mesh both for space and time, $Bi_D^* = 5$. Fo is dimensionless time, x/L is dimensionless distance.

and internal profiles of temperature are almost flat. This allows one to use LPM for the energy balance. Therefore, to monitor the solid temperature it is enough to supplement Equation 3.22 with the following energy balance equation:

$$\frac{dt_m}{d\tau} = \frac{A}{m_s} \frac{1}{c_s + c_{Al} \bar{X}} [q + ((c_A - c_{Al})t_m + \Delta h_{v0})w_D] \quad (3.102)$$

If Equation 3.22 and Equation 3.23 must be solved simultaneously, the problem becomes stiff and requires specialized solvers.

3.7.1.2 Shrinking Solids

3.7.1.2.1 Unrestrained Shrinkage

When solids shrink volumetrically (majority of food products does), their volume is usually related to moisture content by the following empirical law:

$$V = V_s(1 + sX) \quad (3.103)$$

If one assumes that, for instance, a plate shrinks only in the direction of its thickness, the following relationship may be deduced from the above equation:

$$R = R_s(1 + sX) \quad (3.104)$$

where R is the actual plate thickness and R_s is the thickness of absolutely dry plate.

In Eulerian coordinates, shrinking causes an advective mass flux, which is difficult to handle. By changing the coordinate system to Lagrangian, i.e., the one connected with dry mass basis, it is possible to eliminate this flux. This is the principle of a method proposed by Kechaou and Roques (1990). In Lagrangian coordinates Equation 3.91 for one-dimensional shrinkage of an infinite plate becomes:

$$\frac{\partial X}{\partial \tau} = \frac{\partial}{\partial \zeta} \left[\frac{D_{\text{eff}}}{(1 + sX)^2} \frac{\partial X}{\partial \zeta} \right] \quad (3.105)$$

All boundary and initial conditions remain but the BC of Equation 3.94 now becomes

$$\left(\frac{\partial X}{\partial \zeta} \right)_{\zeta=R_s} = - \frac{(1 + sX)^2}{\rho_s D_{\text{eff}}} k_Y (Y^* - Y) \quad (3.106)$$

In Equation 3.105 and Equation 3.106, ζ is the Lagrangian space coordinate, and it changes from 0 to R_s . For the above case of one-dimensional shrinkage the relationship between r and ζ is identical to that in Equation 3.104:

$$r = \zeta(1 + sX) \quad (3.107)$$

The model was proved to work well for solids with $s > 1$ (gelatin, polyacrylamide gel). An exemplary solution of this model for a shrinking gelatin film is shown in Figure 3.14.

3.7.1.2.2 Restrained Shrinkage

For many materials shrinkage accompanying the drying process may be opposed by the rigidity of the solid skeleton or by viscous forces in liquid phase as it is compressed by shrinking external layers. This results in development of stress within the solid. The development of stress is interesting from the point of view of possible damage of dried product by deformation or cracking. In order to account for this, new equations have to be added to Equation 3.10 and Equation 3.11. These are the balance of force equation and liquid moisture flow equation written as

$$G\nabla^2 \mathbf{U} + \frac{G}{1-2\nu} \nabla e - \alpha \nabla p = 0 \quad (3.108)$$

$$\frac{k}{\mu_{Al}} \nabla^2 p = \frac{1}{Q} \frac{\partial p}{\partial \tau} + \alpha \frac{\partial e}{\partial \tau} \quad (3.109)$$

where \mathbf{U} is the deformation matrix, e is strain tensor element, and p is internal pressure (Q and α are constants). The equations were developed by Biot and are explained in detail by Hasatani and Itaya (1996). Equation 3.108 and Equation 3.109 can be

solved together with Equation 3.10 and Equation 3.11 provided that a suitable rheological model of the solid is known. The solution is almost always obtained by the finite element method due to inevitable deformation of geometry. Solution of such problems is complex and requires much more computational power than any other problem in this section.

3.7.2 TWO- AND THREE-DIMENSIONAL MODELS

In fact some supposedly three-dimensional cases can be converted to one-dimensional by transformation of the coordinate system. This allows one to use a finite difference method, which is easy to program. Lima et al. (2001) show how ovoid solids (e.g., cereal grains, silkworm cocoons) can be modeled by a one-dimensional model. This even allows for uniform shrinkage to be considered in the model. However, in the case of two- and three-dimensional models when shrinkage is not negligible, the finite difference method can no longer be used. This is due to unavoidable deformation of corner elements, as shown in Figure 3.15.

The finite element methods have been used instead for two- and three-dimensional shrinking solids (see Perre and Turner, 1999, 2000). So far no commercial software was proven to be able to handle drying problems in this case and all reported simulations were performed by programs individually written for the purpose.

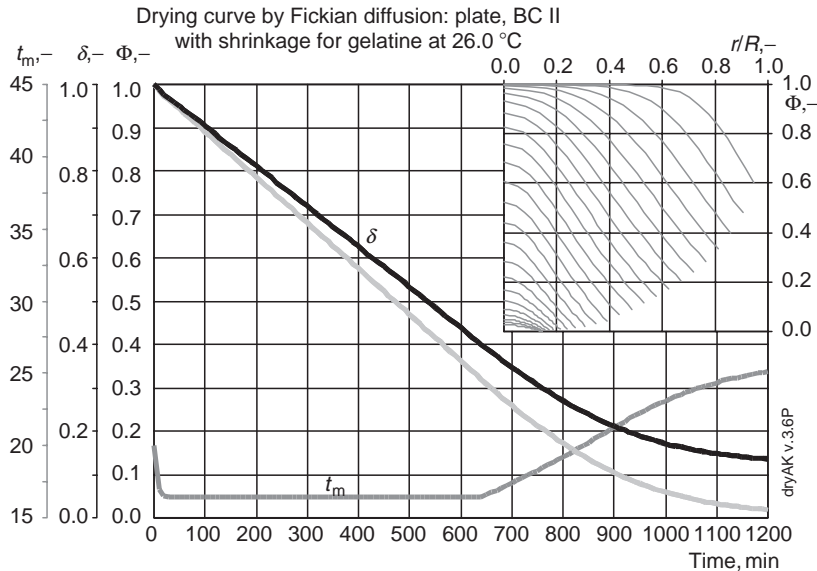


FIGURE 3.14 Solution of a model of drying for a shrinking solid. Gelatin plate 3-mm thick, initial moisture content 6.55 kg/kg. Shrinkage coefficient $s = 1.36$. Main plot shows dimensionless moisture content Φ , dimensionless thickness $\delta = R/R_0$, solid temperature t_m . Insert shows evolution of the internal profiles of Φ .

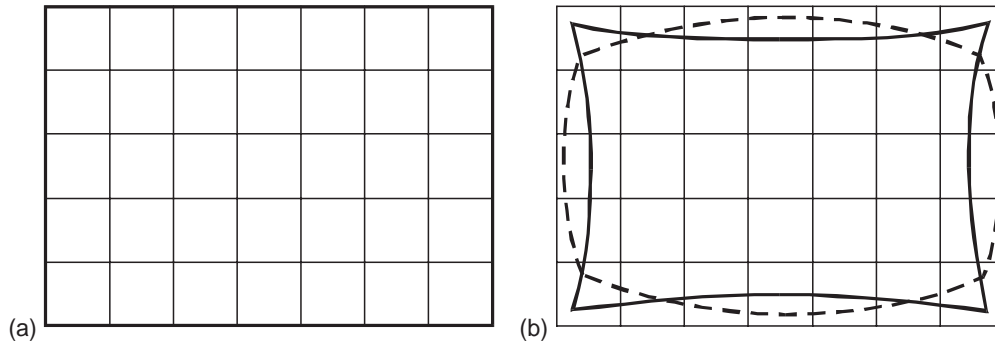


FIGURE 3.15 Finite difference mesh in the case two-dimensional drying with shrinkage: (a) before deformation; (b) after deformation. Broken line—for unrestrained shrinkage, solid line—for restrained shrinkage.

3.7.3 SIMULTANEOUS SOLVING DPM OF SOLIDS AND GAS PHASE

Usually in texts the DPM for solids (e.g., Fick's law) is solved for constant external conditions of gas. This is especially the case when analytical solutions are used. As the drying progresses, the external conditions change. At present with powerful ODE integrators there is essentially only computer power limit for simultaneously solving PDEs for the solid and ODEs for the gas phase. Let us discuss the case when spherical solid particles flow in parallel to gas stream exchanging mass and heat.

The internal mass transfer in the solid phase described by Equation 3.91 will be discretized by a finite difference method into the following set of equations

$$\frac{dX_i}{d\tau} = f(X_{i-1}, X_i, X_{i+1}, v) \quad \text{for } i = 1, \dots, \text{number of nodes} \quad (3.110)$$

where X_i is the moisture content at a given node and v is the vector of process parameters. We will add Equation 3.19 through Equation 3.21 to this set. In the last three equations the space increment dl can be converted to time increment by

$$dl = \frac{S(1 - \varepsilon)\rho_m}{W_s} d\tau \quad (3.111)$$

The resulting set of ODEs can be solved by any ODE solver. The drying rate can be calculated between time steps (Equation 3.112) from temporal change of space-averaged moisture content. As a result one obtains simultaneously spatial profiles of moisture content in the solid as well as longitudinal distribution of parameters in the gas phase. Exemplary results are

shown for cocurrent flash drying of spherical particles in Figure 3.16.

3.8 MODELS FOR BATCH DRYERS

We will not discuss here cases pertinent to startup or shutdown of typically continuous dryers but concentrate on three common cases of batch dryers. In batch drying the definition of drying rate, i.e.,

$$w_D = -\frac{m_s}{A} \frac{dX}{d\tau} \quad (3.112)$$

provides a basis for drying time computation.

3.8.1 BATCH-DRYING OVEN

The simplest batch dryer is a tray dryer shown in Figure 3.17. Here wet solid is placed in thin layers on trays and on a truck, which is then loaded into the dryer.

The fan is started and a heater power turned on. A certain air ventilation rate is also determined. Let us assume that the solid layer can be described by an LPM. The same applies to the air inside the dryer; because of internal fan, the air is well mixed and the case corresponds to case 2d in Figure 3.1. Here, the air humidity and temperature inside the dryer will change in time as well as solid moisture content and temperature. The resulting model equations are therefore

$$m_s \frac{dX}{d\tau} = -w_D A \quad (3.113)$$

$$W_B Y_0 - W_B Y = m_s \frac{dX}{d\tau} + m_B \frac{dY}{d\tau} \quad (3.114)$$

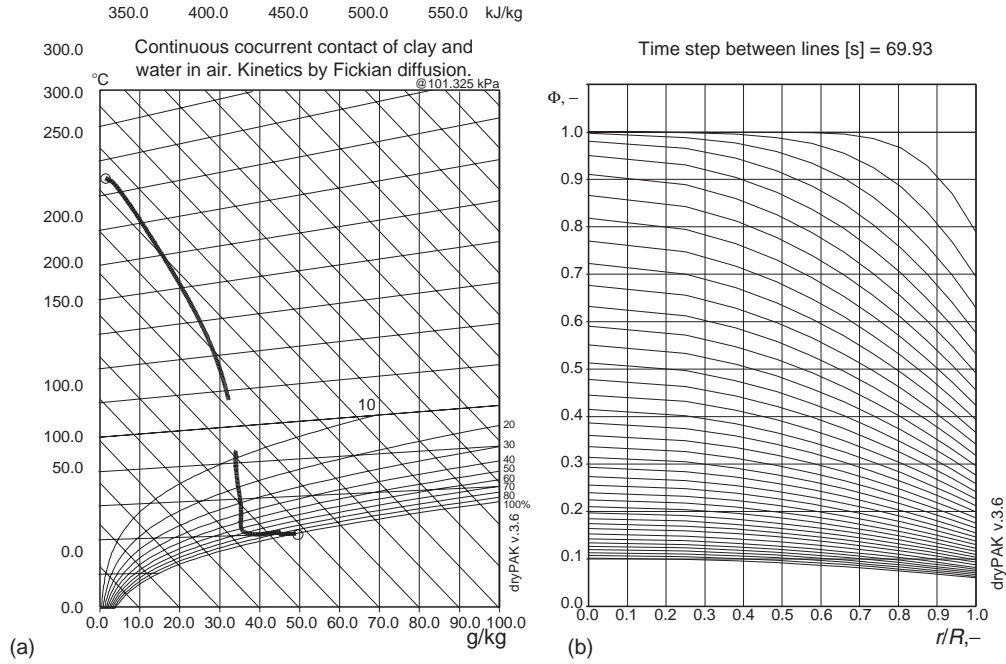


FIGURE 3.16 Cocurrent drying of clay spheres $d = 10$ mm in air at $t_g = 250^\circ\text{C}$. Solid throughput 0.1 kg/s, air throughput 0.06 kg/s. Simultaneous solution for gas phase and solid phase: (a) process trajectories—solid is represented by air in equilibrium with surface; (b) internal moisture distribution profiles.

$$m_S \frac{di_m}{d\tau} = (q - w_D h_{Av})A \quad (3.115)$$

$$W_B i_{g0} - W_B i_g + \Sigma q = m_S \frac{di_m}{d\tau} + m_B \frac{di_g}{d\tau} \quad (3.116)$$

Note that Equation 3.113 is in fact the drying rate definition (Equation 112). In writing these equations we assume that the stream of air exiting the dryer has

the same parameters as the air inside—this is a result of assuming perfect mixing of the air.

This system of equations is mathematically *stiff* because changes of gas parameters are much faster than changes in solid due to the small mass of gas in the dryer. It is advisable to neglect accumulation in the gas phase and assume that gas phase instantly follows changes of other parameters. Equation 3.114 and Equation 3.116 will now have an asymptotic form of algebraic equations. Equation 3.113 through Equation 3.116 can now be converted to the following working form:

$$\frac{dX}{d\tau} = -w_D \frac{A}{m_S} \quad (3.117)$$

$$W_B(Y_0 - Y) + w_D A = 0 \quad (3.118)$$

$$\frac{dt_m}{d\tau} = \frac{1}{c_S + c_{Al}X} \frac{A}{m_S} [q + w_D((c_{Al} - c_A)t_m - \Delta h_{v0})] \quad (3.119)$$

$$W_B[(c_B + c_A Y_0)t_{g0} - (c_B + c_A Y)t_{g0} + (Y - Y_0)c_A t_g] - A[q + w_D c_A(t_g - t_m)] + \Sigma q = 0 \quad (3.120)$$

The system of equations (Equation 3.117 and Equation 3.119) is then solved by an ODE solver for a given set of data and initial conditions. For each time step air parameters Y and t_g are found by solving

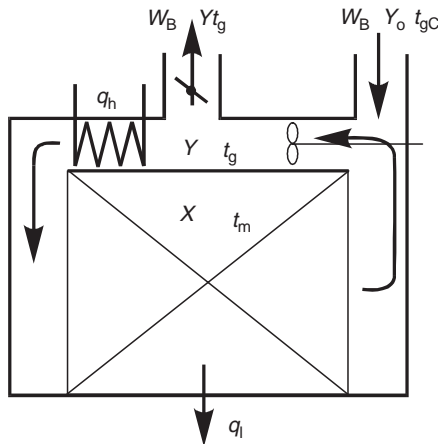


FIGURE 3.17 Schematic of a batchdrying oven.

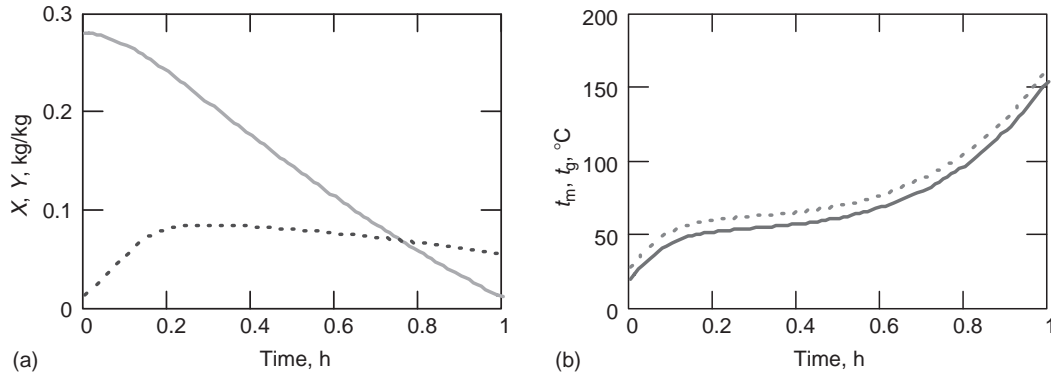


FIGURE 3.18 Solution of a batch oven dryer model—solid dry mass is 90 kg, internal heater power is 20 kW and air ventilation rate is 0.1 kg/s (dry basis); external air humidity is 2 g/kg and temperature 20°C: (a) moisture content X (solid line) and air humidity Y (broken line); (b) material temperature t_m (solid line) and air temperature t_g (broken line).

Equation 3.118 and Equation 3.120. Sample simulation results for this case are plotted in Figure 3.18. Note that at the end of drying, the temperature in the dryer increases excessively due to constant power being supplied to the internal heater. The model may serve as a tool to control the process, e.g., increase the ventilation rate W_B when drying becomes too slow or reduce the heater power when temperature becomes too high as in this case.

3.8.2 BATCH FLUID BED DRYING

In this case the solid phase may be considered as perfectly mixed, so it will be described by an input–output model with accumulation term. On the other hand, the gas phase changes its parameters progressively as it travels through the bed. This situation is shown in Figure 3.19.

Therefore, gas phase will be described by a DPM with no accumulation and the solid phase will be described by an LPM with an accumulation term. The resulting equations are:

$$\frac{dX}{d\tau} = -\frac{a_v}{(1-\varepsilon)\rho_s} \frac{1}{H} \int_0^H w_D dh \quad (3.121)$$

$$\frac{dY}{dh} = \frac{S}{W_B} w_D a_v \quad (3.122)$$

$$\begin{aligned} \frac{dt_m}{d\tau} = & \frac{a_v}{(1-\varepsilon)\rho_s} \frac{1}{c_s + c_{Al}X} \frac{1}{H} \\ & \times \int_0^H [q - ((c_A - c_{Al})t_m + \Delta h_{v0})w_D]dh \end{aligned} \quad (3.123)$$

$$\frac{dt_g}{dh} = -\frac{S}{W_B} a_v \frac{1}{c_B + c_A Y} [q + c_A(t_g - t_m)w_D] \quad (3.124)$$

Equation 3.122 and Equation 3.124 for the gas phase serve only to compute distributions of Y and t_g along bed height, which is necessary to calculate q and w_D . They can easily be integrated numerically, e.g., by the Euler method, at each time step. The integrals in Equation 3.121 and Equation 3.123 can be numerically calculated, e.g., by the trapezoidal rule. This allows Equation 3.121 and Equation 3.123 to be solved by any ODE solver. The model has been solved for a sample case and the results are shown in Figure 3.20.

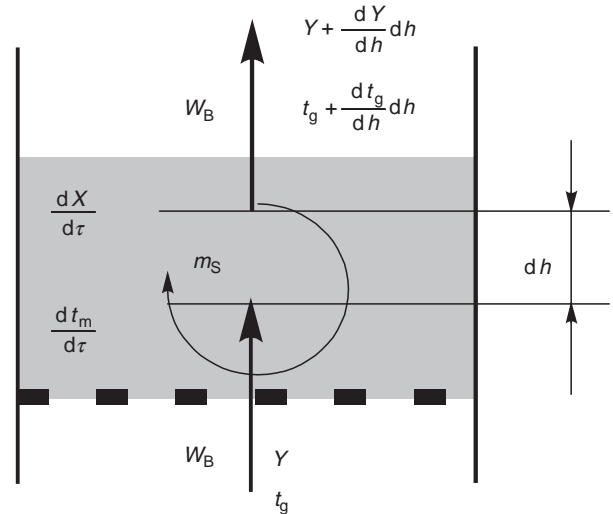


FIGURE 3.19 Schematic of a batch fluid bed dryer.

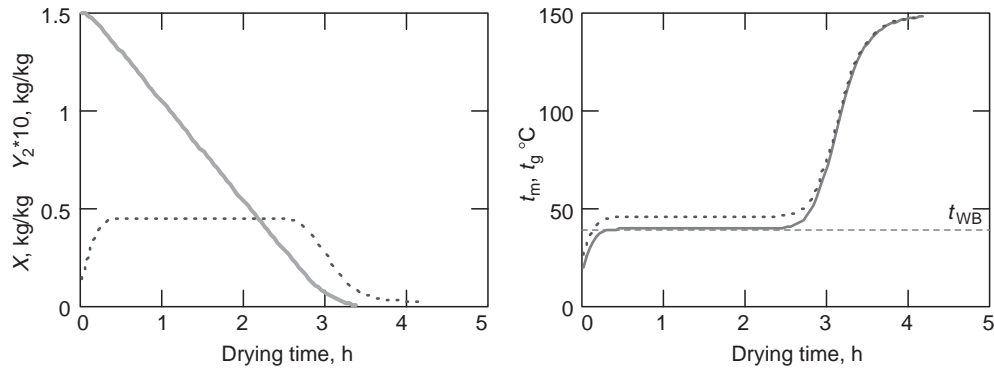


FIGURE 3.20 Temporal changes of solid moisture content and temperature and exit air humidity and temperature in a sample batch fluid bed dryer. Bed diameter 0.6 m, bed height 1.2 m, particle diameter 3 mm, particle density 1200 kg/m³, air temperature 150°C, and humidity 1 g/kg.

3.8.3 DEEP BED DRYING

In deep bed drying solid phase is stationary and remains in the dryer for a certain time while gas phase flows through it continuously (case 2a of Figure 3.1). Drying begins at the inlet end of gas and progresses through the entire bed. A typical desorption wave travels through the bed. The situation is shown schematically in Figure 3.21.

The above situation is described by the following set of equations:

$$\frac{dX}{d\tau} = -\frac{w_D a_V}{(1-\varepsilon)\rho_S} \quad (3.125)$$

$$\frac{dY}{dh} = \frac{S}{W_B} w_D a_V \quad (3.126)$$

$$\begin{aligned} \frac{dt_m}{d\tau} &= \frac{a_V}{(1-\varepsilon)\rho_S} \frac{1}{c_S + c_{AI}X} \\ &= [q - ((c_A - c_{AI})t_m + \Delta h_{v0})w_D] \end{aligned} \quad (3.127)$$

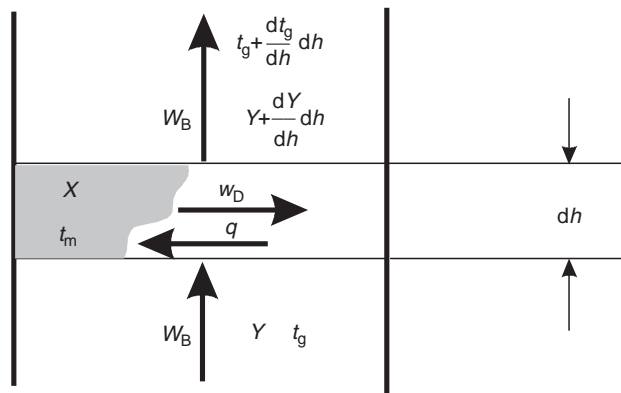


FIGURE 3.21 Schematic of batch drying in a deep layer.

$$\frac{dt_g}{dh} = -\frac{S}{W_B} a_V \frac{1}{c_B + c_A Y} [q + c_A(t_g - t_m)w_D] \quad (3.128)$$

The equations can be solved by finite difference discretization and a suitable numerical technique. Figure 3.22 presents the results of a simulation of drying cereal grains in a thick bed using Mathcad. Note how a desorption wave is formed, and also that the solid in deeper regions of the bed initially takes up moisture from the air humidified during its passage through the entry region.

Given a model together with its method of solution it is relatively easy to vary BCs, e.g., change air temperature in time or switch the gas flow from top to bottom intermittently, and observe the behavior of the system.

3.9 MODELS FOR SEMICONTINUOUS DRYERS

In some cases the dryers are operated in such a way that a batch of solids is loaded into the dryer and it progressively moves through the dryer. New batches are loaded at specified time intervals and at the same moment dry batches are removed at the other end. Therefore, the material is not moving continuously but by step increments. This is a typical situation in a tunnel dryer where trucks are loaded at one end of a tunnel and unloaded at the other, as shown in Figure 3.23.

To simplify the case one can take an LPM for each truck and a DPM for circulating air. As before, we will neglect accumulation in the gas phase but of course consider it in the solid phase. The resulting set of equations is

$$\frac{dX^i}{d\tau} = -\frac{w_{Di} A_i}{m_{Si}} \quad (3.129)$$

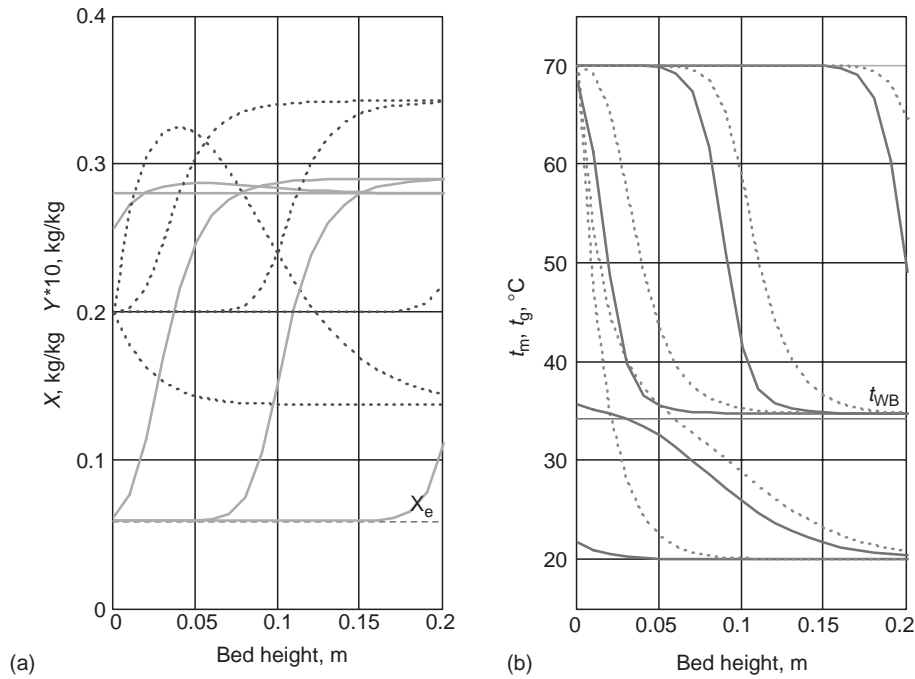


FIGURE 3.22 Simulation of deep bed drying of cereal grains: (a) moisture content profiles (solid lines) and gas humidity profiles (broken lines); (b) material temperature (solid lines) and air temperature (broken lines). Initial solid temperature 20°C and gas inlet temperature 70°C. Profiles are calculated at 0.33, 1.67, 3.33, 6.67, and 11.67 min of elapsed time. X_e is equilibrium moisture content and t_{WB} is wet-bulb temperature.

$$\frac{dY}{dl} = \frac{S}{W_B} w_D a_V \quad (3.130)$$

$$\frac{dt_m^i}{d\tau} = \frac{A_i}{m_{Si}} \frac{1}{c_S + c_{AI} X^i} \times [q_i - ((c_A - c_{AI}) t_m^i + \Delta h_{v0}) w_{Di}] \quad (3.131)$$

$$\frac{dt_g}{dl} = -\frac{S}{W_B} a_V \frac{1}{c_B + c_A Y} [q + c_A (t_g - t_m) w_D] \quad (3.132)$$

where i is the number of a current truck. Additionally, a balance equation for mixing of airstreams at fresh air entry point is required. The semi-steady-state solution is when a new cycle of temporal change of X^i and t_m^i will be identical to the old cycle. In order to converge to a semi-steady state the initial profiles of X^i and t_m^i must be assumed. Usually a linear distribution between the initial and the final values is enough. The profiles are adjusted with each iteration until a cyclic solution is found.

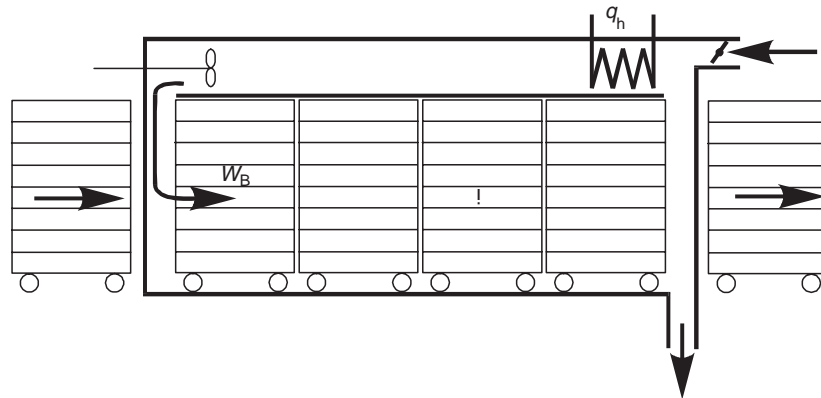


FIGURE 3.23 Schematic of a semicontinuous tunnel dryer.

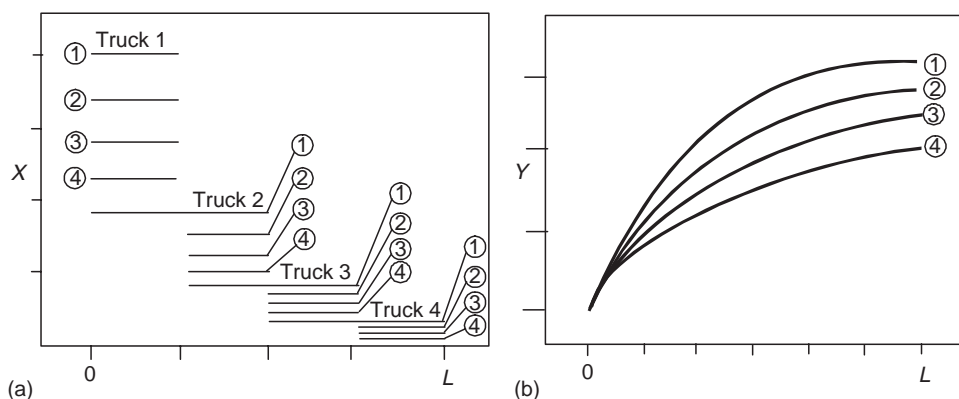


FIGURE 3.24 Schematic of the model solution for semicontinuous tunnel dryer for cocurrent flow of air vs. truck direction—mass transfer only: (a) moisture content in trucks at specified equal time intervals; (b) humidity profiles at specified time intervals. 1, 2, 3, 4—elapsed times.

The solution of this system of equations is schematically shown in Figure 3.24 for semi-steady-state operation and four trucks in the dryer. In each truck moisture content drops in time until the load–unload time interval. Then the truck is moved one position forward so the last moisture content for this truck at former position becomes its initial moisture content at the new position. A practical application of this model for drying of grapes is presented by Caceres-Huambo and Menegalli (2002).

3.10 SHORTCUT METHODS FOR DRYER CALCULATION

When no data on sorptional properties, water diffusivity, shrinkage, etc., are available, dryer design can only be approximate, nevertheless useful, as a first approach. We will identify here two such situations.

3.10.1 DRYING RATE FROM PREDICTED KINETICS

3.10.1.1 Free Moisture

This case exists when drying of the product entirely takes place in the constant drying rate period. It is almost always possible when the solid contains unbound moisture. Textiles, minerals, and inorganic chemicals are examples of such solids.

Let us investigate a continuous dryer calculation. In this case solid temperature will reach, depending on a number of transfer units in the dryer, a value between AST and WBT, which can easily be calculated from Equation 3.49 and Equation 3.50. Now mass and energy balances can be closed over the whole dryer and exit parameters of air and material obtained. Having

these, the averaged solid and gas temperatures and moisture contents in the dryer can be calculated. Finally the drying rate can be calculated from Equation 3.27, which in turn allows one to calculate solid area in the dryer. Various aspect ratios of the dryer chamber can be designed; one should use judgment to calculate dryer cross-section in such a way that air velocity will not cause solid entrainment, etc.

3.10.1.2 Bound Moisture

In this case we can predict drying rate by assuming that it is linear, and at $X = X^*$ drying rate is zero, whereas at $X = X_c$ drying rate is w_{DI} . The equation of drying rate then becomes

$$w_D = w_{DI} \frac{X - X^*}{X_c - X^*} = w_{DI} \Phi \quad (3.133)$$

This equation can be used for calculation of drying time in batch drying. Substituting this equation into Equation 3.112 and integration from the initial X_0 to final moisture content X_f , the drying time is obtained

$$\tau = \frac{m_s}{Aw_{DI}} (X_c - X^*) \ln \frac{X_0 - X^*}{X_f - X^*} \quad (3.134)$$

Similarly, Equation 3.133 can be used in a model of a continuous dryer.

3.10.2 DRYING RATE FROM EXPERIMENTAL KINETICS

Another simple case is when the drying curve has been obtained experimentally. We will discuss both batch and continuous drying.

TABLE 3.3
External RTD Function for Selected Models of Flow

Model of Flow	<i>E</i> Function	
Plug flow	$E(\tau) = \delta(\tau - \tau_r)$	(3.138)
Perfectly mixed flow	$E(\tau) = \frac{1}{\tau_r} e^{-\tau/\tau_r}$	(3.139)
Plug flow with axial dispersion	$E(\tau) = \frac{1}{\sigma\sqrt{2\pi}} \exp\left(-\frac{(\tau - \tau_r)^2}{2\sigma^2}\right)$	(3.140)
	where for $Pe \geq 10$, $\frac{\sigma^2}{\tau_r^2} = \frac{2}{Pe}$	
<i>n</i> -Perfectly mixed uniform beds	$E(\tau) = \frac{n}{\tau_r} \frac{(n(\tau/\tau_r))^{n-1}}{(n-1)!} \exp\left(-n\frac{\tau}{\tau_r}\right)$	(3.141)

3.10.2.1 Batch Drying

We may assume that if the solid size and drying conditions in the industrial dryer are the same, the drying time will also be the same as obtained experimentally. Other simple scaling rules apply, e.g., if a batch fluid bed thickness is double of the experimental one, the drying time will also double.

3.10.2.2 Continuous Drying

Here the experimental drying kinetics can only be used if material flow in the dryer is of plug type. In other words, it is as if the dryer served as a transporter of a batch container where drying is identical to that in the experiment. However, when a certain degree of mixing of the solid phase occurs, particles of the solid phase exiting the dryer will have various residence times and will therefore differ in moisture content. In this case we can only talk of average final moisture content. To calculate this value we will use methods of residence time distribution (RTD) analysis. If the empirical drying kinetics curve can be represented by the following relationship:

$$X = f(X_0, \tau) \quad (3.135)$$

and mean residence time by

$$\tau_r = \frac{m_S}{W_S} \quad (3.136)$$

the average exit solid moisture content can be calculated using the external RTD function *E* as

$$\bar{X} = \int_0^{\infty} E(\tau) X(X_0, \tau) d\tau \quad (3.137)$$

Formulas for *E* function are presented in Table 3.3 for the most common flow models.

Figure 3.25 is an exemplary comparison of a batch and real drying curves. As can be seen, drying time in real flow conditions is approximately 50% longer here.

3.11 SOFTWARE TOOLS FOR DRYER CALCULATIONS

Menshutina and Kudra (2001) present 17 commercial and semicommercial programs for drying calculations that they were able to identify on the market. Only a few of them perform process calculations of dryers including dryer dimensioning, usually for fluid bed dryers. Typically a program for dryer calculations performs balancing of heat and mass and, if dimensioning is possible, the program requires empirical coefficients, which the user has to supply. Similarly, the drying process is designed in commercial process simulators used in chemical and process engineering. A program that does all calculations presented in this chapter does not exist. However, with present-day computer technology, construction of such software is possible; dryPAK (Silva and Correa, 1998; Pakowski, 1999) is a program that evolves in this direction. The main concept in dryPAK is that all models share the same database of humid gas, moist material properties, methods for calculation of drying rate, etc. The results are also visualized in the same way. Figure 3.5, Figure 3.7, Figure 3.14, and Figure 3.16 were in fact produced with dryPAK.

General-purpose mathematical software can greatly simplify solving new models of not-too-complex structure. Calculations shown in Figure 3.9, Figure 3.12, Figure 3.18, Figure 3.20, and Figure 3.22 were produced with Mathcad. Mathcad or MATLAB can

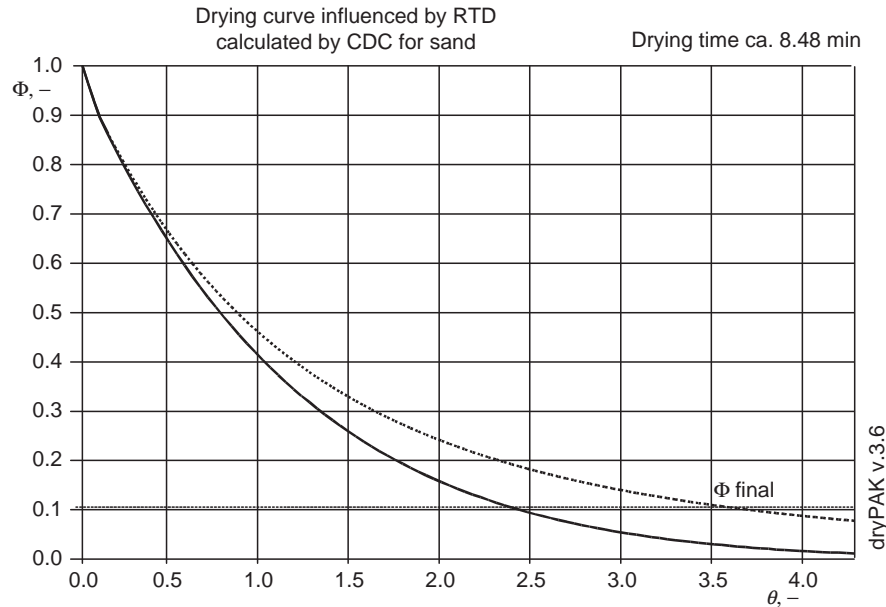


FIGURE 3.25 Experimental (solid line) and actual (broken line) drying kinetic curves for three tanks in series model of flow. θ is the ratio of the actual to mean residence times, Φ is dimensionless moisture content.

significantly reduce the effort involved with numerical solutions of equation systems as they contain a multitude of solvers for both algebraic and differential equations. Problems that would require several days of work can now be solved within hours. To let the reader get acquainted with this tool several Mathcad files containing selected solutions of problems presented in this section will be made available at <http://chemeng.p.lodz.pl/books/HID/>. Both MATLAB and Mathcad offer associated tools for visual modeling of dynamic systems (Simulink and VisiSim, respectively) that make simulation of batch system even easier.

3.12 CONCLUSION

In this chapter we have illustrated how dryer calculations can be made by constructing a model of a dryer and solving it using appropriate numerical methods. Using general-purpose mathematical software solving models is a task that can be handled by any engineer. The results can be obtained in a short time and provide a sound basis for more detailed dryer calculations. For more advanced and specialized dryer design dedicated software should be sought. However, the question of how to obtain the necessary property data of dried materials remains. This question is as important now as it was before since very little has been done in the area of materials databases. The data are spread over the literature and, in the case of unsuccessful search, an experimental determination of the missing data is necessary.

NOMENCLATURE

A	interfacial area of phase contact, m^2
a, b, c	constants of GAB equation
a_v	characteristic interfacial area per unit volume of dryer, $1/\text{m}$
c	specific heat, $\text{kJ}/(\text{kg K})$
D	diffusivity, m^2/s
E	axial dispersion coefficient, m^2/s
E	external RTD function
f	ratio of drying rates in CDC equation
G	shear modulus, Pa
h	specific enthalpy per unit mass of species, kJ/kg
Δh_s	latent heat of sorption, kJ/kg
Δh_v	latent heat of vaporization, kJ/kg
i	specific enthalpy per dry basis, kJ/kg
k	permeability, m^2
k_Y	mass transfer coefficient, $\text{kg}/(\text{m}^2 \text{ s})$
L	total length, m
l	running length, m
M	molar mass, kg/kmol
m	mass holdup, kg
p	vapor pressure, Pa
P_0	total pressure, Pa
q	heat flux, kW/m^2
R	maximum radius, m
R	universal gas constant, $\text{kJ}/(\text{kmol K})$
r	actual radius, m
S	cross-sectional area normal to flow direction, m^2

s	shrinkage coefficient
t	temperature, °C
T	absolute temperature, K
W	mass flowrate, kg/s
w_D	drying rate, kg/(m ² s)
X	moisture content per dry basis, kg/kg
x	coordinate in Cartesian system, m
Y	absolute humidity per dry basis, kg/kg
V	total volume, m ³
α	heat transfer coefficient, kW/(m ² K)
δ	Dirac delta function
ε	voidage
Φ	dimensionless moisture content = $(X - X^*) / (X_c - X^*)$
ϕ	correcting coefficient in Equation 3.27
φ	relative humidity
λ	thermal conductivity, kW/(m K)
μ	viscosity, Pa s
ν	Poisson's ratio
ρ	density, kg/m ³
τ	time, s

SUBSCRIPTS AND SUPERSSCRIPTS

A	moisture
AS	adiabatic saturation
B	dry gas
c	critical (for moisture content)
g	humid gas
i	at interface
m	wet solid
m	mean value
s	at saturation
S	dry solid
WB	wet bulb
v	vapor phase
*	in equilibrium
—	space averaged

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4 Transport Properties in the Drying of Solids

Dimitris Marinos-Kouris and Z.B. Maroulis

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4.1 INTRODUCTION

Drying is a complicated process involving simultaneous heat, mass, and momentum transfer phenomena, and effective models are necessary for process design, optimization, energy integration, and control. The development of mathematical models to describe drying processes has been a topic of many research studies for several decades. Undoubtedly, the observed progress has limited empiricism to a large extent. However, the design of dryers is still a mixture of science and practical experience. Thus the prediction of Luikov that by 1985 “would obviate the need for empiricism in selecting optimum drying conditions,” represented an optimistic perspective, which, however, shows that the efforts must be increased [1]. Presently, more and more sophisticated drying models are becoming available, but a major question that still remains is the measurement or determination of the parameters used in the models. The measurement or estimation of the necessary parameters should be feasible and practical for general applicability of a drying model.

In the early 1970s, Nonhebel and Moss stated that “the choice of drying plant, or design of special plant to meet unprecedented conditions” would require use of 34 parameters [2]. Regardless of the truth of such a statement, that is, of the actual number of parameters necessary for the design of a dryer, there is an obvious need for a large amount of data. Nowadays, the completeness and accuracy of such data reflect to a

large extent our ability to perform effective process design. It should be noted that in spite of the intense activities in the drying literature (*Drying Technology Journal*, *Advances in Drying*, *Drying*, *International Drying Symposium*, etc.), the problem of property data still remains an important one because such data are widely scattered and not systematically evaluated. Moreover, whereas the need “for accurate design data is increasing, the rate of accumulation of new data is not increasing fast enough” [3]. The lack of data is expected to continue and, as noted by Keey, “it is probably unrealistic to expect complete hygrothermal data for materials of commercial interest” [4].

Out of the full set of thermophysical properties necessary for the analysis of drying of a material, this chapter examines only those that are critical. As such, we consider the thermodynamic and transport properties, which are usually incorporated in a drying model as model parameters, and which are:

- Effective moisture diffusivity
- Effective thermal conductivity
- Air boundary heat and mass transfer coefficients
- Drying constant
- Equilibrium material moisture content

Effective thermal conductivity and effective moisture diffusivity are related to internal heat and mass transfer, respectively, while air boundary heat and mass transfer coefficients are related to external heat

and mass transfer, respectively. The above transport properties are usually coefficients in the corresponding flow rate and driving force relationship. The equilibrium material moisture content, on the other hand, is usually related to the mass transfer driving force.

The above transport properties in conjunction with a transport phenomena mechanistic model can adequately describe the drying kinetics, but sometimes an additional property, the drying constant, is also used. The drying constant is essentially a combination of the above transport properties and it must be used in conjunction with the so-called thin-layer model.

Effective moisture diffusivity and effective thermal conductivity are in general functions of material moisture content and temperature, as well as of the material structure. Air boundary coefficients are functions of the conditions of the drying air, that is humidity, temperature, and velocity, as well as system geometry. Equilibrium moisture content of a given material is a function of air humidity and temperature. The drying constant is a function of material moisture content, temperature, and thickness, as well as air humidity, temperature, and velocity.

The required accuracy of the above properties depends on the controlling resistance to heat and mass transfer. If, for example, drying is controlled by the internal moisture diffusion, then the effective moisture diffusivity must be known with high accuracy. This situation is valid when large particles are drying with air of high velocity. Drying of small particles with low velocity of air is controlled by the external mass transfer, and the corresponding coefficient should be known with high accuracy. But there are situations in which heat transfer is the controlling resistance. This happens, for example, in drying of solids with high porosity, in which high mass and low heat transfer rates are obtained.

The purpose of this chapter is to examine the above properties related to drying processes, particularly drying kinetics. Most of the following topics are discussed for each property:

- Definition
- Methods of experimental measurement
- Data compilation
- Effect of various factors
- Theoretical estimation

The statement of Poersch (quoted in Ref. [4]) that it is possible for someone to dry a product based on experience and without theoretical knowledge but not the reverse is worth repeating here. To this we may add the comment that it is impossible to efficiently

dry a product without complete and precise thermo-physical data.

4.2 MOISTURE DIFFUSIVITY

4.2.1 DEFINITION

Diffusion in solids during drying is a complex process that may involve molecular diffusion, capillary flow, Knudsen flow, hydrodynamic flow, or surface diffusion. If we combine all these phenomena into one, the effective diffusivity can be defined from Fick's second law

$$\partial X / \partial t = D \nabla^2 X \quad (4.1)$$

where D (m²/s) is the effective diffusivity, X (kg/kg db) is the material moisture content, and t (s) is the time.

The moisture transfer in heterogeneous media can be conveniently analyzed by using Fick's law for homogeneous materials, in which the heterogeneity of the material is accounted for by the use of an effective diffusivity.

Equation 4.1 shows the time change of the material moisture distribution, that is, it describes the movement of moisture within the solid. The previous equation can be used for design purposes in cases in which the controlling mechanism of drying is the diffusion of moisture.

Pakowski and Mujumdar [5] describe the use of Equation 4.1 for the calculation of the drying rate, whereas Strumillo and Kudra [6] describe its use in calculating the drying time. Solutions of the Fickian equation for a variety of initial and boundary conditions are exhaustively described by Crank [7].

4.2.2 METHODS OF EXPERIMENTAL MEASUREMENT

There is no standard method for the experimental determination of diffusivity. The diffusivity in solids can be determined using the methods presented in Table 4.1. These methods have been developed primarily for polymeric materials [7–9]. Table 4.1 also includes the relevant entries in the “References” section for the application of the methods in food systems.

4.2.2.1 Sorption Kinetics

The sorption (adsorption or desorption) rate is measured with a sorption balance (spring or electrical) whereas the solid sample is kept in a controlled environment. Assuming negligible surface resistance to mass transfer, the method is based on Fick's diffusion equation.

TABLE 4.1
Methods for the Experimental Measurement
of Moisture Diffusivity

Method	Ref.
Sorption kinetics	8
Permeation methods	8
Concentration–distance curves	10–12
<i>Other methods</i>	
Radiotracer methods	8
Nuclear magnetic resonance (NMR)	8, 13, 14
Electron spin resonance (ESR)	8, 15
<i>Drying technique</i>	
Simplified methods	16
Regular regime method	17–19
Numerical solution—regression analysis	See Section 4.7

4.2.2.2 Permeation Method

The permeation method is a steady-state method applied to a film of material. According to this method, the permeation rate of a diffusant through a material of known thickness is measured under constant, well-defined, surface concentrations. The analysis is also based on Fick's diffusion equation.

4.2.2.3 Concentration–Distance Curves

The concentration–distance curves method is based on the measurement of the distribution of the diffusant concentration as a function of time. Light interference methods, as well as radiation adsorption or simply gravimetric methods, can be used for concentration measurements. Various sample geometries can be used, for example semiinfinite solid, two joint cylinders with the same or different material, and so on. The analysis is based on the solution of Fick's equation.

4.2.2.4 Other Methods

Modern methods for the measurement of moisture profiles lead to diffusivity measurement methods. Such methods discussed in the literature are radiotracer methods, nuclear magnetic resonance (NMR), electron spin resonance (ESR), and the like.

4.2.2.5 Drying Methods

The simplified, regular regime, and regression analysis methods are particularly relevant for drying processes. In them, the samples are placed in a dryer and moisture diffusivity is estimated from drying data. All the drying methods are based on Fick's

equation of diffusion, and they differ with respect to the solution methodology. The following analysis is considered.

4.2.2.5.1 Simplified Methods

Fick's equation is solved analytically for certain sample geometries under the following assumptions:

Surface mass transfer coefficient is high enough so that the material moisture content at the surface is in equilibrium with the air drying conditions. Air drying conditions are constant. Moisture diffusivity is constant, independent of material moisture content and temperature.

The analytical solution for slab, spherical, or cylindrical samples is used in the analysis. Several alternatives exist concerning the methodology of estimation of diffusivity using the above equations. They are discussed in the COST 90bis project of European Economic Community (EEC) [16]. These alternatives differ essentially on the variable on which a regression analysis is applied.

4.2.2.5.2 Regular Regime Method

The regular regime method is based on the experimental measurement of the *regular regime curve*, which is the drying curve when it becomes independent of the initial concentration profile. Using this method, the concentration-dependent diffusivity can be calculated from one experiment.

4.2.2.5.3 Numerical Solution—Regression Analysis Method

The regression analysis method can be considered as a generalization of the other two types of methods. It can estimate simultaneously some additional transport properties; it is analyzed in detail in [Section 4.7](#).

4.2.3 DATA COMPILATION

Effective diffusivities, reported in the literature, have been usually estimated from drying or sorption rate data. Experimental data are scarce because of the effect of the experimental method, the method of analysis, the variations in composition and structure of the examined materials, and so on. Data of effective diffusion coefficients are available for inorganic materials [20], polymers [8], and foods [21,22].

[Table 4.2](#) gives some literature values of the effective diffusivity of moisture in various materials. A number of data from the above-mentioned bibliographic entries are also included in [Table 4.2](#). New data up to 1992 are also incorporated. Foods are the

TABLE 4.2
Effective Moisture Diffusivity in Some Materials

Classification ^a	Material	Water Content (kg/kg db)	Temperature (°C)	Diffusivity (m ² /s)	Ref.
<i>Food</i>					
1	Alfalfa stems	<3.70	26	2.6E-12–2.6E-09	23
2	Apple	0.12	60	6.5E-12–1.2E-10	24
		0.15–7.00	30–76	1.2E-10–2.6E-10	25
3	Avocado		31–56	1.1E-10–3.3E-10	26
4	Beet		65	1.5E-09	26
5	Biscuit	0.10–0.65	20–100	9.4E-10–9.7E-08	27
6	Bread	0.10–0.70	20–100	2.5E-09–5.5E-07	27
7	Carrot	0.03–11.6	42–80	9.0E-10–3.3E-09	28
8	Corn	0.05–0.23	40	1.0E-12–1.0E-10	29
		0.19–0.27	36–62	7.2E-11–3.3E-10	30
9	Fish muscle	0.05–0.30	30	8.1E-11–3.4E-10	31
10	Garlic	0.20–1.60	22–58	1.1E-11–2.0E-10	32
11	Milk foam	0.20	40	1.1E-09	33
	Milk skim	0.25–0.80	30–70	1.5E-11–2.5E-10	34
12	Muffin	0.10–0.65	20–100	8.4E-10–1.5E-07	27
13	Onion	0.05–18.7	47–81	7.0E-10–4.9E-09	35
14	Pasta, semolina	0.01–0.25	40–125	3.0E-13–1.5E-10	36
	Pasta, corn based	0.10–0.40	40–80	5.0E-11–1.3E-10	37
	Pasta, durum wheat	0.16–0.35	50–90	2.5E-12–5.6E-11	38
15	Pepper, green	0.04–16.2	47–81	5.0E-10–9.2E-09	35
16	Pepperoni	0.19	12	4.7E-11–5.7E-11	39
17	Potato	0.60	54	2.6E-10	40
		<4.00	65	4.0E-10	41
		0.15–3.50	65	1.7E-09	42
		0.01–7.20	39–82	5.0E-11–2.7E-09	43
18	Rice	0.18–0.36	60	1.3E-11–2.3E-11	44
		0.28–0.64	40–56	1.0E-11–6.9E-11	45
19	Soybeans, defatted	0.05	30	2.0E-12–5.4E-12	46
20	Starch, gel	0.10–0.30	25	1.0E-12–2.3E-11	47
		0.20–3.00	30–50	1.0E-10–1.2E-09	48
		0.75	25–140	1.0E-10–1.5E-09	49
	Starch granular	0.10–0.50	25–140	5.0E-10–3.0E-09	49
21	Sugar beet	2.50–3.60	40–80	4.0E-10–1.3E-09	50, 51
22	Tapioca root	0.16–1.95	97	9.0E-10	52, 53
23	Turkey	0.04	22	8.0E-15	54
24	Wheat	0.12–0.30	21–80	6.9E-12–2.8E-10	55
		0.13–0.20	20	3.3E-10–3.7E-09	56
<i>Other materials</i>					
1	Asbestos cement	0.10–0.60	20	2.0E-09–5.0E-09	20
2	Avicel (FMC Corp.)		37	5.0E-09–5.0E-08	57
3	Brick powder	0.08–0.16	60	2.5E-08–2.5E-06	58
4	Carbon, activated		25	1.6E-05	59
5	Cellulose acetate	0.05–0.12	25	2.0E-12–3.2E-12	60
6	Clay brick	0.20	25	1.3E-08–1.4E-08	61
7	Concrete	0.10–0.40	20	5.0E-10–1.2E-08	20
	Concrete, pumice	0.20	25	1.8E-08	61
8	Diatomite	0.05–0.50	20	3.0E-09–5.0E-09	20
9	Glass wool	0.10–1.80	20	2.0E-09–1.5E-08	20
	Glass spheres, 10 µm	0.01–0.22	60	1.84E-8 ± 0.94E-8	16
10	Hyde clay	0.10–0.40		5.0E-09–1.0E-08	62
11	Kaolin clay	<0.50	45	1.5E-08–1.5E-07	20
12	Model system		68	3.1E-09	63

continued

TABLE 4.2 (continued)
Effective Moisture Diffusivity in Some Materials

Classification ^a	Material	Water Content (kg/kg db)	Temperature (°C)	Diffusivity (m ² /s)	Ref.
13	Peat	0.30–2.50	45	4.0E-08–5.0E-08	20
14	Sand	<0.15	45	8.0E-08–1.5E-07	20
	Sand, sea	0.07–0.13	60	2.5E-08–2.5E-06	58
	Sand	0.05–0.10		1.0E-07–1.0E-06	64
15	Silica alumina	0.59–1.18	60	2.5E-08–2.5E-06	58
16	Silica gel		25	3.0E-06–5.6E-06	59
17	Tobacco leaf		30–50	3.2E-11–8.1E-11	65
18	Wood, soft		40–90	5.0E-10–2.5E-09	66
	Wood, yellow poplar	1.00	100–150	1.0E-08–2.5E-08	67

^aClassification number for each material used in Figure 4.1.

most investigated materials in the literature, and they are presented separately. Table 4.2 was prepared for the needs of this chapter, that is, to show the range of variation of diffusivity for various materials and not to present some experimental values. That is why most of the data are presented as ranges.

The data of Table 4.2 are further displayed in Figure 4.1 through Figure 4.4. The moisture diffusivity is plotted versus the number of material for food and other materials in Figure 4.1. Diffusivities in foods have values in the range 10^{-13} to 10^{-6} m²/s, and most of them (82%) are accumulated in the region 10^{-11} to 10^{-8} . Diffusivities of other materials have values in the range 10^{-12} to 10^{-5} , whereas most of them (58%) are accumulated in the region 10^{-9} to 10^{-7} . These results are also clarified in the histograms of Figure 4.2. Diffusivities in foods are less than those in other materials. This is because of the complicated biopolymer structure of food and, probably, the stronger binding of water in them.

The influence of material moisture content and temperature from the statistical point of view is shown in Figure 4.3 and Figure 4.4. Figure 4.3 shows the diffusivities versus the material moisture content for all the materials. The positive effect of material moisture content on diffusivity is evident. The same trend is noted in Figure 4.4 with regard to the temperature. It should be noted that the observed trends in the previous figures are the result of examining different materials at various temperatures and moistures and from various sources. The influence of material moisture content and temperature for each material is discussed in the next section.

In general, comparison among diffusivities reported in the literature is difficult because of the different methods of estimation and the variation of composition, especially for foods. However, on the

basis of Figure 4.3 and Figure 4.4, it is concluded that the differences in diffusivity among materials are less than that between temperature or material moisture content of the same material. Diffusivities of other solutes in various materials are also presented in the literature (e.g., see Ref. [68]).

4.2.4 FACTORS AFFECTING DIFFUSIVITY

Moisture diffusivity depends strongly on temperature and, often, very strongly on the moisture content, but there are few reliable figures. In porous materials the void fraction affects diffusivity significantly, and the pore structure and distribution do so even more.

The temperature dependence of the diffusivity can generally be described by the Arrhenius equation, which takes the form

$$D = D_0 \exp(-E/RT) \quad (4.2)$$

where D_0 (m²/s) is the Arrhenius factor, E (kJ/kmol) is the activation energy for diffusion, R (kJ/(kmol K)) the gas constant, and T (K) the temperature.

The moisture content dependence of the diffusivity can be introduced in the Arrhenius equation by considering either the activation energy or the Arrhenius factor as an empirical function of moisture. Both modifications can be considered simultaneously. Other empirical equations not based on the Arrhenius equation can be used.

The moisture diffusivity is an increasing function of the temperature and moisture of the material. Yet, in certain categories of polymers, deviation from this kind of behavior has been observed. For instance, for several of the less hydrophilic polymers (e.g., polymethacrylates and polycrylates) the moisture diffusivity decreases with increasing water content. On the

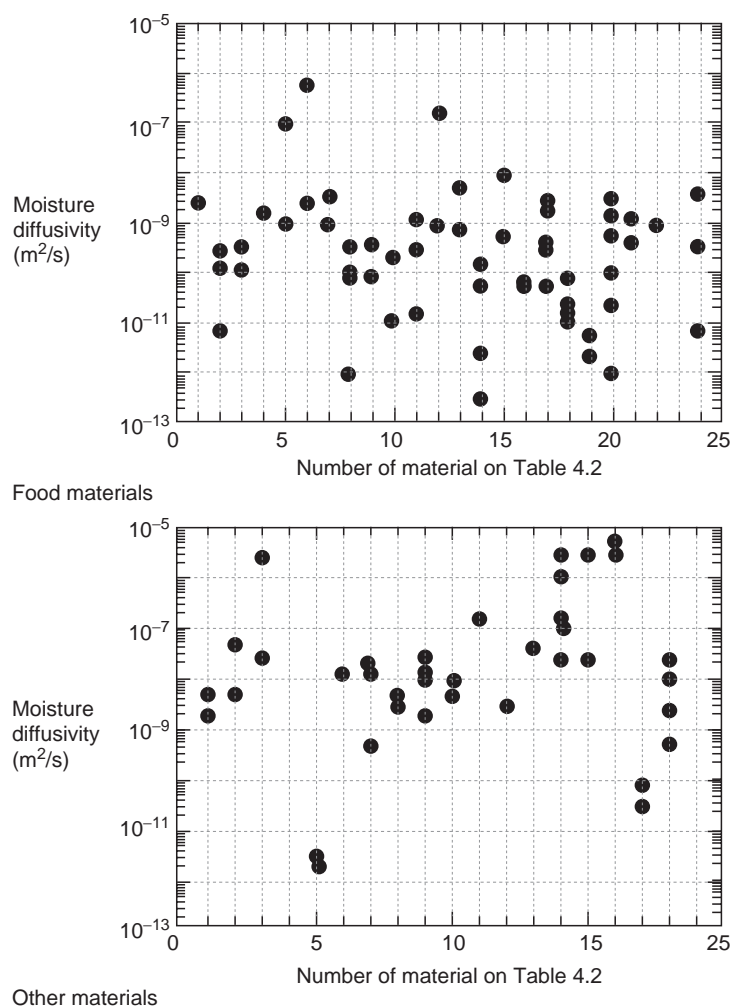


FIGURE 4.1 Moisture diffusivity in various materials (data from Table 4.2).

other hand, the moisture diffusivity appears to be independent of the concentration—and hence constant—for some hydrophobic polyolefins.

Table 4.3 gives some relationships that describe simultaneous dependence of the diffusivity upon temperature and moisture. Some rearrangement of the equations proposed has been done in order to present them in a uniform format. Table 4.4 lists parameter values for typical equations of Table 4.3.

Equation T3.1 through Equation T3.4 in Table 4.3 suggest that the material moisture content can be taken into account by considering the preexponential factor of the Arrhenius equation as a function of material moisture content. Polynomial functions of first order can be considered (Equation T3.1), as well as of higher order (Equation T3.2 or Equation T3.3). The exponential function can also be used (Equation T3.4).

Equation T3.5 and Equation T3.6 in Table 4.3 are obtained by considering the activation energy for

diffusion as a function of material moisture content. Equation T3.7 through Equation T3.10 are not based on the Arrhenius form. They are empirical and they use complicated functions concerning the discrimination of the moisture and temperature effects (except, of course, Equation T3.7). Equation T3.11 is more sophisticated as it considers different diffusivities of bound and free water and introduces the functional dependence of material moisture content on the binding energy of desorption. Equation T3.12 introduces the effect of porosity on moisture diffusivity.

With regard to the number of parameters involved (a significant measure concerning the regression analysis), it is concluded that at least three parameters are needed (Equation T3.1, Equation T3.5, and Equation T3.7).

Equation T3.5 and Equation T3.7 in Table 4.3 were applied to potato and clay brick, respectively, and the results are presented in Figure 4.5. Both materials exhibit typical behavior. Diffusivity at low

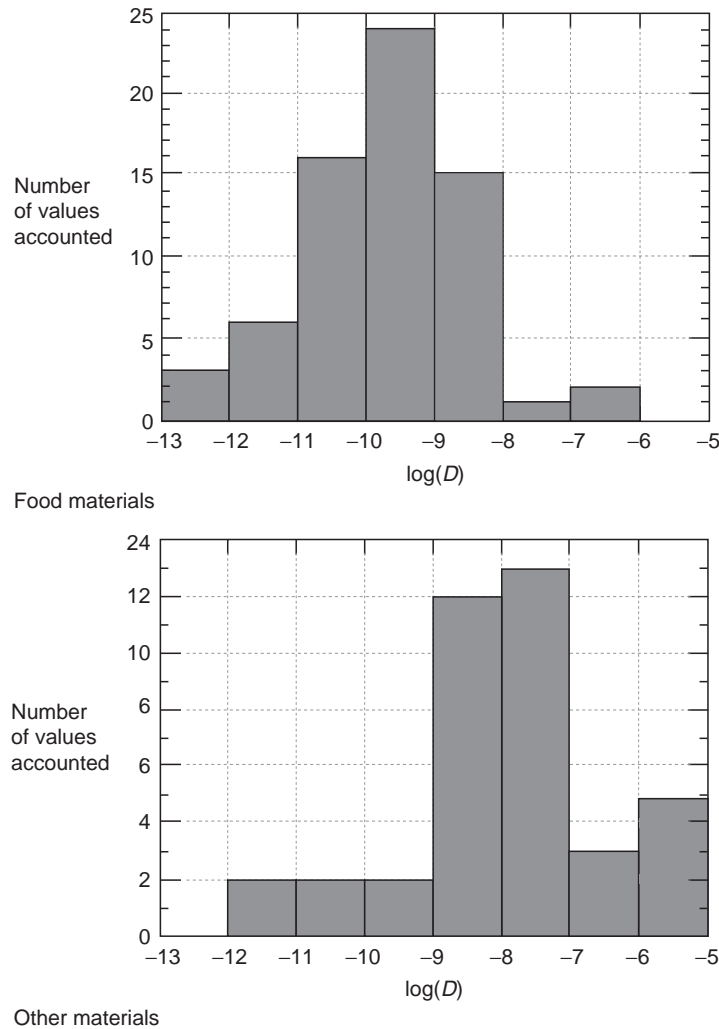


FIGURE 4.2 Histograms of diffusivities in various materials (data from Table 4.2).

moisture content shows a steep descent when the moisture content decreases.

The equations listed in Table 4.3 resulted from fitting to experimental data. The reason for the success of this procedure is the apparent simple dependence of diffusivity upon the material moisture content and temperature, which, as stated above, can be described even by three parameters only. The equations of Table 4.3 have been chosen by the respective researchers as the most appropriate for the material listed.

A single relation for the dependence of diffusivity upon the material moisture content and temperature general enough so as to apply to all the materials would be especially useful. It is expected that such a relation will be proposed soon.

The effect of pore structure and distribution on moisture diffusion can be examined by considering the material as a two-(or multi-) phase (dry material, water, air in voids, etc.) system and by considering

some structural models to express the system geometry. Although a lot of work has been done in the analogous case of thermal conductivity, little attention has been given to the case of moisture diffusivity, and even less experimental validation of the structural models has been obtained. The similarity, however, of the relevant transport phenomena (i.e., heat and mass transfer) permits, under certain restrictions, the use of conclusions derived from one area in the other. Thus, the literature correlations for the estimation of the effective diffusion coefficient, in many cases, had been initially developed for the thermal conductivity in porous media [79].

4.2.5 THEORETICAL ESTIMATION

The prediction of the diffusion coefficients of gases from basic thermophysical and molecular properties is possible with great accuracy using the Chapman–Enskog

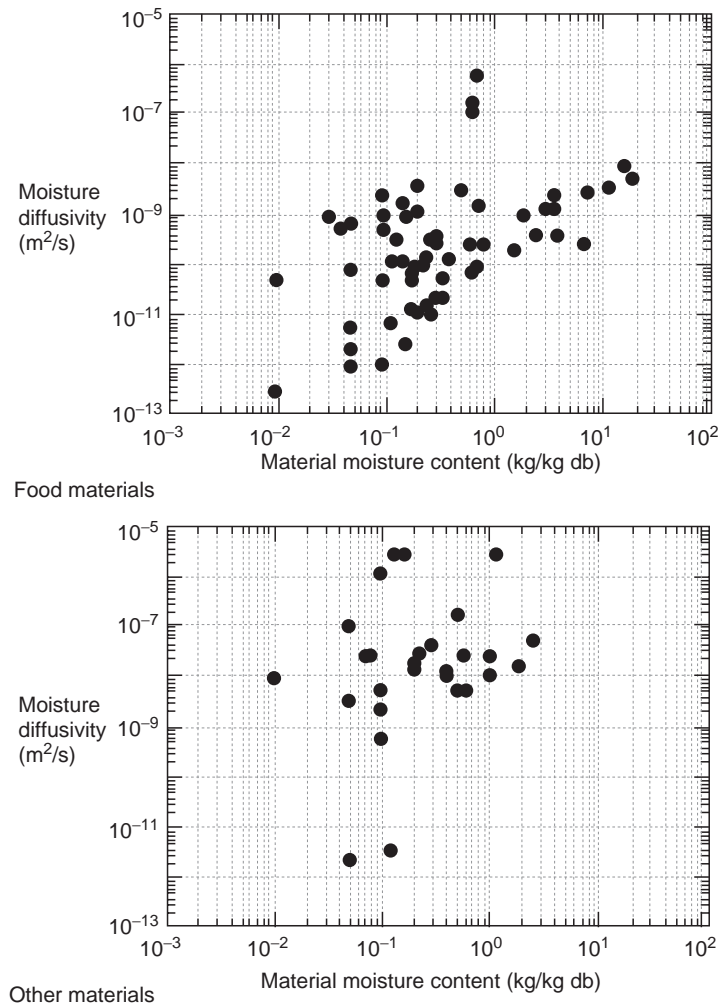


FIGURE 4.3 Moisture diffusivity versus material moisture content (data from Table 4.2).

kinetic theory. Diffusivities in liquids, on the other hand, in spite of the absence of a rigorous theory, can be estimated within an order of magnitude from the well-known equations of Stokes and Einstein (for large spherical molecules) and Wilke (for dilute solutions).

Diffusion of gases, vapors, and liquids in solids, however, is a more complex process than the diffusion in fluids because of the heterogeneous structure of the solid and its interactions with the diffusing components. As a result, it has not yet been possible to develop an effective theory for the diffusion in solids. Usually, diffusion in solids is handled by the researchers in a manner analogous to heat conduction. In the following paragraphs typical methods are described for the development of semiempirical correlations for diffusivity.

For the estimation of the diffusion coefficient in isotropic macroporous media, the relation

$$D = (\delta \varepsilon / \tau^2) D_A \quad (4.3)$$

has been proposed [79]. In this equation, ε is the porosity, τ is the tortuosity, δ is the constrictivity, and D_A is the vapor diffusivity in air in the absence of porous media. In spite of its simplicity, Equation 4.3 will not attain practical utility unless it is validated with additional pore space models, its parameters (ε , τ , δ) determined for a large number of systems, and the effect of the solid's moisture properly accounted for.

An equation has been derived relating the effective diffusivity of porous foodstuffs to various physical properties such as molecular weight, bulk density, vapor space permeability, water activity as a function of material moisture content, water vapor pressure, thermal conductivity, heat of sorption, and temperature [80]. A predictive model has been proposed to obtain effective diffusivities in cellular foods. The

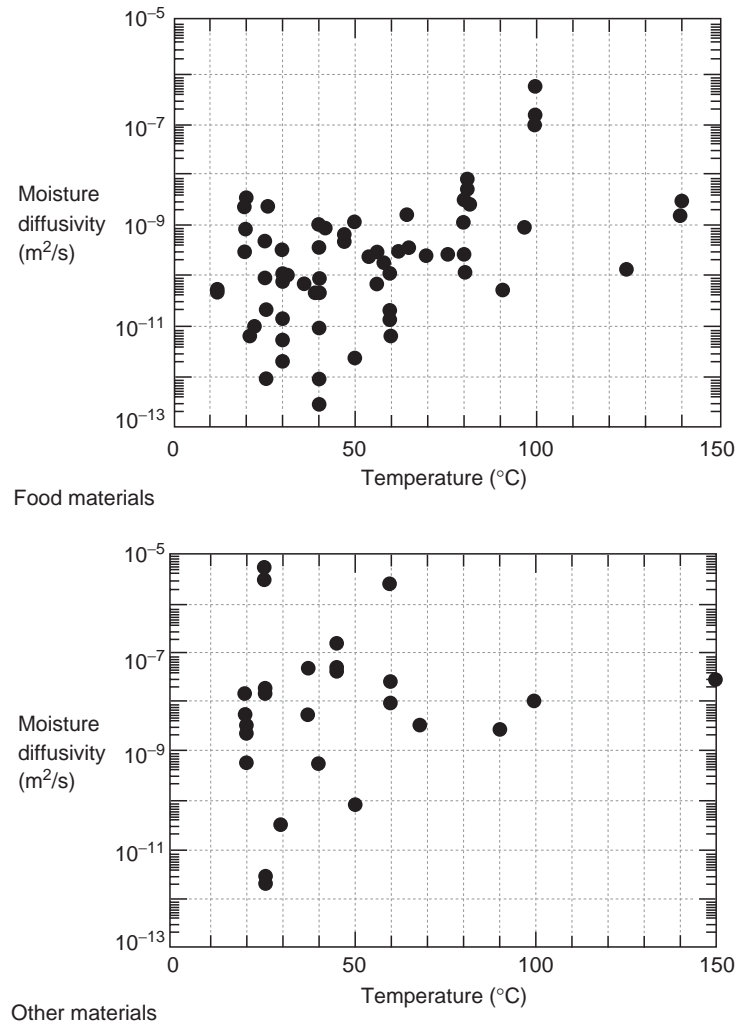


FIGURE 4.4 Moisture diffusivity versus material temperature (data from Table 4.2).

method requires data for composition, binary molecular diffusivities, densities, membrane and cell wall permeabilities, molecular weights, and water viscosity and molar volume [81]. The effect of moisture upon the effective diffusivity is taken into account via the binding energy of sorption in an equation suggested in Ref. [77].

4.3 THERMAL CONDUCTIVITY

4.3.1 DEFINITION

The *thermal conductivity* of a material is a measure of its ability to conduct heat. It can be defined using Fourier's law for homogeneous materials:

$$\partial T / \partial t = (k / \rho c_p) \nabla^2 T \quad (4.4)$$

where k is the thermal conductivity (kW/(m K)), ρ is the density (kg/m³), c_p is the specific heat of the

material (kJ/(kg K)), T is the temperature (K), and t is the time (s). The quantity $(k / \rho c_p)$ is the thermal diffusivity. For heterogeneous materials, the effective thermal conductivity is used in conjunction with Fourier's law.

Equation 4.4 is used in cases in which heat transfer during drying takes place through conduction (internally controlled drying). This, for example, is the situation when drying large particles, relatively immobile, that are immersed in the heat transfer medium.

As far as heat and mass transfer is concerned, the drying process is internally controlled whenever the respective Biot number (Bi_H , Bi_M) is greater than 1 [5].

4.3.2 METHODS OF EXPERIMENTAL MEASUREMENT

The effective thermal conductivity can be determined using the methods presented in Table 4.5, which includes the relevant references. Measurement techniques for thermal conductivity can be grouped into

TABLE 4.3
Effect of Material Moisture Content and Temperature on Diffusivity

Equation No.	Materials of Application	Equation	No. of Parameters	Ref.
T3.1	Apple, carrot, starch	$D(X,T) = a_0 \exp(a_1 X) \exp(-a_2/T)$	3	49, 69, 70
T3.2	Bread, biscuit, muffin	$D(X,T) = a_0 \exp\left(\sum_{i=1}^3 a_i X^i\right) \exp(-a_2/T)$	5	27
T3.3	Polyvinylalcohol	$D(X,T) = a_0 \exp\left(\sum_{i=1}^{10} a_i X^i\right) \exp(-a_2/T)$	12	71
T3.4	Vegetables	$D(X,T) = a_0 \exp(-a_1/X) \exp(-a_2/T)$	3	72
T3.5	Glucose, coffee extract, skim milk, apple, potato, animal feed	$D(X,T) = a_0 \exp[-a_1(1/T - 1/a_2)]$ $a_1 = a_{10} + a_{11} \exp(-a_{12}X)$	5	18
T3.6	Silica gel	$D(X,T) = a_0 \exp(-a_1/T) a_1 = a_{10} + a_{11}X$	3	73
T3.7	Clay brick, burned clay, pumice concrete	$D(X,T) = a_0 X^{a_1} T^{a_2}$	3	61
T3.8	Corn	$D(X,T) = a_0 \exp(a_1 X) \exp(-a_2/T) a_1 = a_{11}T + a_{10}$	4	30
T3.9	Rough rice	$D(X,T) = a_1 \exp(a_2 X) a_1 = a_{10} \exp(a_{11}T),$ $a_2 = a_{20} \exp(a_{21}T + a_{22}T^2)$	5	74, 75
T3.10	Wheat	$D(X,T) = a_0 + a_1 X + a_2 X^2 a_0 = a_{01} \exp(a_{02}T),$ $a_1 = a_{11} \exp(a_{12}T), a_2 = a_{21} \exp(a_{22}T)$	6	76
T3.11	Semolina, extruded	$D(X,T) = a_0 \exp(-a_2/T) \frac{a_2 \exp(-a_3/T)}{1 + a_2 \exp(-a_3/T)}$	4	77
T3.12	Porous starch	$D(X,T) = (a_0 + a_1 X^{a_2}) \exp(-a_3/T) a_0 = F(\varepsilon)$	>5	78

D , moisture diffusivity; X , material moisture content; T , temperature; a_i , constants; ε , porosity.

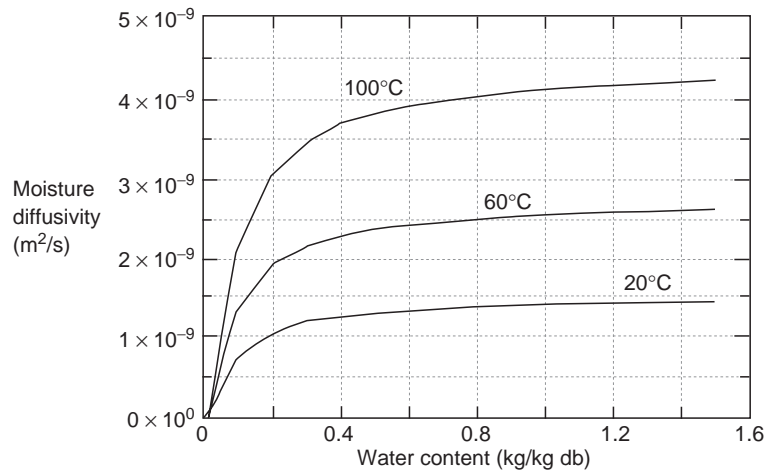
steady-state and transient-state methods. Transient methods are more popular because they can be run for as short as 10 s, during which time the moisture migration and other property changes are kept minimal.

4.3.2.1 Steady-State Methods

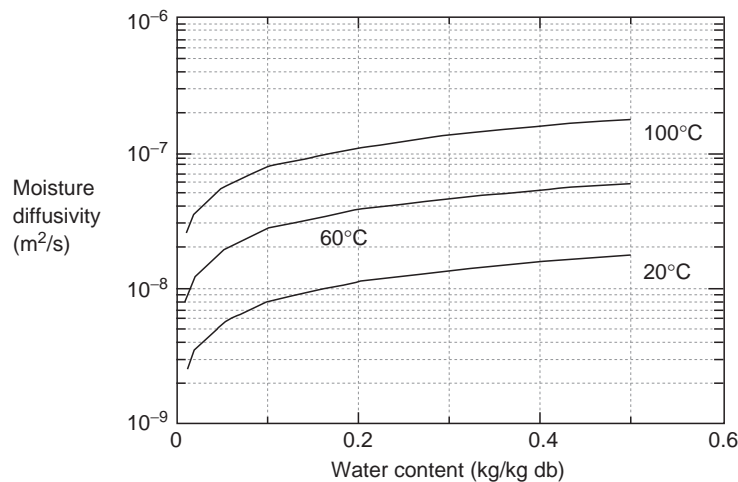
In steady-state methods, the temperature distribution of the sample is measured at steady state, with the sample placed between a heat source and a heat sink.

TABLE 4.4
Application Examples

Material	Equation	Constants	Ref.
Clay brick, burned clay	$D = D_0 (T/T_0)^{a_T} (X/X_0)^{a_X}$	$D_0 = 7.36 \times 10^{-9} \text{ m}^2/\text{s}, T_0 = 273 \text{ K}, a_T = 9.5,$ $X_0 = 0.35 \text{ kg/kg db}, a_X = 0.5 \text{ for clay brick};$ $D_0 = 1.11 \times 10^{-9} \text{ m}^2/\text{s}, T_0 = 273 \text{ K}, a_T = 6.5,$ $X_0 = 0.40 \text{ kg/kg db}, a_X = 0.5 \text{ for burned clay}$	61
Polyvinylalcohol	$D = D_0 \exp[-E/R(1/T - 1/T_0)],$ $D_0 = \sum a_i X^i$	$T_0 = 298 \text{ K}, E = 3.05 \times 10^4 \text{ J/mol},$ $R = 8.314 \text{ J/(mol K)}, a_0 = -0.104015 \times 10^2,$ $a_1 = 0.363457 \times 10^2, a_2 = -0.469291 \times 10^3,$ $a_3 = 0.634869 \times 10^4, a_4 = -0.517559 \times 10^5,$ $a_5 = 0.250188 \times 10^6, a_6 = -0.747613 \times 10^6,$ $a_7 = 0.139929 \times 10^7, a_8 = -0.159715 \times 10^7,$ $a_9 = 0.101503 \times 10^7, a_{10} = -0.274672 \times 10^6$	71
Potato, carrot	$D = D_0 \exp(-X_0/X) \exp(-T_0/T)$	$D_0 = 2.41 \times 10^{-7} \text{ m}^2/\text{s}, X_0 = 7.62 \times 10^{-2} \text{ kg/kg db},$ $T_0 = 1.49 \times 10^{+3} \text{ }^\circ\text{C for potato}; D_0 = 2.68 \times 10^{-4} \text{ m}^2/\text{s},$ $X_0 = 8.92 \times 10^{-2} \text{ kg/kg db}, T_0 = 3.68 \times 10^{+3} \text{ }^\circ\text{C for carrot}$	72
Silica gel	$D = D_0 \exp((E_0 - E_1 X)/T)$	$D_0 = 5.71 \times 10^{-7} \text{ m}^2/\text{s}, E_0 = 2450 \text{ K}, E_1 = 1400 \text{ K/(kg/kg db)}$	73



Potato



Clay brick

FIGURE 4.5 Effect of material moisture content and temperature on moisture diffusivity. Data for potato are from Kiranoudis, C.T., Maroulis, Z.B., and Marinos-Kouris, D., *Drying Technol.*, 10(4), 1097, 1992 and data for clay brick are from Haertling, M., in *Drying '80*, Vol. 1, A.S. Mujumdar (Ed.), Hemisphere Publishing, New York, 1980, pp. 88–98.

Different geometries can be used, those for longitudinal heat flow and radial heat flow.

4.3.2.2 Longitudinal Heat Flow (Guarded Hot Plate)

The longitudinal heat flow (guarded hot plate) method is regarded as the most accurate and most widely used apparatus for the measurement of thermal conductivity of poor conductors of heat. This method is most suitable for dry homogeneous specimens in slab forms. The details of the technique are given by the American Society for Testing and Materials (ASTM) Standard C-177 [82].

4.3.2.3 Radial Heat Flow

Whereas the longitudinal heat flow methods are most suitable for slab specimens, the radial heat flow techniques are used for loose, unconsolidated powder or granular materials. The methods can be classified as follows:

- Cylinder with or without end guards
- Sphere with central heating source
- Concentric cylinder comparative method

4.3.2.4 Unsteady State Methods

Transient-state or unsteady-state methods make use of either a line source of heat or plane sources of heat.

TABLE 4.5
Methods for the Experimental Measurement
of Thermal Conductivity

Method	Ref.
Steady-state method	
Longitudinal heat flow (guarded hot plate)	82
Radial heat flow	83
Unsteady-state method	
Fitch	84, 85
Plane heat source	86
Probe method	87, 88

In both cases, the usual procedure is to apply a steady heat flux to the specimen, which must be initially in thermal equilibrium, and to measure the temperature rise at some point in the specimen, resulting from this applied flux [83]. The Fitch method is one of the most common transient methods for measuring the thermal conductivity of poor conductors. This method was developed in 1935 and was described in the National Bureau of Standards Research Report No. 561. Experimental apparatus is commercially available.

4.3.2.5 Probe Method

The probe method is one of the most common transient methods using a line heat source. This method is simple and quick. The probe is a needle of good thermal conductivity that is provided with a heater wire over its length and some means of measuring the temperature at the center of its length. Having the probe embedded in the sample, the temperature response of the probe is measured in a step change of heat source and the thermal conductivity is estimated using the transient solution of Fourier's law. Detailed descriptions as well as the necessary modifications for the application of the above-mentioned methods in food systems are given in Refs. [83,89,90].

4.3.3 DATA COMPILATION

Despite the limited data of effective moisture diffusivity, a lot of data are reported in the literature for thermal conductivity. Data for mainly homogeneous materials are available in handbooks such as the *Handbook of Chemistry and Physics* [91], the *Chemical Engineers' Handbook* [92], *ASHRAE Handbook of Fundamentals* [93], Rohsenow and Choi [94], and many others. For foods and agricultural products, data are available in Refs. [83,88,95–97]. For selected pharmaceutical materials, data are presented by Pakowski and Mujumdar [98].

Some data for thermal conductivity are presented in Table 4.6. These values are distributed as shown in Figure 4.6. The distribution is different from that of moisture diffusivity (Figure 4.2), which is normal. For thermal conductivity, the values are uniformly distributed in the range 0.25 to 2.25 W/(m K), whereas a lot of data are accumulated below 0.25 W/(m K).

4.3.4 FACTORS AFFECTING THERMAL CONDUCTIVITY

The thermal conductivity of homogeneous materials depends on temperature and composition, and empirical equations are used for its estimation. For each material, polynomial functions of first or higher order

TABLE 4.6
Effective Thermal Conductivity in Some Materials

Material	Temperature (°C)	Thermal Conductivity (W/(m K))	Ref.
Aerogel, silica	38	0.022	94
Asbestos	427	0.225	94
Bakelite	20	0.232	94
Beef, 69.5% water	−18	0.622	99
Beef fat, 9% water	−10	0.311	100
Brick, common	20	0.173–0.346	94
Brick, fire clay	800	1.37	94
Carrots	−15 to −19	0.622	101
Concrete	20	0.813–1.40	94
Corkboard	38	0.043	94
Diatomaceous earth	38	0.052	94
Fiber-insulating board	38	0.042	94
Fish	−20	1.50	100
Fish, cod, and haddock	−20	1.83	102
Fish muscle	−23	1.82	103
Glass, window	20	0.882	94
Glass wool, fine	38	0.054	94
Glass wool, packed	38	0.038	94
Ice	0	2.21	94
Magnesia	38	0.067	94
Marble	20	2.77	94
Paper		0.130	94
Peach	18–27	1.12	104
Peas	18–27	1.05	104
Peas	−12 to −20	0.501	101
Plums	−13 to −17	0.294	101
Potato	−10 to −15	1.09	101
Potato flesh	18–27	1.05	104
Rock wool	38	0.040	94
Rubber, hard	0	0.150	94
Strawberries	18–27	1.35	104
Turkey breast	−25	0.167	100
Turkey leg	−25	1.51	100
Wood, oak	21	0.207	94

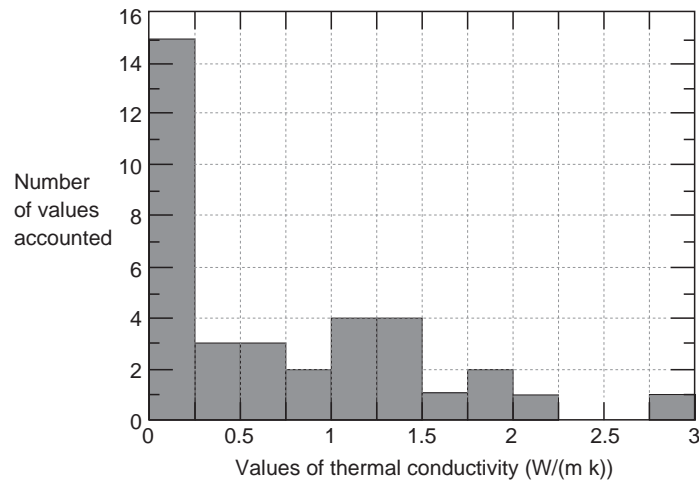


FIGURE 4.6 Distribution of thermal conductivity values (data from Table 4.5).

are used to express the temperature effect. A large number of empirical equations for the calculation of thermal conductivity as a function of temperature and humidity are available in the literature [83,92].

For heterogeneous materials, the effect of geometry must be considered using structural models. Utilizing Maxwell's and Eucken's work in the field of electricity, Luikov et al. [105] initially used the idea of an elementary cell, as representative of the model structure of materials, to calculate the effective thermal conductivity of powdered systems and solid porous materials. In the same paper, a method is proposed for the estimation of the effective thermal conductivity of mixtures of powdered and solid porous materials.

Since then, a number of structural models have been proposed, some of which are given in Table 4.7. The perpendicular model assumes that heat conduction

is perpendicular to alternate layers of the two phases, whereas the parallel model assumes that the two phases are parallel to heat conduction. In the mixed model, heat conduction is assumed to take place by a combination of parallel and perpendicular heat flow. In the random model, the two phases are assumed to be mixed randomly. The Maxwell model assumes that one phase is continuous, whereas the other phase is dispersed as uniform spheres. Several other models have been reviewed in Refs. [107,110,111], among others.

The use of some of these structural models to calculate the thermal conductivity of a hypothetical porous material is presented in Figure 4.7. The parallel model gives the larger value for the effective thermal conductivity, whereas the perpendicular model gives the lower value. All other models predict values in between. The use of structural models has been

TABLE 4.7
Structural Models for Thermal Conductivity in Heterogeneous Materials

Model	Equation	Ref.
Perpendicular (series)	$1/k = (1 - \varepsilon)/k_1 + \varepsilon/k_2$	106,107
Parallel	$k = (1 - \varepsilon)k_1 + \varepsilon k_2$	106,107
Mixed	$1/k = \frac{1 - F}{(1 - \varepsilon)k_1 + \varepsilon k_2} + F \left(\frac{1 - \varepsilon}{k_1} + \frac{\varepsilon}{k_2} \right)$	106,107
Random	$k = k_1^{(1-\varepsilon)} k_2^\varepsilon$	106,107
Effective medium theory	$k = k_1 [b + (b^2 + 2(k_1/k_2)(Z - 2))^{1/2}]$ $b = [Z(1 - \varepsilon)/2 - 1 + (k_2/k_1)(\varepsilon Z/2 - 1)]/(Z - 2)$	108
Maxwell	$k = \frac{k_2[k_1 + 2k_2 - 2(1 - \varepsilon)(k_2 - k_1)]}{k_1 + 2k_2 + (1 - \varepsilon)(k_2 - k_1)}$	109

k , Effective thermal conductivity; k_1 , thermal conductivities of phase 1; ε , void fraction of phase 2; F , Z , parameters.

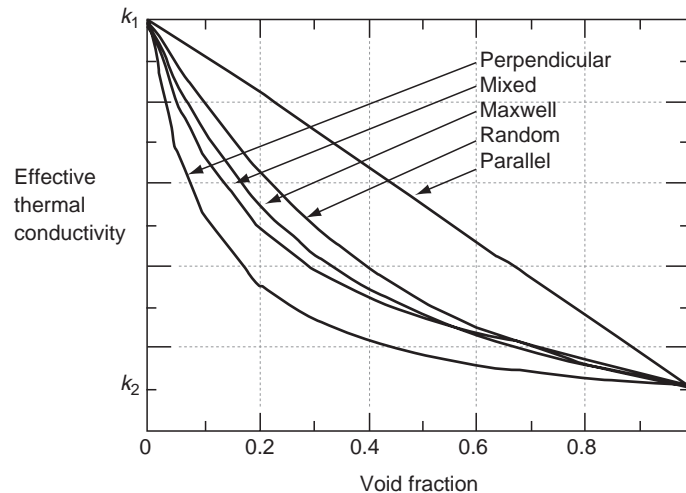


FIGURE 4.7 Effect of geometry on the thermal conductivity of heterogeneous materials using structural models.

successfully extended to foods [108,112], which exhibit a more complex structure than that of other materials, whereas this structure often changes during the heat conduction.

A systematic general procedure for selecting suitable structural models, even in multiphase systems, has been proposed in Ref. [113]. This method is based on a model discrimination procedure. If a component has unknown thermal conductivity, the method estimates the dependence of the temperature on the unknown thermal conductivity, and the suitable structural models simultaneously.

An excellent example of applicability of the above is in the case of starch, a useful material in extrusion. The granular starch consists of two phases, the wet granules and the air–vapor mixture in the intergranular space. The starch granule also consists of two phases, the dry starch and the water. Consequently, the thermal conductivity of the granular starch depends on the thermal conductivities of pure materials (i.e., dry pure starch, water, air, and vapor, all functions of temperature) and the structures of granular starch and the starch granule. It has been shown that the parallel model is the best model for both the granular starch and the starch granule [113]. These results led to simultaneous experimental determination of the thermal conductivity of dry pure starch versus temperature. Dry pure starch is a material that cannot be isolated for direct measurement.

4.3.5 THEORETICAL ESTIMATION

As in the case of the diffusion coefficient, the thermal conductivity in fluids can be predicted with satisfactory accuracy using theoretical expressions, such as the

formulas of Chapman and Enskog for monoatomic gases, of Eucken for polyatomic ones, or of Bridgman for pure liquids. The thermal conductivity of solids, however, has not yet been predicted using basic thermophysical or molecular properties, just like the analogous diffusion coefficient. Usually, the thermal conductivities of solids must be established experimentally since they depend upon a large number of factors that cannot be easily measured or predicted.

A large number of correlations are listed in the literature for the estimation of thermal conductivity as a function of characteristic properties of the material. Such relations, however, have limited practical utility since the values of the necessary properties are not readily available.

A method has been developed for the prediction of thermal conductivity as a function of temperature, porosity, material skeleton thermal conductivity, thermal conductivity of the gas in the porous, mechanical load on the porous material, radiation, and optical and surface properties of the material's particles [105]. The method produced satisfactory results for a wide range of materials (quartz sand, powdered Plexiglas, perlite, silica gel, etc.).

It has been proposed that the thermal conductivity of wet beads of granular material be estimated as a function of material content and the thermal conductivity of each of the three phases [114]. The results of the method were validated in a small number of materials such as crushed marble, slate, glass, and quartz sand.

Empirical equations for estimating the thermal conductivity of foods as a function of their composition have been proposed in the literature. In particular, it has been suggested that the thermal

conductivity of foods is a first-degree function of the concentrations of the constituents (water, protein, fat, carbohydrate, etc.) [97].

4.4 INTERPHASE HEAT AND MASS TRANSFER COEFFICIENTS

4.4.1 DEFINITION

The interphase heat transfer coefficient is related to heat transfer through a relative stagnant layer of the flowing air, which is assumed to adhere to the surface of the solid during drying (generally heating or cooling). It may be defined as the proportionality factor in the equation (Newton's law)

$$Q = h_H A (T_A - T) \quad (4.5)$$

where h_H (kW/(m² K)) is the surface heat transfer coefficient at the material–air interface, Q (kW) is the rate of heat transfer, A (m²) is the effective surface area, T (K) is the solid temperature at the interface, and T_A (K) is the bulk air temperature.

By analogy, a surface mass transfer coefficient can be defined using the following equation:

$$J = h_M A (X_A - X_{AS}) \quad (4.6)$$

where h_M (kg/(m² s)) is the surface mass transfer coefficient at the material–air interface, J (kg/s) is the rate of mass transfer, A (m²) is the effective surface area, X_{AS} (kg/kg) and X_A (kg/kg) are the air humidities at the solid interface and the bulk air.

Equation 4.5 and Equation 4.6 are used in cases in which the drying is externally controlled. This occurs when the Biot number (Bi_H , Bi_M) for heat and mass transfer is less than 0.1 [5].

Volumetric heat and mass transfer coefficients are often used instead of surface heat and mass transfer coefficients. They can be defined using the equations

$$h_{VH} = \alpha h_H \quad (4.7)$$

$$h_{VM} = \alpha h_M \quad (4.8)$$

where α is the specific surface defined as follows:

$$\alpha = A/V \quad (4.9)$$

where A (m²) is the effective surface area and V (m³) is the total volume of the material.

Different coefficients can be defined using different driving forces.

4.4.2 METHODS OF EXPERIMENTAL MEASUREMENT

The methods of experimental measurement of heat and mass transfer coefficients are summarized in Table 4.8, and resulted mainly from heat and mass transfer investigations in packed beds. Heat transfer techniques are either steady or unsteady state. In steady-state methods, the heat flow is measured together with the temperatures, and the heat transfer coefficient is obtained using Newton's law. Three different methods for heating are presented in Table 4.8. In unsteady-state techniques, the temperature of the outlet air is measured as a response to variations of the inlet air temperature. A transient model incorporating the heat transfer coefficient is used for analysis. Step, pulse, or cyclic temperature variations of the input air temperature have been used. Drying experiments during the constant drying rate period have also been used for estimating heat and mass transfer coefficients. A generalization of this method for simultaneous estimation of transport properties using drying experiments is presented in Section 4.7.

4.4.3 DATA COMPILATION

All the data available in the literature are in the form of empirical equations, and they are examined in the next section.

4.4.4 FACTORS AFFECTING THE HEAT AND MASS TRANSFER COEFFICIENTS

Both heat and mass transfer coefficients are influenced by thermal and flow properties of the air and, of course, by the geometry of the system. Empirical equations for various geometries have been proposed

TABLE 4.8
Methods for the Experimental Measurement of Heat and Mass Transfer Coefficients

Method	Ref.
<i>Steady-state heating methods</i>	
Material heating	115
Wall Heating	116
Microwave heating	117
<i>Unsteady-state heating methods</i>	
Step change of input air temperature	118,119
Pulse change of input air temperature	120,121
Cyclic temperature variation of input air	122,123
Constant rate drying experiments	124,125
Simultaneous estimation of transport properties using drying experiments	See Section 4.7

in the literature. Table 4.9 summarizes the most popular equations used for drying. The empirical equations incorporate dimensionless groups, which are defined in Table 4.10. Some nomenclature needed for understanding Table 4.9 is also included in Table 4.10.

Equation T9.1 through Equation T9.5 in Table 4.9 are the most widely used equations in estimating heat and mass transfer coefficients for simple geometries (packed beds, flat plates).

For packed beds, the literature contains many references. In 1965, Barker reviewed 244 relevant papers [183]. The equation suggested by Whitaker [130] is selected and presented in Table 4.9 as Equation

T9.7. It has been obtained by fitting to data of several investigators (see Refs. [126,127]). Equation T9.6 for flat plates comes from the same investigation [130], and it is also included in Table 4.9. In drying of granular materials, the equations reviewed in Ref. [136] should be examined.

Rotary dryers are usually controlled by heat transfer. Thus, Equation T9.8 through Equation T9.10 in Table 4.9 are proposed in Ref. [131] for the estimation of the corresponding heat transfer coefficients.

Heat and mass transfer in fluidized beds have been discussed in Refs. [6,137–140]. The latter reviewed the most important correlations and proposed Equation

TABLE 4.9
Equations for Estimating Heat and Mass Transfer Coefficients

Equation No.	Geometry	Equation	Ref.
T9.1	Packed beds (heat transfer)	$j_H = 1.06Re^{-0.41}$ $350 < Re < 4000$	126
T9.2	Packed beds (mass transfer)	$j_M = 1.82Re^{-0.51}$ $40 < Re < 350$	127
T9.3	Flat plate (heat transfer, parallel flow)	$j_H = 0.036Re^{-0.2}$ $500,000 < Re$	128
T9.4	Flat plate (heat transfer, parallel flow)	$h_H = 0.0204G^{0.8}$ $0.68 < G < 8.1$; $45 < T < 150^\circ\text{C}$	129
T9.5	Flat plate (heat transfer, perpendicular flow)	$h_H = 1.17 G^{0.37}$ $1.1 < G < 5.4$	129
T9.6	Flat plate (heat transfer, parallel flow)	$Nu = 0.036(Re^{0.8} - 9200)Pr^{0.43}$ $1.0 \times 10^5 < Re < 5.5 \times 10^6$	130
T9.7	Packed beds (heat transfer)	$Nu' = (0.5Re^{1/2} + 0.2Re^{2/3})Pr^{1/3}$ $2 \times 10^3 < Re' < 8 \times 10^3$	130
T9.8	Rotary dryer (heat transfer)	$j_H = 1.0Re^{-0.5}Pr^{1/3}$	131
T9.9	Rotary dryer (heat transfer)	$Nu = 0.33Re^{0.6}$	131
T9.10	Rotary dryer (heat transfer)	$h_{vH} = 0.52G^{0.8}$	131
T9.11a	Fluidized beds (heat transfer)	$Nu = 0.0133Re^{1.6}$ $0 < Re < 80$	6
T9.11b	Fluidized beds (heat transfer)	$Nu = 0.316Re^{0.8}$ $80 < Re < 500$	6
T9.12a	Fluidized beds (mass transfer)	$Sh = 0.374Re^{1.18}$ $0.1 < Re < 15$	6
T9.12b	Fluidized beds (mass transfer)	$Sh = 2.01Re^{0.5}$ $15 < Re < 250$	6
T9.13	Droplets in spray dryer (heat transfer)	$Nu = 2 + 0.6Re^{1/2}Pr^{1/3}$ $2 < Re < 200$	132
T9.14	Droplets in spray dryer (mass transfer)	$Sh = 2 + 0.6Re^{1/2}Sc^{1/3}$ $2 < Re < 200$	132
T9.15	Spouted beds (heat transfer)	$Nu = 5.0 \times 10^{-4} Re_s^{1.46} (u/u_s)^{1/3}$	6
T9.16	Spouted beds (mass transfer)	$Sh = 2.2 \times 10^{-4} Re^{1.45} (D/H_0)^{1/3}$	6
T9.17	Pneumatic dryers (heat transfer)	$Nu = 2 + 1.05Re^{1/2}Pr^{1/3}Gu^{0.175}$ $Re < 1000$	6
T9.18	Pneumatic dryers (mass transfer)	$Sh = 2 + 1.05Re^{1/2}Pr^{1/3}Gu^{0.175}$ $Re < 1000$	6
T9.19	Impingement drying	Several equations for various configurations	133–135

For nomenclature, see Table 4.10.

TABLE 4.10
Dimensionless Groups of Physical Properties

Name	Definition
Biot for heat transfer	$Bi_H = h_H d / 2k$
Biot for mass transfer	$Bi_M = h_M d / 2\rho\Delta$
Gukhman number	$Gu = (T_A - T) / T_A$
Heat transfer factor	$j_H = St Pr^{2/3}$
Mass transfer factor	$j_M = (h_M / u_A \rho_A) Sc^{2/3}$
Nusselt number	$Nu = h_H d / k_A$
Prandtl number	$Pr = c_p \mu / k_A$
Reynolds number	$Re = u_A \rho_A d / \mu$
Schmidt number	$Sc = \mu / \rho_A D_A$
Sherwood number	$Sh = h_M d / \rho_A D_A$
Stanton number	$St = h_H / u_A \rho_A c_p$

c_p , specific heat (kJ/(kg·K)); d , particle diameter (m); D , diffusivity in solid (m²/s); D_A , vapor diffusivity in air (m²/s); ε , void fraction in packed bed; G , mass flow rate of air (kg/(m² s)); h_H , heat transfer coefficient (kW/(m² K)); h_M , mass transfer coefficient (kg/(m² s)); h_{VH} , volumetric heat transfer coefficient (kW/(m³ K)); h_{VM} , volumetric mass transfer coefficient (kg/(m³ s)); k , thermal conductivity of solid (kW/(m·K)); k_A , thermal conductivity of air (kW/(m·K)); μ , dynamic viscosity of air (kg/(m·s)); Nu , $Nu' = Nu \varepsilon / (1 - \varepsilon)$; ρ_A , density of air (kg/m³); Re , $Re' = Re (1 - \varepsilon)$; Re_s , Re based on u_s instead of u ; T_A , air temperature (°C); T , material temperature (°C); u_A , air velocity (m/s); u_s , air velocity for incipient spouting (m/s).

T9.11 and Equation T9.12 of Table 4.9 for the calculation of heat and mass transfer coefficients, respectively. Further information for fluidized bed drying can be found in Ref. [141].

Vibration can intensify heat and mass transfer between the particles and gas. The following correction has been suggested for the heat and mass transfer coefficients when vibration occurs [6]

$$h_{H'} = h_H (A' f' / u_A)^{0.65} \quad (4.10)$$

$$h_{M'} = h_M (A' f' / u_A)^{0.65} \quad (4.11)$$

where u (m/s) is the air velocity, A (m) the vibration amplitude, and f (s⁻¹) the frequency of vibration. Further information on vibrated bed dryers can be found in Ref. [142].

For spray dryers, the popular equation of Ranz and Marshall [132] is presented in Table 4.9 (Equation T9.13 and Equation T9.14). They correlated data obtained for suspended drops evaporating in air.

Heat and mass transfer in a spouted bed has not been fully investigated yet because of the complex

character of the flow path of the particles in a bed with zones under different aerodynamic conditions [6]. However, Equation T9.15 and Equation T9.16 of Table 4.9 can be used.

Heat transfer coefficients for pneumatic dryers have been reviewed in Ref. [6]. The majority of authors examined and use an equation similar to Equation T9.13 and Equation T9.14 of Table 4.9 for spray dryers. For immobile particles, the exponent of the Re number is close to 0.5 and for free-falling particles, it is 0.8. Equation T9.17 of Table 4.9 is proposed. The mass transfer coefficient could be estimated by the analogy $Sh = Nu$ [6]. In extensive reviews [133–135], correlations for estimating heat and mass transfer coefficients in impingement drying under various configurations are discussed.

The calculated heat and mass transfer coefficients using some of the equations presented in Table 4.9 are plotted versus air velocity with some simplifications in Figure 4.8 and Figure 4.9. These figures can be used to estimate approximately the heat and mass transfer coefficients for various dryers. The simplifications made for the construction of these figures concern the drying air and material conditions. For instance, the air temperature is taken as 80°C, the air humidity as 0.010 kg/kg db, and the particle size as 10 mm (typical drying conditions). For other conditions, the equations of Table 4.9 should be used.

4.4.5 THEORETICAL ESTIMATION

No theory is available for estimating the heat and mass transfer coefficients using basic thermophysical properties. The analogy of heat and mass transfer can be used to obtain mass transfer data from heat transfer data and vice versa. For this purpose, the Chilton–Colburn analogies can be used [129]

$$j_M = j_H = f/2 \quad (4.12)$$

where f is the well-known Fanning friction factor for the fluid, and j_H and j_M are the heat and mass transfer factors defined in Table 4.10. Discrepancies of the above classical analogy have been discussed in Ref. [143].

In air conditioning processes, the heat and mass transfer analogy is usually expressed using the Lewis relationship

$$h_H / h_M = c_p \quad (4.13)$$

where c_p (kJ/(kg K)) is the specific heat of air.

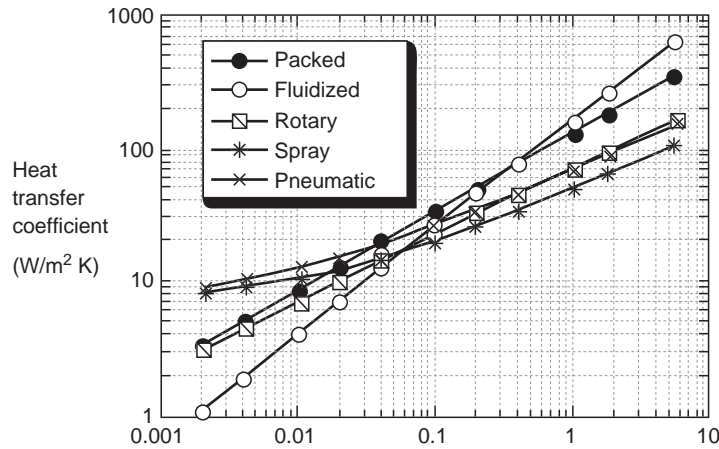


FIGURE 4.8 Heat transfer coefficients versus air velocity for some dryers (particle size 10 mm; drying conditions $T_A = 80^\circ\text{C}$, $X_A = 10 \text{ g/kg db}$).

4.5 DRYING CONSTANT

4.5.1 DEFINITION

The transport properties discussed above (moisture diffusivity, thermal conductivity, interface heat, and mass transfer coefficients) describe completely the drying kinetics. However, in the literature sometimes (mainly in foods, especially in cereals) instead of the above transport properties, the drying constant K is used. The drying constant is a combination of these transport properties.

The drying constant can be defined using the so-called thin-layer equation. Lewis suggested that during the drying of porous hygroscopic materials, in the falling rate period, the rate of change in material

moisture content is proportional to the instantaneous difference between material moisture content and the expected material moisture content when it comes into equilibrium with the drying air [144]. It is assumed that the material layer is thin enough or the air velocity is high so that the conditions of the drying air (humidity, temperature) are kept constant throughout the material. The thin-layer equation has the following form:

$$-dX/dt = K(X - X_e) \quad (4.14)$$

where X (kg/kg db) is the material moisture content, X_e (kg/kg db) is the material moisture content in equilibrium with the drying air, and t (s) is the time.

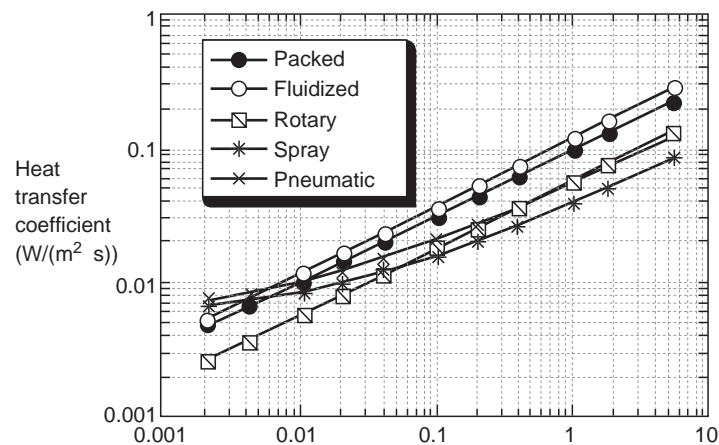


FIGURE 4.9 Mass transfer coefficients versus air velocity for some dryers (particle size 10 mm; drying conditions $T_A = 80^\circ\text{C}$, $X_A = 10 \text{ g/kg db}$).

A review of several other thin-layer equations can be found in Refs. [76,145].

Equation 4.14 constitutes an effort toward a unified description of the drying phenomena regardless of the controlling mechanism. The use of similar equations in the drying literature is ever increasing. It is claimed, for example, that they can be used to estimate the drying time as well as for the generalization of the drying curves [6].

The drying constant K is the most suitable quantity for purposes of design, optimization, and any situation in which a large number of iterative model calculations are needed. This stems from the fact that the drying constant embodies all the transport properties into a simple exponential function, which is the solution of Equation 4.14 under constant air conditions. On the other hand, the classical partial differential equations, which analytically describe the four prevailing transport phenomena during drying (internal–external, heat–mass transfer), require a lot of time for their numerical solution and thus are not attractive for iterative calculations.

4.5.2 METHODS OF EXPERIMENTAL MEASUREMENT

The measurement of the drying constant is obtained from drying experiments. In a drying apparatus, the air temperature, humidity, and velocity are controlled and kept constant, whereas the material moisture content is monitored versus time. The drying constant is estimated by fitting the thin-layer equation to experimental data.

4.5.3 FACTORS AFFECTING THE DRYING CONSTANT

The drying constant depends on both material and air properties as it is a phenomenological property representative of several transport phenomena. So, it is a function of material moisture content, temperature, and thickness, as well as air humidity, temperature, and velocity.

Some relationships describing the effect of the above factors on the drying constant are presented in Table 4.11. Equation T11.1 and Equation T11.2 are Arrhenius-type equations, which take into account the temperature effect only. The effect of water activity can be considered by modifying the activation energy (Equation T11.1) on the preexponential factor (Equation T11.2). Equation T11.1 and Equation T11.2 consider the same factors in a different form. Equation T11.4 takes into account only the air velocity effect, whereas Equation T11.5 considers all the factors affecting the drying constant. Table 4.12 lists parameter values for typical equations of Table 4.11.

Equation T11.2 and Equation T11.5 were applied to shelled corn [150] and to green pepper [35], respectively, and the results are presented in Figure 4.10. The effects of air temperature and velocity, as well as particle dimensions, are shown for green pepper drying, whereas the air temperature and the small air–water activity effects are shown for the low air temperature drying of wheat.

4.5.4 THEORETICAL ESTIMATION

It is impossible to estimate an empirical constant using theoretical arguments. The estimation of an

TABLE 4.11
Effect of Various Factors on the Drying Constant

Equation No.	Materials of Application	Equation	Ref.
T11.1a	Grains, barley, various tropical agricultural products	$K(T_A) = b_0 \exp[-b_1/T_A]$	75,146,147
T11.1b	Barley, wheat	$K(T_A) = b_0 \exp[-b_1/(b_2 + b_3 T_A)]$	148
T11.2a	Melon	$K(a_w, T_A) = b_0 \exp[-(b_1 + b_2 a_w)/T_A]$	149
T11.2b	Corn, shelled	$K(a_w, T_A) = b_0 \exp(-b_1 a_w) \exp[-b_2/(b_3 + b_4 T_A)]$	150
T11.3a	Rice	$K(a_w, T_A) = b_0 + b_1 T_A - b_2 a_w$	151
T11.3b	Wheat	$K(a_w, T_A) = b_0 + b_1 T_A^2 - b_2 a_w$	152
T11.4	Carrot	$K(u_A) = \exp(-b_1 + b_2 \ln u_A)$	153
T11.5	Potato, onion, carrot, pepper	$K(a_w, T_A, d, u_A) = b_0 a_w^{b_1} T_A^{b_2} d^{b_3} u_A^{b_4}$	35

K , Drying constant; T_A , temperature; u_A , air velocity; a_w , water activity; d , particle diameter; b_1 , parameters.

TABLE 4.12
Application Examples

Material	Equation	Constants	Ref.
Shelled corn	$K = b_0 \exp(-b_1 a_w) \exp[-b_2/(b_3 + b_4 T_A)]$ $0.1 < a_w < 0.6, 23.5 < T_A < 56.9^\circ\text{C}$	$b_0 = 170/\text{s}, b_1 = 1.15, b_2 = 8259,$ $b_3 = 492, b_4 = 1.8/^\circ\text{C}$	150
Green pepper	$K = b_0 X_A^{b_1} T_A^{b_2} d^{b_3} u_A^{b_4}$ $0.006 < X_A < 0.022 \text{ kg/kg db},$ $60 < T_A < 90^\circ\text{C}, 0.005 < d < 0.015 \text{ m}, 3 < u_A < 5 \text{ m/s}$	$b_0 = 1.11 \times 10^{-8}/\text{s}, b_1 = 9.03 \times 10^{-2},$ $b_2 = 1.54, b_3 = -0.982, b_4 = 0.293$	35

Source: From Brunauer, S., Deming, L.S., Deming, W.E., and Teller, E., *Am. Chem. Soc. J.*, 62, 1723, 1940. With permission.

empirical constant using theoretical arguments has little, if any, meaning. Nevertheless, if we assume that for some drying conditions the controlling mechanism is the moisture diffusion in the material, then the drying constant can be expressed as a function

of moisture diffusivity. For slabs, for example, the following equation is valid:

$$K = \pi^2 D/L^2 \quad (4.15)$$

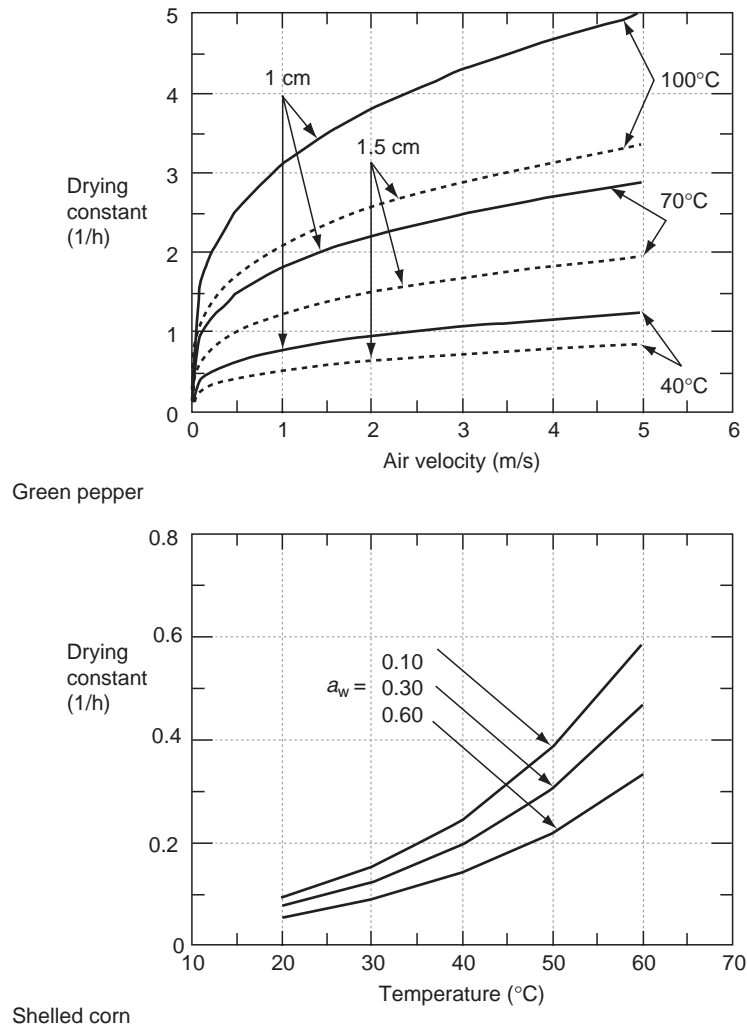


FIGURE 4.10 Effect of various factors on the drying constant. Data for green pepper are from Kiranoudis, C.T., Maroulis, Z.B., and Marinos-Kouris, D., *Drying Technol.*, 10(4), 995, 1992 and data for shelled corn are from Westerman, P.W., White, G.M., and Ross, I.J., *Trans. ASAE*, 16, 1136, 1973.

where D (m^2/s) is the effective diffusivity and L (m) is the thickness of the slab.

4.6 EQUILIBRIUM MOISTURE CONTENT

4.6.1 DEFINITION

A knowledge of the state of thermodynamic equilibrium between the surrounding air and the solid is a basic prerequisite for drying, as it is for any similar mass transfer situation.

The moisture content of the material when it comes into equilibrium with drying air is a useful property included in most drying models. The relation between equilibrium material moisture content and the corresponding water activity for a given temperature is known as the *sorption isotherm*. The water activity a_w at the pressures and temperatures that usually prevail during drying is equal to the relative humidity of air.

The equilibrium moisture of a material can be attained either by adsorption or by desorption, as expressed by the respective isotherms of Figure 4.11. The usually observed deviation of the two curves is due to the phenomenon of hysteresis, which has not yet been quantitatively described. Many explanations for the phenomenon have been put forth that converge in that there are more active sites during the desorption than during adsorption. It is clear from Figure 4.11 that the desorption isotherm is the curve to use for the process of drying.

In essence, the sorption isotherms express the minimum value of material moisture content that can be reached by a solid during drying in relation to the relative humidity of the drying air. On the basis of such isotherms, the equilibrium material moisture

content can be calculated. Such equilibrium values are necessary for the formulation of the mass transfer driving forces.

Moreover, the isotherms determine the proper storage environment and the packaging conditions, especially for foods. Through the isotherms, the isosteric heat of sorption can be determined and, hence an accurate prediction can be made of the energy requirements for the drying of a solid. The utility of the isotherm is extended to the determination of the moisture sorption mechanism as well as to the degree of bound water.

Brunauer et al. [154] classified the sorption isotherms into five different types (see Figure 4.12). The sorption isotherms of the hydrophilic polymers, such as natural fibers and foods, are of type II. The isotherms of the less hydrophilic rubbers, plastics, synthetic fibers, and foods rich in soluble components are of type III. The isotherms of certain inorganic materials (such as aluminum oxides) are of type IV. For many materials, however, the sorption isotherms cannot be properly classified since they belong to more than one type.

4.6.2 METHODS OF EXPERIMENTAL MEASUREMENT

A comprehensive review of existing experimental measuring methods is given in Refs. [155,156]. Sorption isotherms can be determined according to two basic principles, gravimetric and hygrometric.

4.6.2.1 Gravimetric Methods

During the measurement, the air temperature and the water activity are kept constant until the moisture content of the sample attains the constant equilibrium

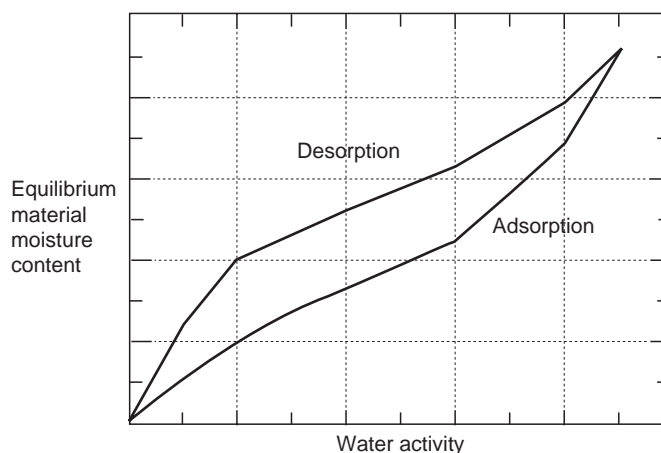


FIGURE 4.11 Hysteresis between adsorption and desorption isotherms.

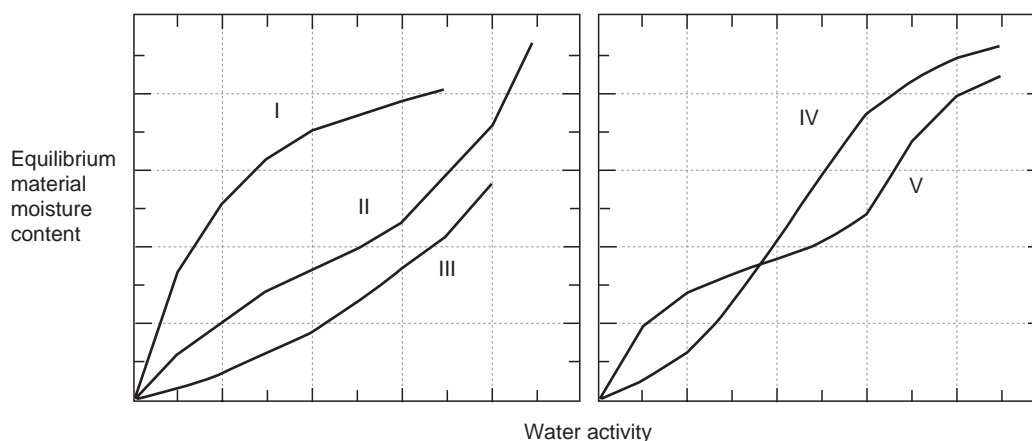


FIGURE 4.12 The five types of isotherms. (From Brunauer, S., Deming, L.S., Deming, W.E., and Teller, E., *Am. Chem. Soc. J.*, 62, 1723, 1940.)

value. The air may be circulated (dynamic methods) or stagnant (static). The material weight may be registered continuously (continuous methods) or discontinuously (discontinuous methods).

4.6.2.2 Hygrometric Methods

During the measurement, the material moisture content is kept constant until the surrounding air attains the constant equilibrium value. The air–water activity is measured via hygrometer or manometer.

The working group in the COST 90bis Project has developed a reference material (microcrystalline cellulose, MCC) and a reference method for measuring water sorption isotherms, and conducted a collaborative study to determine the precision (repeatability and reproducibility) with which the sorption isotherm of the reference material may be determined by the reference method. A detailed procedure for the resulting standardized method was presented, and the factors influencing the results of the method were discussed [157–159].

4.6.3 DATA COMPILATION

A large volume of data of equilibrium moisture content appears in the literature. Data for more than 35 polymeric materials, such as natural fibers, proteins, plastics, and synthetic fibers, are given in Ref. [8]. Isotherms for 32 materials (organic and inorganic) are also given in Ref. [92]. The literature is especially rich in sorption isotherms of foods due to the fact that the value of water activity is a critical parameter for food preservation safety and quality.

A bibliography on sorption isotherms of food materials is presented in Ref. [160]. The collection

comprises 2200 references, including about 900 papers with information on equilibrium moisture content of foods in defined environments. The papers are listed alphabetically according to the names of the first author, but they are also grouped according to product.

Additional bibliographies should also be mentioned. The *Handbook of Food Isotherms* contains more than 1000 isotherms, with a mathematical description of over 800 [161]. About 460 isotherms were obtained from the monograph of Ref. [162]. Data on sorption properties of selected pharmaceutical materials are presented in Ref. [98].

4.6.4 FACTORS AFFECTING THE EQUILIBRIUM MOISTURE CONTENT

Equilibrium material moisture content depends upon many factors, among which are the chemical composition, the physical structure, and the surrounding air conditions. A large number of equations (theoretical, semiempirical, empirical) have been proposed, none of which, however, can describe the phenomenon of hysteresis. Another basic handicap of the equations is that their applicability is not satisfactory over the entire range of water activity ($0 \leq a_w \leq 1$).

Table 4.13 lists the best-known isotherm equations. The Langmuir equation can be applied in type I isotherm behavior. The Brunauer–Emmet–Tetter (BET) equation has been successfully applied to almost all kinds of materials, but especially to hydrophilic polymers for $a_w < 0.5$. The Halsey equation is suitable for materials of types I, II, and III. The Henderson equation is less versatile than that of Halsey. For cereal and other field crops, the Chung

TABLE 4.13
Effect of Water Activity and Temperature on the Equilibrium Moisture Content

Equation Name	Equation	Ref.
Langmuir	$a_w \left(\frac{1}{X} - \frac{1}{b_0} \right) = \frac{1}{b_0 b_1}$	163
Brunauer–Emmet–Tetter (BET)	$\frac{a_w}{(1 - a_w)X} = \frac{1}{b_0 b_1} + \frac{b_1 - 1}{b_0 b_1} a_w$	164
Halsey	$a_w = \exp \left[-\frac{b_1}{RT} \left(\frac{X}{b_2} \right)^{b_3} \right]$	165
Henderson	$1 - a_w = \exp[-b_1 T X^{b_2}]$	166
Chung and Pfof	$\ln a_w = -\frac{b_1}{RT} \exp(-b_2 X)$	167
Chen and Clayton	$\ln a_w = -b_1 T^{b_2} \exp(-b_3 T^{b_2} X)$	168
Iglesias and Chirife	$\ln a_w = -\exp[(b_1 T + b_2) X^{b_3}]$	169
Guggenheim–Anderson–de Boer (GAB)	$X = \frac{b_0 b_1 b_2 a_w}{(1 - b_1 a_w)(1 - b_1 a_w + b_1 b_2 a_w)}$ $b_1 = b_{10} \exp(b_{11}/RT)$, $b_2 = b_{20} \exp(b_{21}/RT)$	170, 171

X , Equilibrium material moisture content; a_w , water activity; T , temperature; b_1 , parameters.

and Pfof equation is considered suitable, whereas that of Iglesias and Chirife has been successfully applied on isotherms of type III (i.e., foods rich in soluble components).

The Guggenheim–Anderson–de Boer (GAB) equation is considered as the most versatile model, capable of application to situations over a wide range of water activities ($0.1 < a_w < 0.9$) and to various materials (inorganic, foods, etc.). The GAB equation is probably the most suitable for process analysis and design of drying because of its reliability, its simple mathematical form, and its wide use (with materials and water activity ranges). Table 4.14 lists parameter values of the GAB equation for some foods.

Two selected food materials are presented as an example in Figure 4.13. Potatoes exhibit a typical behavior. Equilibrium material moisture content is increased [172]. Raisins, on the other hand, exhibit an inverse temperature effect at large water activities [173]. As shown in Figure 4.13, potatoes and raisins exhibit sorption isotherms of types II and III, respectively.

The isotherms at 25°C for some organic and inorganic materials are presented in Figure 4.14 [92]. In Figure 4.14, one can observe the various isotherm types, like type I for activated charcoal and silica gel, type II for leather, type III for soap, and so on.

Various regression analysis methods for fitting the above equations to experimental data have been discussed in the literature. The direct nonlinear regression exhibits several advantages over indirect

TABLE 4.14
Application of the Guggenheim–Anderson–de Boer Model to Some Fruits and Vegetables

Material	b_0	$b_{10} \times 10^5$	b_{11}	b_{20}	b_{21}
Potato	8.7	1.86	34.1	5.68	6.75
Carrot	21.2	5.94	28.9	8.03	5.49
Tomato	18.2	1.99	34.5	5.52	6.70
Pepper	21.1	1.46	33.4	5.56	6.56
Onion	20.2	2.30	32.5	5.79	6.43
Raisin	12.5	0.17	22.4	1.77	−1.53
Fig	11.7	0.05	25.2	1.77	−1.55
Prune	13.3	0.07	23.9	1.82	−1.65
Apricot	15.1	0.11	21.1	2.13	−2.05

Source: From Kiranoudis, C.T., Maroulis, Z.B., Tsami, E., and Marinos-Kouris, D., *J. Food Eng.*, 20(1), 55, 1992; Maroulis, Z.B., Tsami, E., Marinos-Kouris, D., and Saravacos, G.D., *J. Food Eng.*, 7(1), 63, 1988.

nonlinear regression [173]. Linear regression, on the other hand, can give highly erroneous results and should be avoided [174]. When there exist differences in the variance of the data, the direct nonlinear weighted regression method should be used [175].

4.7 SIMULTANEOUS ESTIMATION OF HEAT AND MASS TRANSPORT PROPERTIES FROM DRYING EXPERIMENTS

4.7.1 PRINCIPLES OF ESTIMATION

In the previous sections, methods of experimental determination of heat and mass transport properties have been discussed. These methods use special apparatus and are based on the equation of definition of the corresponding property. This section discusses the experimental determination of these properties from drying experiments. Some relevant techniques have been already discussed by Molnar [125]. However, a generalized method based on model-building techniques is presented here. The method uses a drying experimental apparatus and estimates the heat and mass transport properties as parameters of a drying model that incorporates these properties [28,43,176–180]. An outline of the method is described below.

First, an experimental drying apparatus is used. In such an apparatus, the air passes through the drying material and the air humidity, temperature, and velocity are controlled, whereas the material moisture content and, eventually, the material temperature are monitored versus time. Second, a mathematical

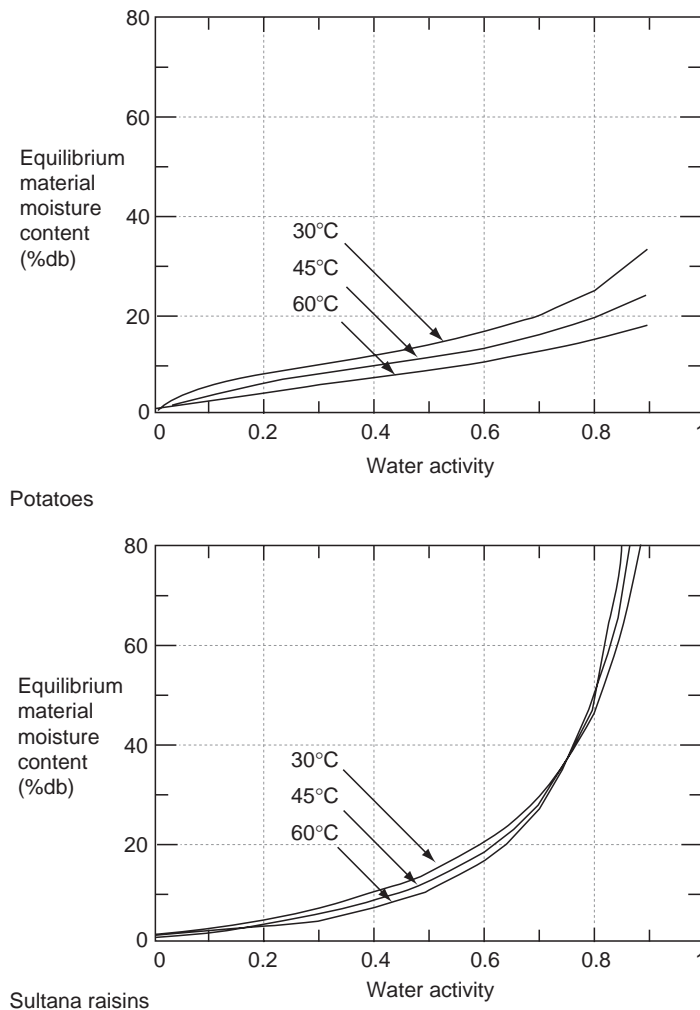


FIGURE 4.13 Effect of air–water activity and temperature on equilibrium material moisture content for two foods. (Data for potatoes from Kiranoudis, C.T., Maroulis, Z.B., Tsami, E., and Marinos-Kouris, D., *J. Food Eng.*, 20(1), 55, 1992 and data for sultana raisins from Maroulis, Z.B., Tsami, E., Marinos-Kouris, D., and Saravacos, G.D., *J. Food Eng.*, 7(1), 63, 1988.)

model that takes into account the controlling mechanisms of heat and mass transfer is considered. This model includes the heat and mass transport properties as model parameters or, even more, includes the functional dependence of the relevant factors on the transport properties. Third, a regression analysis procedure is used to obtain the transport properties as model parameters by fitting the model to experimental data of material moisture content and temperature.

Theoretically, all the properties describing the drying kinetics could be estimated simultaneously. We can define the drying kinetics (in an analogous manner to reaction kinetics) as the dependence of factors affecting the drying on the drying rate. Drying is not a chemical reaction, but it involves simultaneous heat and mass transfer phenomena. Consequently, the

properties describing these phenomena describe the drying process as well.

If, for example, the phenomena considered are

The moisture diffusion in the solid toward its external surface

The vaporization and convective transfer of the vapor into the airstream

The conductive heat transfer within the solid mass

The convective heat transfer from the air to the solid's surface

then the following properties describe the drying kinetics:

Effective moisture diffusivity

Air boundary mass transfer coefficient

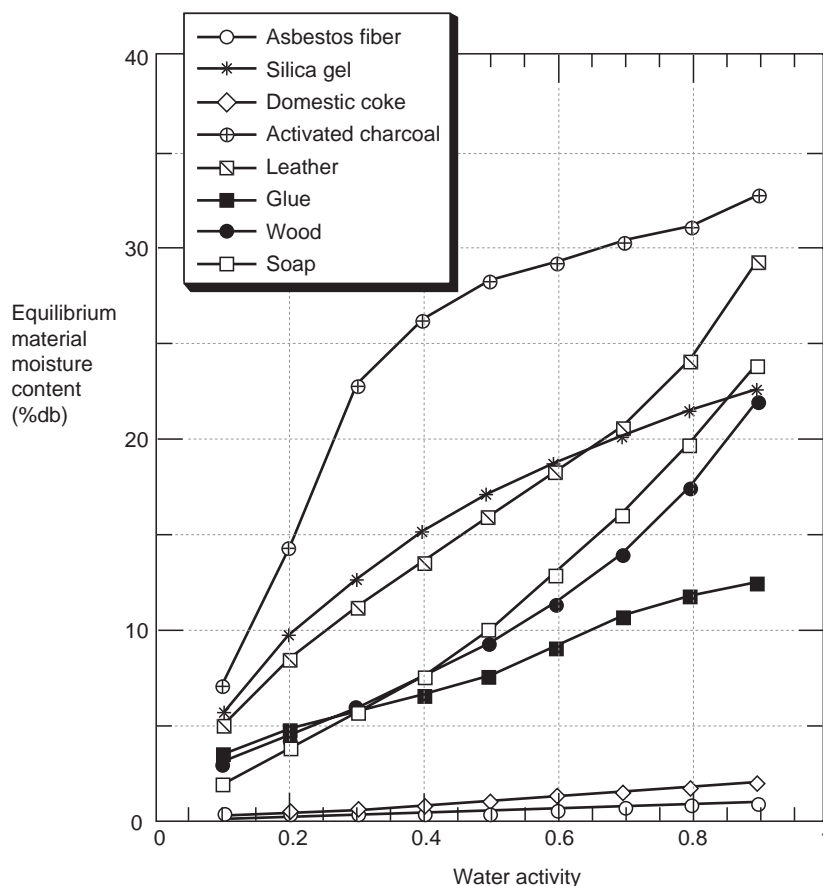


FIGURE 4.14 Equilibrium material moisture content for some organic and inorganic materials. (Data from Perry, R.H. and Chilton, C.H., *Chemical Engineers' Handbook*, 4th and 5th ed., McGraw-Hill, New York, 1963, 1973.)

Effective thermal conductivity
Air boundary heat transfer coefficient

and consequently they can be estimated.

Alternatively, if the drying constant is assumed to describe the drying kinetics by the thin-layer equation, then the drying constant can be estimated using this method.

4.7.2 EXPERIMENTAL DRYING APPARATUS

A typical drying apparatus is shown in Figure 4.15. The apparatus consists of two parts, the air conditioning section and the measuring section. The air conditioning section includes the heater, the humidifier, and the fan, which are handled via a temperature, a humidity, and a flow controller, respectively. In the measuring section, the air properties, that is, temperature, humidity, and velocity, as well as the material properties (weight and temperature) are continuously recorded. The use of a computer for online measurement and control is preferable.

4.7.3 THE DRYING MODEL

An information flow diagram for a drying model appropriate for this method is shown in Figure 4.16. This model can calculate the material moisture content and temperature as a function of position and time whenever the air humidity, temperature, and velocity are known as a function of time, together with the model parameters. If the model takes into account the controlling mechanisms of heat and mass transfer, then the transport properties (moisture diffusivity, thermal conductivity, boundary heat and mass transfer coefficients) are included in the model as parameters. If the dependence of drying conditions (material moisture content, temperature, and thickness, as well as air humidity, temperature, and velocity) on transport properties is also considered, then the constants of the relative empirical equations are considered as model parameters. In Figure 4.16 the part of the model that contains equations for the heat and mass transfer phenomena is termed the *process model*, whereas the equations describing the dependence

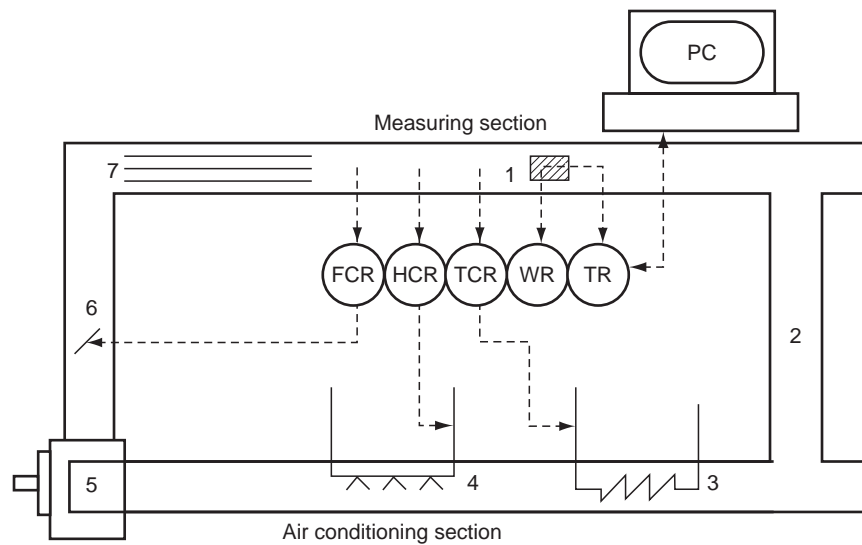


FIGURE 4.15 Typical experimental drying apparatus: (1) sample; (2) air recirculating duct; (3) heater; (4) humidifier; (5) fan; (6) valve; (7) straighteners; FCR, airflow control and recording; HCR, air humidity control and recording; TCR, air temperature control and recording; WR, sample weight recording; TR, sample temperature recording; PC, personal computer, for on-line measurement and control.

of drying conditions on transport properties form the *properties model*.

In the process model, each mechanism of heat and mass transfer is expressed using a driving force and a transport property as a coefficient of proportionality between the rate and the corresponding driving force. In the properties model, several formulas can be considered. Some assumptions have been suggested in the previous sections.

4.7.4 REGRESSION ANALYSIS

The parameters of a model can be estimated by fitting the model to experimental data [181,182]. Using the

model of [Section 4.7.3](#), two regression analysis procedures can be applied [43]: transport properties estimation and transport properties equations estimation.

4.7.4.1 Transport Properties Estimation

It is assumed that during the drying experiments the drying conditions are not varying very much with time, and the transport properties can be considered constant (not functions of the drying conditions). The transport properties are estimated as parameters of the process model by fitting it to experimental data. Only the properties of the controlling mechanisms can be obtained. Consequently, the precision and

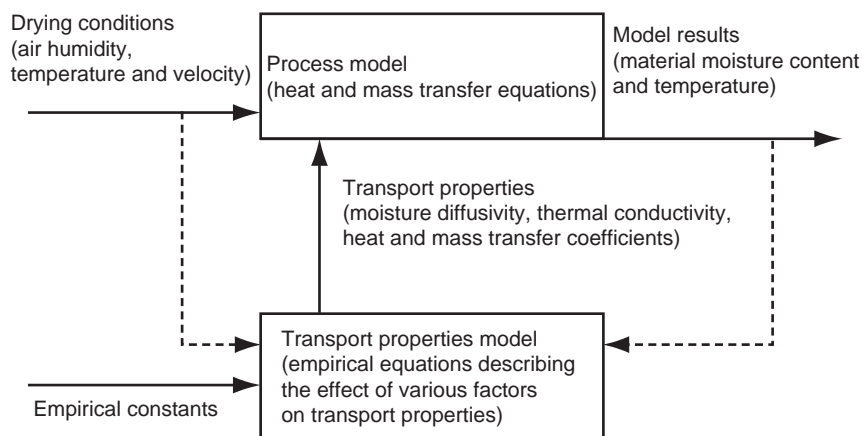


FIGURE 4.16 Model information flow diagram.

correlations of the estimates should be examined. A model discrimination procedure is suggested to discard the noncontrolling mechanisms.

4.7.4.2 Transport Properties Equations Estimation

Several empirical equations describing the dependence of transport properties on various factors are tested using a model discrimination procedure. The constants of the empirical equations are estimated as parameters of the total model (process model plus properties model) by fitting it to experimental data.

The information flow diagram for the regression analysis proposed is shown in Figure 4.17.

4.7.5 APPLICATION EXAMPLE

The method described above is applied to a wide set of experimental data in potato drying [43].

4.7.5.1 Experimental Drying Apparatus

An experimental drying apparatus similar to that shown in Figure 4.15 was used [35]. In each experiment, the air–water activity, temperature, and velocity were controlled, and the material moisture content and temperature were monitored versus time. A total number of 100 experiments were performed for three different particle dimensions (5, 10, and 15 mm) at five air temperature (60, 65, 70, 75, and 80°C), three air velocities (3, 4, and 5 m/s) and at air humidity ranging from 6 to 22 g/kg db.

4.7.5.2 Drying Model

A mathematical drying model involving simultaneous heat and mass transfer is considered for the analysis [43]. The model considered has the following form:

Moisture diffusion into the solid

$$\partial(\rho X)/\partial t = \nabla(\rho D \nabla X) \quad (4.16)$$

$$D = a_0 \exp(-a_1/X) \exp(-a_2/T) \quad (4.17)$$

Boundary layer vapor transfer

$$-(\rho D \nabla X) = h_M(a_{we} - a_w) \quad (4.18)$$

$$h_M = c_0 X_A^{c_1} T_A^{c_2} u_A^{c_3} \quad (4.19)$$

Heat conduction in the solid

$$\partial(\rho h_s)/\partial t = \nabla(k \nabla T) \quad (4.20)$$

$$k = b_0 \exp(-b_1/X) \exp(-b_2/T) \quad (4.21)$$

Boundary layer heat transfer

$$-(k \nabla T) = h_H(T - T_A) - \Delta H_s h_M(a_{we} - a_w) \quad (4.22)$$

$$h_H = d_0 X_A^{d_1} T_A^{d_2} u_A^{d_3} \quad (4.23)$$

where X and T are the material moisture content and temperature, respectively, u_A , a_w , and T_A are the air velocity, water activity, and temperature, respectively. The thermophysical and thermodynamic properties, material density ρ , material specific enthalpy h_s , heat of vaporization of water ΔH_s , and equilibrium air–water activity a_{we} are known functions of material moisture content and temperature.

The transport properties, moisture diffusivity D , and thermal conductivity k are functions of material moisture content and temperature, whereas heat and mass transfer coefficients, h_H , h_M , are functions of air velocity, water activity, and temperature.

The following adjustable constants are introduced to the relevant properties model: a_i , b_i , c_i , d_i .

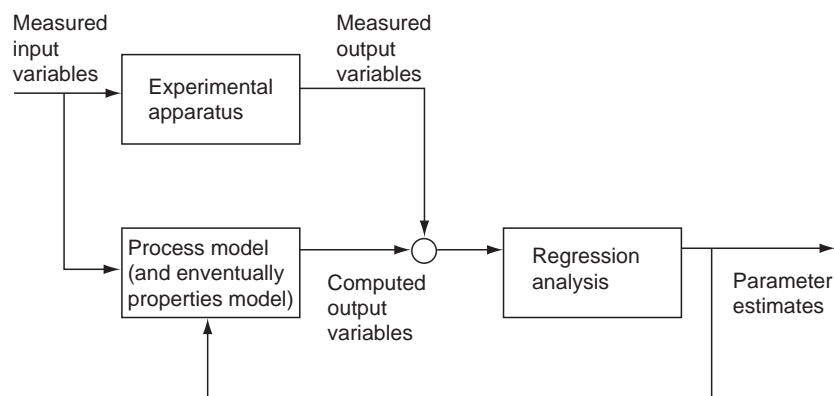


FIGURE 4.17 Regression analysis information flow diagram.

4.7.5.3 Regression Analysis

If X_i and T_i are the experimental values of material moisture content and temperature and $X_{i,c}$ and $T_{i,c}$ are the corresponding calculated values using the mathematical model, then the relative deviations between experimental and calculated values (relative residuals) can be defined as follows:

$$e_{X_i} = (X_{i,c} - X_i)/X_i \quad (4.24)$$

$$e_{T_i} = (T_{i,c} - T_i)/T_i \quad (4.25)$$

The relative standard deviations between experimental and calculated values of material moisture content S_X and temperature S_T are defined as follows:

$$S_{X^2} = \sum e_{X_i^2} / N \quad (4.26)$$

$$S_{T^2} = \sum e_{T_i^2} / N \quad (4.27)$$

where N is the number of experimental points (including different measurements and different experiments).

A linear combination of S_X and S_T is used for parameter estimation and the resulting S_X , S_T are used for model validation [180,182]. The regression analysis is performed simultaneously on all experiments.

4.7.5.4 Results

The application of the proposed method proved that:

The moisture diffusivity is a function of material moisture content and temperature.

The thermal conductivity is high and cannot be estimated from these experiments.

Heat and mass transfer coefficients are constant in the region of experimentation.

More specifically, the results obtained are as follows:

$$D = 2.94 \times 10^{-7} \text{m}^2/\text{s} \exp(-1.58 \times 10^3 K/T) \exp(-6.72 \times 10^{-2} \text{kg}/\text{kg}/X) \quad (4.28)$$

$$h_M = 5.84 \times 10^{-7} \text{kg}/\text{m}^2/\text{s} \quad (4.29)$$

$$h_H = 1.64 \times 10^{-1} \text{W}/\text{m}^2 \quad (4.30)$$

The resulting model calculates the material moisture content and temperature close to experimental values and it is considered satisfactory.

4.8 TRANSPORT PROPERTIES OF FOODS

The transport properties of foods received much attention in the literature [184–188]. The main results presented by Saravacos and Maroulis [188] are summarized in this section. The results refer to moisture diffusivity and thermal conductivity. Recently published values of moisture diffusivity and thermal conductivity in various foods were retrieved from the literature and were classified and analyzed statistically to reveal the influence of material moisture content and temperature. Empirical models relating moisture diffusivity and thermal conductivity to material moisture content and temperature were fitted to all examined data for each material. The data were screened carefully using residual analysis techniques. A promising model was proposed based on an Arrhenius-type effect of temperature, which uses a parallel structural model to take into account the effect of material moisture content.

4.8.1 MOISTURE DIFFUSIVITY

A total of 175 papers were retrieved from the literature from which 1773 data were obtained. These data refer to more than 100 food materials classified into 11 food categories. Among the available data only 19 materials have more than ten data, which come from more than three publications. The resulting model is summarized in Table 4.15 and the results

TABLE 4.15
Mathematical Model for Calculating Moisture Diffusivity in Foods as a Function of Moisture Content and Temperature

Proposed mathematical model

$$D = \frac{1}{1+X} D_0 \exp\left[-\frac{E_0}{R} \left(\frac{1}{T} - \frac{1}{T_r}\right)\right] + \frac{X}{1+X} D_i \exp\left[-\frac{E_i}{R} \left(\frac{1}{T} - \frac{1}{T_r}\right)\right]$$

where D is the moisture diffusivity (m^2/s), X is the material moisture content (kg/kg db), T the material temperature ($^{\circ}\text{C}$), T_r is the reference temperature (60°C), and $R = 0.0083143$ is the ideal gas constant $\text{kJ}/(\text{mol K})$

Adjustable model parameters

D_0 (m^2/s)	diffusivity at moisture $X = 0$ and temperature $T = T_r$
D_i (m^2/s)	diffusivity at moisture $X = \infty$ and temperature $T = T_r$
E_0 (kJ/mol)	activation energy for diffusion in dry material at $X = 0$
E_i (kJ/mol)	activation energy for diffusion in wet material at $X = \infty$

TABLE 4.16
Parameter Estimates of the Proposed Mathematical Model

Material	No. of Papers	No. of Data	D_i (m ² /s)	D_o (m ² /s)	E_i (kJ/mol)	E_o (kJ/mol)	s.d. (m ² /s)
<i>Cereal products</i>							
—Corn	4	26	4.40E-09	0.00E+00	0.0	10.4	1.48E-10
Dent	3	15	1.19E-08	0.00E+00	49.4	73.1	3.30E-10
Grains	3	28	1.15E-09	6.66E-11	10.2	57.8	3.17E-10
Kernel	4	25	5.87E-10	5.32E-10	0.0	33.8	1.88E-11
Pericarp	3	13	1.13E-09	0.00E+00	10.0	5.0	2.34E-11
—Pasta	3	21	1.39E-09	0.00E+00	16.2	2.0	7.71E-12
<i>Rice</i>							
Kernel	3	12	9.75E-09	0.00E+00	12.5	2.0	5.52E-11
—Rough rice	7	35	2.27E-09	0.00E+00	12.7	0.7	3.66E-11
—Wheat	6	22	1.94E-09	1.30E-09	0.0	46.3	9.53E-11
<i>Fruits</i>							
—Apple	8	39	7.97E-10	1.16E-10	16.7	56.6	1.92E-10
—Banana	4	34	2.03E-09	4.66E-10	9.9	4.6	1.77E-10
<i>Grapes</i>							
Seedless	3	32	5.35E-09	0.00E+00	34.0	10.4	1.45E-10
—Raisins	3	10	8.11E-10	1.05E-10	21.4	50.1	6.88E-11
<i>Model foods</i>							
—Amioca	4	49	1.52E-08	1.52E-08	0.0	33.3	1.02E-09
—Hylon-7	5	48	1.96E-08	1.96E-08	0.0	24.2	3.87E-09
<i>Vegetables</i>							
—Carrot	9	90	2.47E-09	1.54E-09	13.9	11.3	1.69E-09
—Garlic	4	22	5.33E-10	1.68E-11	15.4	7.1	7.43E-11
—Onion	4	31	1.45E-08	0.00E+00	70.2	10.4	1.58E-09
—Potato	16	106	1.57E-09	4.31E-10	44.7	76.9	4.02E-10

of parameter estimation are presented in Table 4.16. Figure 4.18 through Figure 4.21 present the model-calculated values for selected food materials as a function of moisture content and temperature.

The regression procedure was applied simultaneously to all the data of each material, regardless of the data sources. Thus, the results are not based on the data of only one author and, consequently, they are of higher accuracy and general applicability.

The diffusivity parameters D_o and D_i of the proposed model vary in the range of 10^{-10} to 10^{-8} m²/s. It should be noted that the self-diffusivity of water is approximately 10^{-9} m²/s, and the moisture diffusivity in bone-dry food material should be lower (in our analysis, by a factor of 100).

Low moisture diffusivities are found in nonporous and sugar-containing foods, whereas higher values of moisture diffusivity characterize porous food materials. Diffusivities higher than the self-diffusivity of water are indicative of vapor diffusion in porous solids.

The moisture diffusivity increases, in general, with increasing moisture content. Temperature has a positive effect, which depends strongly on the food material. The energy of activation for diffusion E of water is, in general, higher in the dry food materials. Some observed exceptions may be explained by the prevailing type of diffusion. Thus, lower values of activation energy for diffusion are expected for porous foods, where vapor diffusion is important. In general, temperature has a stronger effect on diffusivity in liquids and solids than in the gaseous state.

4.8.2 THERMAL CONDUCTIVITY

A total of 146 papers were retrieved from the literature from which 1210 data were obtained. These data refer to more than 100 food materials classified into 11 food categories. Among the available data only 13 materials have more than 10 data, which come from more than three publications. This procedure is applied to these data and the results of parameter

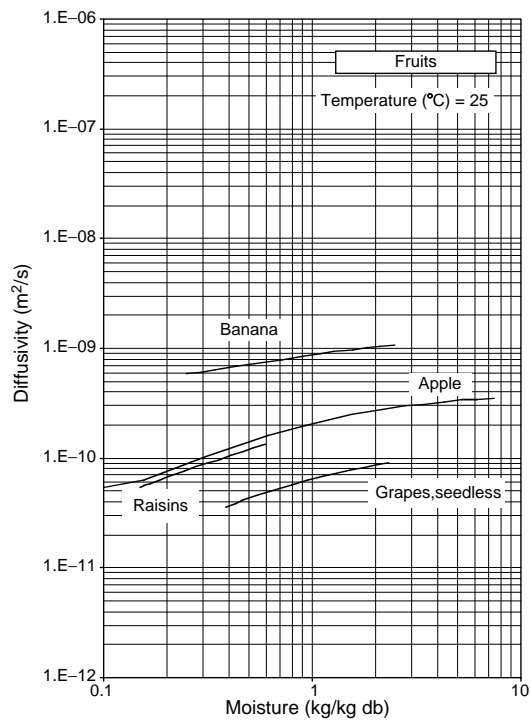


FIGURE 4.18 Predicted values of moisture diffusivity of fruits at $25^\circ C$.

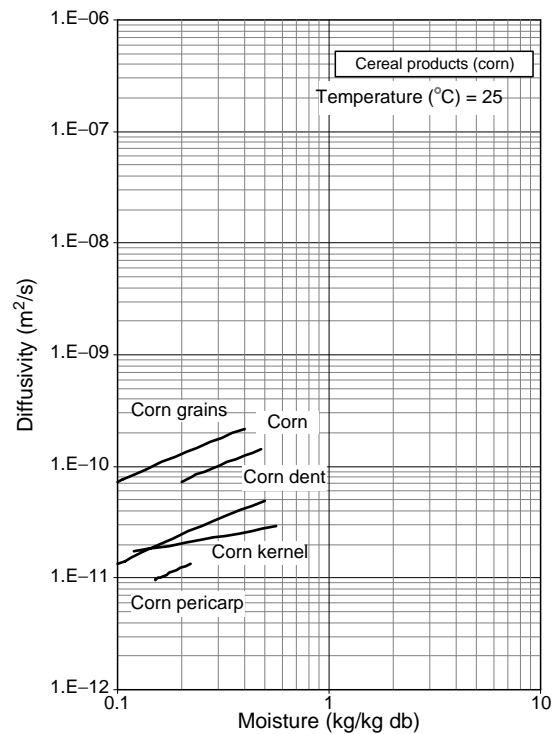


FIGURE 4.20 Predicted values of moisture diffusivity of corn at $25^\circ C$.

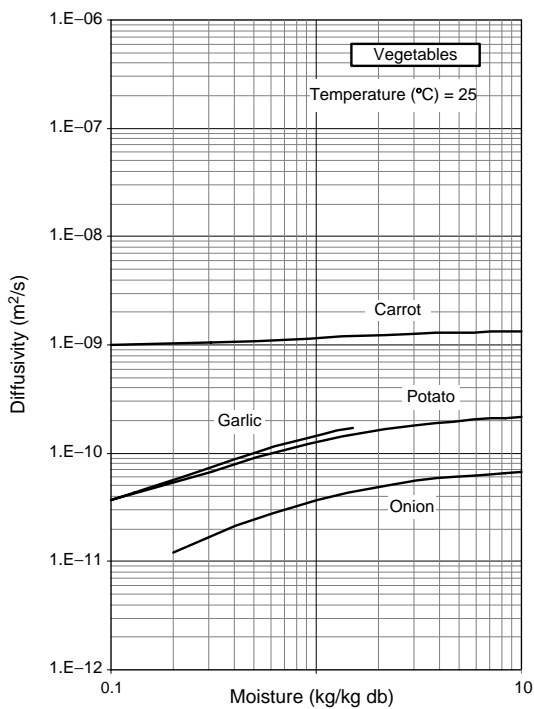


FIGURE 4.19 Predicted values of moisture diffusivity of vegetables at $25^\circ C$.

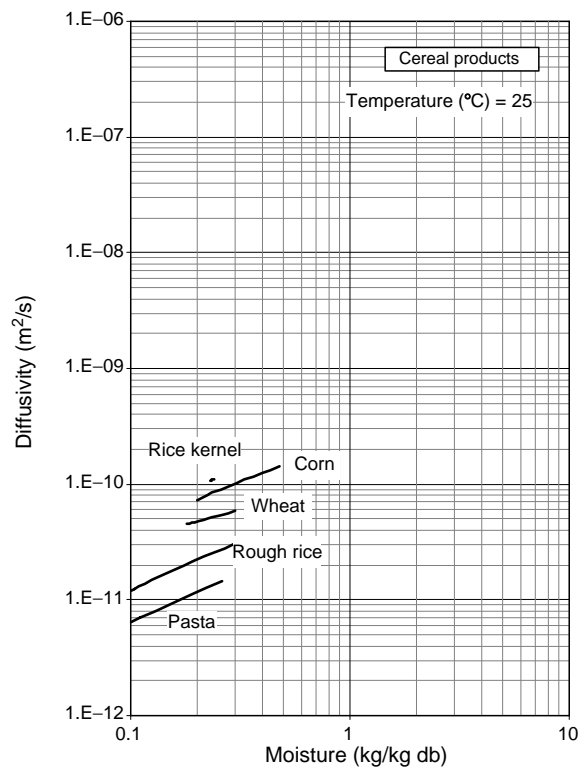


FIGURE 4.21 Predicted values of moisture diffusivity of cereal products at $25^\circ C$.

TABLE 4.17
Mathematical Model for Calculating Thermal Conductivity in Foods as a Function of Moisture Content and Temperature

Proposed mathematical model

$$\lambda = \frac{1}{1+X} \lambda_0 \exp \left[-\frac{E_0}{R} \left(\frac{1}{T} - \frac{1}{T_r} \right) \right] + \frac{X}{1+X} \lambda_i \exp \left[-\frac{E_i}{R} \left(\frac{1}{T} - \frac{1}{T_r} \right) \right]$$

where λ is the thermal conductivity (W/(m K)), X is the material moisture content (kg/kg db), T the material temperature (°C), $T_r = 60^\circ\text{C}$ a reference temperature, and $R = 0.0083143$ the ideal gas constant kJ/(mol K)

Adjustable model parameters

λ_0 (W/(m K)) thermal conductivity at moisture $X = 0$ and temperature $T = T_r$

λ_i (W/(m K)) thermal conductivity at moisture $X = \infty$ and temperature $T = T_r$

E_0 (kJ/mol) activation energy for heat conduction in dry material at $X = 0$

E_i (kJ/mol) activation energy for heat conduction in wet material at $X = \infty$

estimation are presented in Table 4.17 and Table 4.18. Figure 4.22 through Figure 4.24 present the model-calculated values for selected food materials as a function of moisture content and temperature.

Thermal conductivity parameters λ_0 and λ_i vary in the range of 0.05 to 1.0 W/(m K). It should be noted that the thermal conductivity of air is about 0.026 W/(m K), whereas that of water is 0.60 W/(m K). Values of thermal conductivity of foods higher than 0.60 W/(m K) are normally found in frozen food materials ($\lambda_{\text{ice}} = 2$ W/(m K)).

The thermal conductivity increases, in general, with increasing moisture content. Temperature has a positive effect, which depends strongly on the food material. The energy of activation for heat conduction E is, in general, higher in dry food materials.

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NOMENCLATURE

A_0	effective surface area for heat and mass transfer, m^2
A'	vibration amplitude, m
a_i	constants in equations of Table 4.3 and in Equation 4.17, various units of measure
a_w	air–water activity

TABLE 4.18
Parameter Estimates of the Proposed Mathematical Model

Material	No. of Papers	No. of Data	λ_i (W/(m K))	λ_0 (W/(m K))	E_i (kJ/mol)	E_0 (kJ/mol)	s.d. (W/(m K))
<i>Cereal products</i>							
Corn	3	15	1.580	0.070	7.2	5.0	0.047
<i>Fruits</i>							
Apple	12	68	0.589	0.287	2.4	11.7	0.114
Orange	4	13	0.642	0.106	1.3	0.0	0.007
Pear	5	15	0.658	0.270	2.4	1.9	0.016
<i>Model foods</i>							
Amioca	5	29	0.718	0.120	3.2	14.4	0.037
Starch	4	24	0.623	0.243	0.3	0.4	0.006
Hylon	3	21	0.800	0.180	9.9		0.072
<i>Vegetables</i>							
Potato	12	37	0.611	0.049	0.0	47.0	0.059
Tomato	5	28	0.680	0.220	0.2	5.0	0.047
<i>Dairy</i>							
Milk	5	33	0.665	0.212	1.7	1.9	0.005
<i>Meat</i>							
Beef	6	37	0.568	0.280	2.2	3.2	0.017
<i>Other</i>							
Rapeseed	3	35	0.239	0.088	3.6	0.6	0.023
<i>Baked products</i>							
Dough	3	15	0.800	0.273	2.7	0.0	0.183

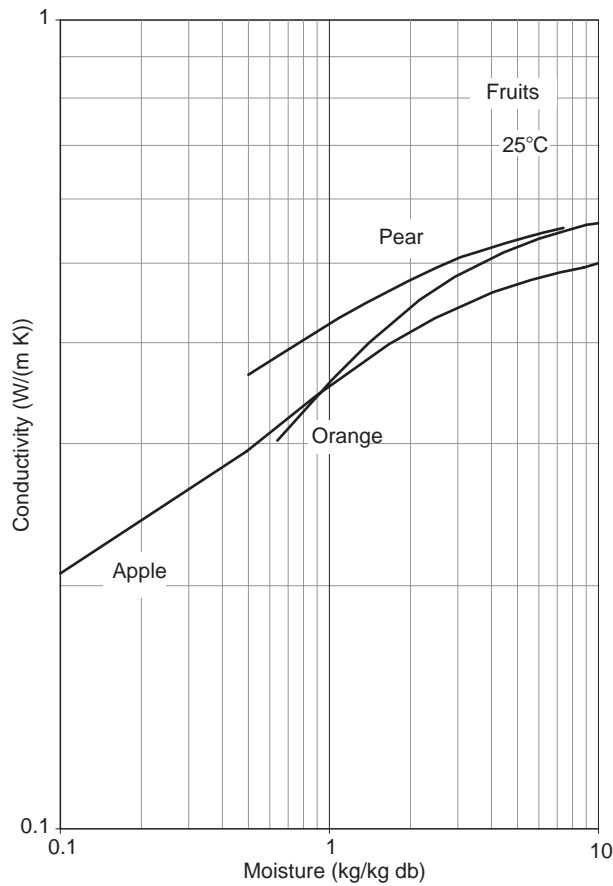


FIGURE 4.22 Predicted values of thermal conductivity of fruits at 25°C.

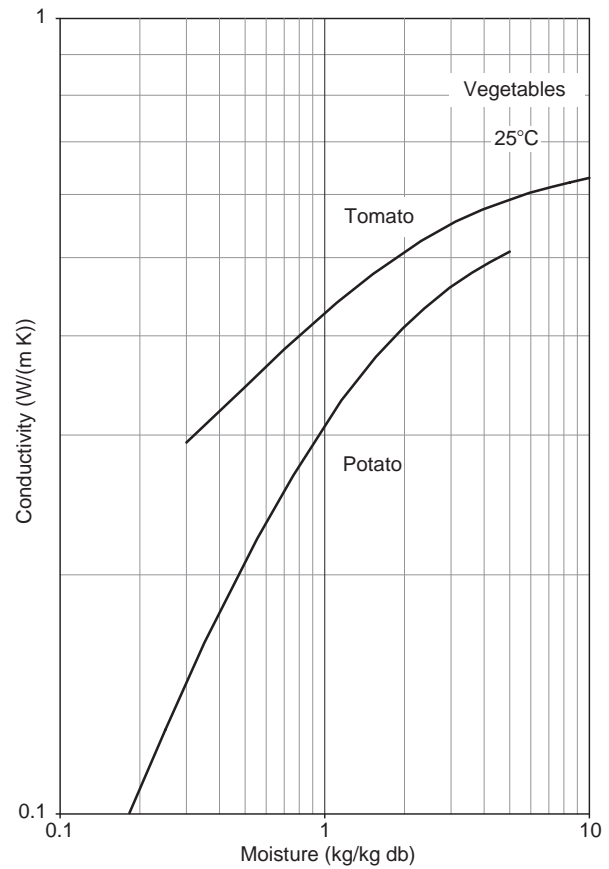


FIGURE 4.23 Predicted values of thermal conductivity of vegetables at 25°C.

a_{we} equilibrium air–water activity
 b_i constants in equations of Table 4.11 and Table 4.12 and in Equation 4.21, various units of measure
 Bi_H Biot number for heat transfer (see Table 4.10)
 Bi_M Biot number for mass transfer (see Table 4.10)
 c_i constants in Equation 4.19, various units of measure
 c_p specific heat, kJ/(kg · K)
 ΔH_s latent heat of vaporization, kJ/kg
 d particle diameter, m
 d_i constants in Equation 4.23, various units of measure
 D diffusivity in solids, m²/s
 D_A vapor diffusivity in air, m²/s
 D' diameter of spouted bed, m
 db dry base
 D_O Arrhenius factor in Equation 4.2, m²/s
 E activation energy in Arrhenius equation, kJ/kmol
 e_{Ti} relative deviation between experimental and calculated values of material temperature, °C

e_{Xi} relative deviation between experimental and calculated values of material moisture content, kg/kg db
 F constant (Table 4.7)
 f friction factor
 f' vibration frequency, 1/s
 G mass flow rate of air, kg/(m² s)
 Gu Gukhman number (see Table 4.10)
 h_H heat transfer coefficient, kW/(m² K)
 h_M mass transfer coefficient, kg/(m² s)
 H_0 static bed height for spouted beds, m
 h_s specific enthalpy, kJ/kg
 h_{vH} volumetric heat transfer coefficient, kW/(m³ K)
 h_{vM} volumetric mass transfer coefficient kg/(m³ s)
 J rate of mass transfer, kg/s
 j_H heat transfer factor (see Table 4.10)
 j_M mass transfer factor (see Table 4.10)
 K drying constant, 1/s, 1/h
 k effective thermal conductivity, kW/(m · K)
 k_A thermal conductivity of air, kW/(m · K)
 k_i thermal conductivity of phase i , kW/(m² · K)
 L slab thickness, m

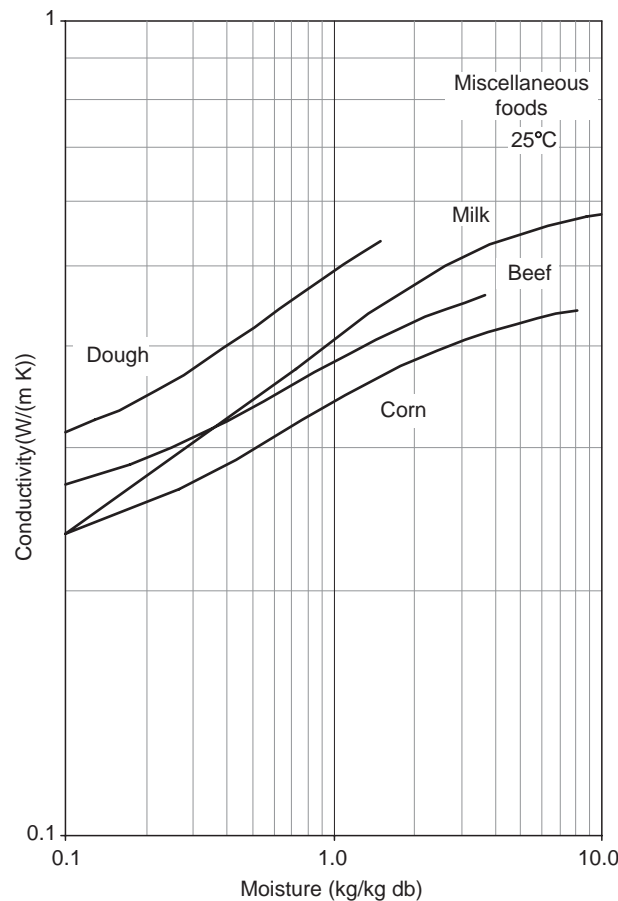


FIGURE 4.24 Predicted values of thermal conductivity of miscellaneous foods at 25°C.

N	number of measurements
Nu	Nusselt number (see Table 4.10)
Nu'	$Nu' = Nu \varepsilon / (1 - \varepsilon)$
Pr	Prandtl number (see Table 4.10)
Q	rate of heat transfer, kW/s
R	gas constant, kJ/(kmol · K)
Re	Reynolds number (see Table 4.10)
Re'	$Re' = Re (1 - \varepsilon)$
Re_s	Re based on u_s instead of u_A
Sc	Schmidt number (see Table 4.10)
Sh	Sherwood number (see Table 4.10)
St	Stanton number (see Table 4.10)
S_T	standard deviation between experimental and calculated values of material temperature, °C
S_X	standard deviation between experimental and calculated values of material moisture content, kg/kg db
t	time, s
T_A	air temperature, °C
T	material temperature, °C
T_i	experimental value of material temperature during measurement i , °C

$T_{i,c}$	calculated value of material temperature during measurement i , °C
u_A	air velocity, m/s
u_s	air velocity for incipient spouting, m/s
V	total volume of the material, m ³
w_b	wet base
X	material moisture content, kg/kg db
X_A	air humidity, kg/kg db
X_{AS}	air humidity at the solid interface, kg/kg db
X_e	equilibrium material moisture content, kg/kg db
X_i	experimental value of material moisture content during measurement i , kg/kg db
$X_{i,c}$	calculated value of material moisture content during measurement i , kg/kg db
O	Thermal conductivity at moisture $X = O$
OO	Thermal conductivity at moisture $X = OO$
Z	constant (Table 4.7)

Greek Symbols

α	specific area, m ² /m ³
ε	void fraction (porosity)
δ	constrictivity
μ	dynamic viscosity of air, kg/ms
ρ_A	density of air, kg/m ³
ρ	density of material, kg/m ³
τ	tortuosity

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5 Spreadsheet-Aided Dryer Design

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5.1 INTRODUCTION

Spreadsheet software has become an indispensable tool for engineers, because of the availability of personal computers, ease of use, and adaptability to many types of problems. Spreadsheet software has achieved great popularity because of its availability for microcomputers at reasonable cost, the ease of learning and using the software, and its flexible applications to many problems.

Furthermore, general-purpose spreadsheet software can be used effectively in process design (Maroulis and Saravacos, 2003). For example, Microsoft Excel with Visual Basic for Applications is an effective tool for process design. Spreadsheets offer sufficient process model “hospitality.” They are connected easily and online with charts and graphic objects, resulting in powerful and easy-to-use graphical interfaces. Excel also supports mathematical and statistical tools. For instance, Solver is an excellent tool for solving sets of equations and performing optimization. Databases are effectively and easily accessed. In addition, Visual Basic for Applications offers a powerful object-oriented programming language, capable of constructing commercial graphics interfaces.

It is the objective of this chapter to present step-by-step procedures in order to allow application of various dryer models into the Excel environment. This chapter refers to two main topics. The principles for solution of a process design problem are presented first and then the principles for Excel implementation are described.

The reader needs to become familiar with the following topics regarding Excel software, using the related literature:

- Modeling and spreadsheets
- Analyzing the Solver
- Sensitivity analysis using Excel tables
- Controls and dialog boxes to input data
- Graphics to get the results
- Databases
- Visual Basic as a programming language

5.2 PRINCIPLES AND TECHNIQUES OF SPREADSHEET-AIDED PROCESS DESIGN

Computer-aided design is based on computer simulators, whereas computer simulators are based on process modeling. The basic terms, such as modeling, simulation, and design, are defined in [Table 5.1](#). Modeling is the procedure of translating the physical laws of a process to mathematical equations to analyze or design the process. Simulation is the appropriate software, which predicts the real performance of a process. It is based on mathematical modeling plus the appropriate graphics interface in a computer environment. Design is a procedure of sizing and rating a process in order to achieve specific goals, such as economic production, product quality, and protection of the environment.

Modeling and simulation are useful tools in process design. [Table 5.2](#) summarizes a step-by-step procedure for process modeling, whereas [Table 5.3](#)

TABLE 5.1
Basic Definitions

Modeling: is the procedure to translate the physical laws of a process to mathematical equations

Simulation: is the appropriate software which guesses the real performance of a process

Design: is a procedure to size and rate a process in order to obtain a specific goal

Sizing: given the process specifications calculate the equipment size and characteristics

Rating: given the process specifications and the equipment size and characteristics calculate the operating conditions

summarizes a step-by-step procedure for process simulation. These steps are further analyzed.

The equations constituting a model describe the physical laws, which apply to the process. They are derived from material and energy balances, thermodynamic equilibrium relationships, transport phenomena, geometry, equipment characteristics, etc. Generally some assumptions are also built into the model.

A degrees-of-freedom analysis is shown in Table 5.4 and in Figure 5.1. Suppose that M variables are incorporated into the mathematical model of N equations; generally, M is greater than or equal to N , and the difference $M-N$ corresponds to the degrees of freedom of the process. The degrees of freedom is characteristic of the process. In process design, some variables have given values, due to the design specifications, and the remainder corresponds to design variables.

The number of design variables is characteristic of the problem. Several different problems could be defined for every process (see, for example, Table 5.5). The values for the design variables are decided by the design engineer. The remainder $N \times N$ set of equations is solved by using mathematical techniques. In chemical and food engineering the resulting system is sparse, that is every variable appears in a few equations. In that case the system can be solved

TABLE 5.2
Process Modeling

1. Process model formulation
 2. Degrees-of-freedom analysis
 3. Alternative problem formulations
 4. Problem-solution algorithm
 5. Cost estimation and project evaluation analysis
 6. Process optimization
-

TABLE 5.3
Process Simulation Procedure in a Spreadsheet Environment

1. Model development in a spreadsheet
 2. Implementation of alternative problem solutions or optimization procedures
 3. Development of graphics interface
-

sequentially (down triangle matrix) or by using a few trial variables.

The above approach is suitable for implementation in a spreadsheet environment. The resulting simulator has generally the outline presented in Figure 5.2. Four different units are distinguished, with each one developed in a different sheet (Maroulis and Saravacos, 2002).

The “Process Model Worksheet” is the heart of the system calculations. It contains the process model. When no interactions are needed, the model solution uses only worksheet functions. In that case, when any change in input variables (free variables) occurs, the solution is obtained automatically on this worksheet.

Since the use of the simulator requires the solution of different problems, several different problems are formulated in the “Problem Solution Visual Basic Module.” Their solution is based on the simplest problem of the process model worksheet above, and uses the Solver or the Goal Seek utilities of Excel via a Visual Basic program, to obtain a solution for the alternative problems.

All technical and required data are retrieved from the “Database worksheet,” which contains all the required information in the form of “data lists.” These data are extended and modified via appropriate dialog boxes.

“Graphics interface worksheet” is a user-friendly way for human-machine communication. It usually consists of three parts: (a) Problem specifications: The specifications and the required data for the problem

TABLE 5.4
Degrees-of-Freedom Analysis

Total number of variables	M	
Total number of equations	N	
Degrees of freedom	$F = M - N$	Process characteristic
Degrees of freedom	F	
Problem specifications	K	
Design variables	$D = F - K$	Problem characteristic

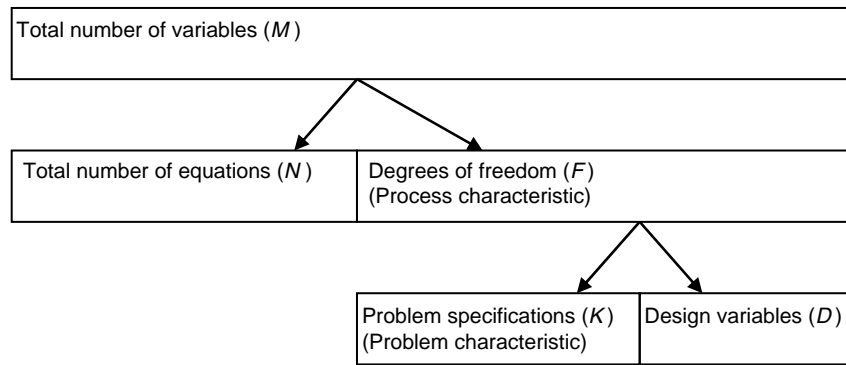


FIGURE 5.1 Degrees-of-freedom analysis.

to be solved are entered by the user or estimated from the databases. Data are inserted via dialog boxes or buttons for changing some important magnitudes. (b) Problem-type selection: The type of problem to be solved is selected via buttons. (c) Results presentation: The results are obtained automatically, and are presented in the form of tables or charts. Since these charts are updated automatically, the user has at his disposal all the information needed for sizing, rating, sensitivity analysis, or comparison of alternative solutions.

The following steps comprise an Excel implementation procedure:

1. Workbook preparation
2. Process modeling in a spreadsheet
3. Using “Solver” for process optimization
4. Using graphs and tables for presentation of the results
5. Introducing dialog boxes and controls to modify data
6. Toward an integrated graphics interface

Step 1: Workbook Preparation

Create a new workbook and name it to describe the process, e.g., “BeltDryer.xls.” Insert and name blank sheets are presented in [Table 5.6](#).

Step 2: Process Modeling in a Spreadsheet

Into the spreadsheet “Process” consider seven separate ranges, as it is presented in [Table 5.7](#).

Each range consists of three columns and several rows, one row for every variable in the range. In each range the first column contains the variable names, the second the variable values or variable formulas, and the third the units used. Name all cells in second columns according to the names in the first column. (You can use the “Ctrl+Shift+F3” option.)

TABLE 5.5
Some Typical Problems

<i>Direct</i>	
Given	the characteristics of input streams the equipment characteristics the operating conditions
Calculate	the characteristics of the output streams
<i>Design</i>	
Given	the characteristics of input streams the characteristics of output streams the equipment characteristics the operating conditions
Calculate	
<i>Rating</i>	
Given	the characteristics of input streams the characteristics of output streams the equipment characteristics the operating conditions
Calculate	
<i>Identification</i>	
Given	the characteristics of input streams the characteristics of output streams the operating conditions
Calculate	the equipment characteristics

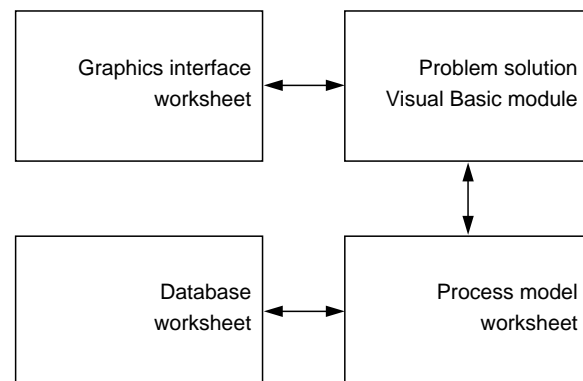


FIGURE 5.2 Simulator architecture on a spreadsheet environment.

TABLE 5.6
Sheets in “BeltDryer.xls” Workbook

Sheet Name	Purpose
<i>Spreadsheets</i>	
Process	Process model
Flow sheet	Process flow sheet
Report	Summary report of results
Control	Graphics interface
<i>Visual Basic Modules</i>	
Optimize	Process optimization subroutines
Controls	Subroutines for dialog boxes and controls
<i>Dialog box sheets</i>	
Spec	Process specifications
Tech	Technical data
Cost	Economical data

The ranges “Technical Data,” “Design Variables,” “Process Specifications,” and “Economic Data” contain only data. The ranges “Process

Model,” “Process Constraints,” and “Economic Model” contain formulas.

Having inserted data and formulas, the process model implementation has been completed. The resulting spreadsheet “Process” looks like that presented in Figure 5.3. The cell ranges can be colored with different colors. The drawn arrows show the information flow in the spreadsheet.

The spreadsheet process model is now ready for use. Any changes in process data, economic data, process specifications, design variables are taken into account and the results are updated immediately.

Any optimization technique, graphical or tabulated reports, any scenario analysis, or sensitivity analysis, any sophisticated graphics interface can be based on the “Process” spreadsheet. Some examples follow.

Step 3: Using “Solver” for Process Optimization

Create a Visual Basic subroutine with the name “optimum” in the “Optimize” module. The appropriate code is shown in Table 5.8.

TABLE 5.7
Cell Content in “Process” Spreadsheet

Range Name	Content
Technical data	Data
Design variables	Data
Process specifications	Data
Economic data	Data
Process model	Formulas
Process constraints	Formulas
Economic model	Formulas

TABLE 5.8
Visual Basic Subroutine for Process Optimization

```

Sub optimum()
1 Sheets("Process").Activate
2 SolverReset
3 SolverOk SetCell:= Range("objective"), MaxMinVal:= 1,
  ByChange:= Range("variables")
4 SolverAdd CellRef:= Range("constraints"), Relation:= 3,
  FormulaText:= 0#
5 SolverSolve UserFinish:= True
6 Beep
End Sub

```

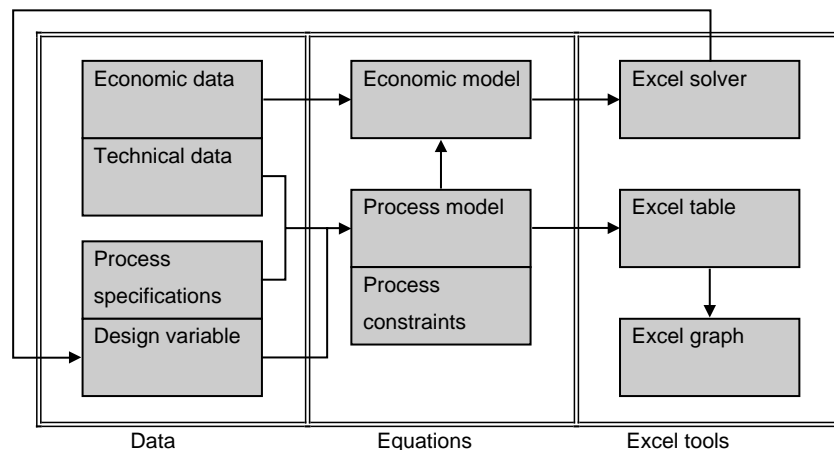


FIGURE 5.3 Model implementation in the “Process” spreadsheet.

Statement 1 activates the “Process” spreadsheet. Statement 2 resets the Solver. Statement 3 selects the cell with the name “objective” to be the objective function [SetCell:=Range(“objective”)], requires the minimization of the objective function [MaxMinVal:=2], and selects the range “variables” to be the decision variables [ByChange:=Range(“variables”)]. Statement 4 suggests that all cells in the range “constraints” [CellRef:=Range(“constraints”)] must be greater than [Relation:=3] zero [FormulaText:=0#]. Statement 5 activates the solver to find the optimum.

The above-mentioned cell names must be defined. Thus, in the sheet “Process” name:

- The cells that contain the values of the design variables as “variables”
- The cells that contain the process constraints as “constraints”
- The cell that contains the profit as “objective”

In the sheet “Process” insert a new button, name it “optimizer” and assign it to the subroutine “optimum.”

Press the button “optimizer” and the optimum is reached in a few seconds.

Step 4: Using Excel Tables and Charts for Presentation of the Results

The process design results can be further analyzed using the tools “Tables” and “Charts” supported by Excel.

For example, a process flow sheet can easily be constructed in Excel as follows: in the sheet “Flow sheet” draw a flow sheet by using the drawing toolbar. Any information concerning process conditions can be inserted in cells near the desired point of the flow sheet. For each piece of information there need to be three cells, one for the variable name, one for the variable value, and one for the variable units. That is, to insert a stream flow rate, select a cell near the icon of the stream arrow and insert the symbolic name of the stream flow rate, i.e., “F=,” in a neighboring cell insert the formula “=F” to get the value from the “Process” sheet, and in another cell, nearby, insert the units, i.e., “kg/s.” You can add any information you like. Any changes in data are updated immediately.

In order to plot the effect of the design variable (*X*) on a technical (*Y*) and an economic (*Z*) variable the following steps can be used: construct a one-dimensional Excel table in which the “Column Input Cell” is the cell with the name “X.” The second and third output columns refer to the cells “Y” and “Z,” respectively. Next construct a “XY(Scatter)” chart in

which the first column of the table corresponds to *x*-values and the second to *y*-values. Similarly, construct a second “XY(Scatter)” chart in which the first column of the table corresponds to *x*-values and the third to *y*-values.

Any other tabulated results or desired reports can be easily obtained as follows: select a spreadsheet to incorporate the required information. Insert text or graphics as you like. Get the information from the “Process” sheet, as described previously in the flow sheet construction procedure.

Step 5: Introducing Dialog Boxes and Controls to Modify Data

A dialog box can be used to modify the values of process specifications, which are included in the range “Process Specifications” in the spreadsheet “Process.”

In the Dialog Module “db_spec” insert for every variable one “Label” (from the toolbar “forms”) for its description, one “Edit Box” (from the toolbar “forms”) for its value, and one “Label” for its units. Name all the Edit Boxes with the name of the corresponding variable.

In the Visual Basic Module “vb_controls” type a subroutine to use the dialog box in the sheet “d_spec,” as described in Table 5.9.

In the spreadsheet “Process” insert a button, name it “specifications,” and assign it to the subroutine “DialogSpecifications.”

Press the button “specifications” and a dialog box appears in order to modify data for process specifications.

A scroll bar can be used for each design variable in order to modify the values of the design variables, which are included in the range “Design Variables” in the spreadsheet “Process.”

TABLE 5.9
A Subroutine to Activate the Dialog Box

```
Sub DialogSpecifications( )
dbName = "d_spec"
DialogSheets(dbName).EditBoxes("W").Text = Range("W").Value
DialogSheets(dbName).EditBoxes("Xo").Text = Range("Xo").Value
DialogSheets(dbName).EditBoxes("Yo").Text = Range("Yo").Value
If DialogSheets(dbName).Show Then
Range("W").Value = DialogSheets(dbName).EditBoxes("W").Text
Range("Xo").Value = DialogSheets(dbName).EditBoxes("Xo").Text
Range("Yo").Value = DialogSheets(dbName).EditBoxes("Yo").Text
End If
Beep
End Sub
```

A scroll bar, in order to handle the variable X , can be inserted as follows:

- Insert the scroll bar icon from the toolbar “forms”
- Insert the minimum allowable value in a cell named “X.min”
- Insert the maximum allowable value in a cell named “X.max”
- Insert the coded value in a cell named “X.CV”

The coded value ranges between 0 and 100 and is defined as follows:

$$X.CV = (X - X.min) / (X.max - X.min) * 100$$

- Insert a scroll bar from the toolbar “forms” and assign the “Cell Link” (in the “Format Object” menu) to the coded value “X.CV”
- Replace the content of the cell named “X” with the following formula:

$$= X.min + X.CV * (X.max - X.min) / 100$$

It must be noted that the range “variables” which is handled by the solver during optimization must be redefined to refer to coded values, instead of the actual values. This modification guarantees the proper performance of the optimization and of scroll bars.

Step 6: Toward an Integrated Graphics Interface

Any desired graphics interface can be developed in the spreadsheet “Control.” It can be constructed as follows:

- Draw a process flow sheet in sheet “Controls,” as described in Step 5
- Insert buttons to appear and disappear the crucial graphs
- Insert buttons to activate the desired dialog boxes
- Insert scroll bars to modify the desired process variables
- Insert buttons to solve different problems, e.g., process optimization.

The user has now at his disposal a process simulator. He can enter data via scroll bars or dialog boxes and observe the results via buttons, which activate the desired graphs or reports.

The graphics interface could be further improved to look professional using appropriate programming code in Visual Basic.

5.3 DESIGN OF A CONVEYOR BELT DRYER

In this section a design approach is described for a conveyor belt dryer (Maroulis and Saravacos, 2003).

5.3.1 PROCESS DESCRIPTION

A typical flow sheet of a conveyor belt dryer is presented in Figure 5.4. The wet feed at flow rate F (kg/s db), temperature T_0 (°C), and humidity X_0 (kg/kg db) is distributed on the belt as it enters the dryer. The dried product exits the dryer at the same flow rate on dry basis F (kg/s db), temperature T (°C), and moisture content X (kg/kg db). The belt is moving at a velocity u (m/s) and requires an electrical power E_b (kW). The drying air enters the dryer at a flow rate F_f (kg/s db), temperature T (°C), and humidity Y (kg/kg db). The drying air temperature is controlled in the heater, and the drying air humidity is controlled through the flow rate of the fresh air F_a (kg/s db). An electrical power E_f (kW) is expended by the fan and a thermal power Q (kW) is expended by the heater. The air conditions for design can be considered constant due to the high air recirculation.

5.3.2 PROCESS MODEL

A mathematical model of the process presented in Figure 5.4 is summarized in Table 5.10.

Equation T10.1 calculates the vapor pressure at drying temperature, whereas Equation T10.2 is the psychrometric equation. Equation T10.1 and Equation T10.2 are used to calculate the water activity at drying conditions (i.e., temperature T and air humidity Y). Equation T10.3 calculates the equilibrium material moisture content at drying conditions, whereas Equation T10.4 estimates the drying time constant at drying conditions. Both Equation T10.3 and Equation T10.4 are used in Equation T10.5, which calculates the required drying time.

Equation T10.6 and Equation T10.7 constitute the moisture balance at the dryer. Equation T10.6 refers to solid, and Equation T10.7 to air. The thermal energy requirements for drying are summarized

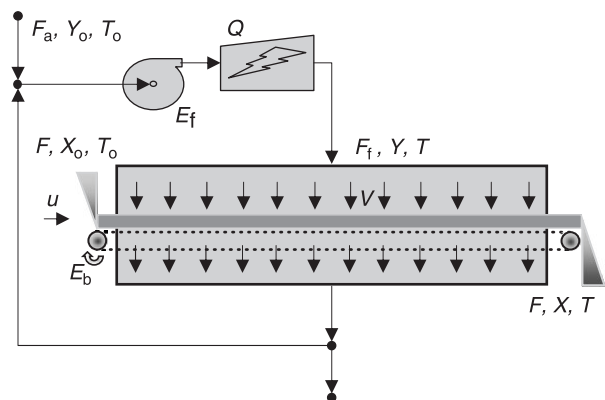


FIGURE 5.4 Schematic representation of a belt dryer.

TABLE 5.10
Belt Dryer Model

Psychrometric equations

$$P_s = \exp [a_1 - a_2/(a_3 + T)] \quad (\text{T10.1})$$

$$Y = ma_w P_s / (P - a_w P_s) \quad (\text{T10.2})$$

Drying kinetics

$$X_c = b_1 \exp[b_2/(273+T)] [a_w/(1-a_w)]^{b_3} \quad (\text{T10.3})$$

$$t_c = c_0 d^{c_1} V^{c_2} T^{c_3} Y^{c_4} \quad (\text{T10.4})$$

$$t = -t_c \ln[(X - X_c)/(X_0 - X_c)] \quad (\text{T10.5})$$

Material balance

$$W = F(X_0 - X) \quad (\text{T10.6})$$

$$W = F_a(Y - Y_0) \quad (\text{T10.7})$$

Thermal energy requirements

$$Q_{we} = F(X_0 - X) [\Delta H_0 - (C_{PL} - C_{PV})T] \quad (\text{T10.8})$$

$$Q_{sh} = F[C_{PS} + X_0 C_{PL}] (T - T_0) \quad (\text{T10.9})$$

$$Q_{ah} = F_a [C_{PA} + Y_0 C_{PV}] (T - T_0) \quad (\text{T10.10})$$

$$Q = Q_{we} + Q_{sh} + Q_{ah} \quad (\text{T10.11})$$

Air heater

$$Q = A_s U_s (T_s - T) \quad (\text{T10.12})$$

Belt dryer

$$M = tF(1 + X_0) \quad (\text{T10.13})$$

$$M = (1 - \varepsilon) \rho_s H \quad (\text{T10.14})$$

$$H = Z_0 DL \quad (\text{T10.15})$$

$$A_b = LD \quad (\text{T10.16})$$

$$u_b = L/t \quad (\text{T10.17})$$

Fan

$$\Delta P = f_1 Z_0 V^2 \quad (\text{T10.18})$$

$$F_i = \rho_a VDL \quad (\text{T10.19})$$

$$E_f = \Delta P F_i / \rho_a \quad (\text{T10.20})$$

Belt driver

$$E_b = e_1 L(1 + X_0)F \quad (\text{T10.21})$$

Electrical energy requirements

$$E = E_b + E_f \quad (\text{T10.22})$$

Performance indices

$$n = Q_{we}/Q \quad (\text{T10.23})$$

$$r = W/A_b \quad (\text{T10.24})$$

in Equation T10.8 through Equation T10.11. Equation T10.8 refers to water evaporation, Equation T10.9 to solids heating, Equation T10.10 to rejected air heating, and Equation T10.11 refers to the total energy required by the heater.

Equation T10.12 is used for sizing the heater. Equation T10.13 through Equation T10.17 are used for sizing the belt.

Equation T10.13 correlates the residence time with the mass holdup, and Equation T10.14 the mass holdup with the volume holdup. These equations are valid for all dryer types. Equation T10.15 is the geometrical distribution of the volume holdup on the belt. Equation T10.16 calculates the required belt area, and Equation T10.17 the required belt velocity to obtain the desired residence time.

Equation T10.18 through Equation T10.20 are used for sizing the fan. Equation T10.18 calculates the pressure loss of air through the loaded belt. Equation T10.19 correlates the airflow with the air velocity. Equation T10.20 estimates the required electrical power to operate the fan.

Equation T10.21 estimates the required electrical power to move the belt. Equation T10.22 calculates the required total electrical power.

Finally, Equation T10.23 and Equation T10.24 define two crucial dryer performance indices. Equation T10.23 defines the dryer thermal performance, whereas Equation T10.24 calculates the evaporating capacity per unit belt area.

Thirty-seven variables presented in Table 5.11 are involved in the model of 24 equations presented in Table 5.10. The corresponding technical data are summarized in Table 5.12. The process specifications of a typical design problem are presented in Table 5.13, whereas a degrees-of-freedom analysis is shown in Table 5.14, which results in four design variables. Table 5.15 suggests a selection of design variables and the corresponding solution algorithm is presented in Table 5.16. The total annualized cost (TAC) presented in Table 5.17 is used as objective function in process optimization. The required cost data are summarized in Table 5.18.

5.4 EXCEL IMPLEMENTATION OF A BELT DRYER DESIGN

In this section the dryer design model presented in Section 5.3 is implemented in an Excel environment according to the principles and techniques presented in Section 5.2.

Steps 1–3 of Section 5.2 are applied and the dryer model is created on the spreadsheet “process” as shown in Figure 5.5. The ranges “Technical Data,” “Process Specifications,” “Design Variables,” and “Cost Data” contain data according to Table 5.12, Table 5.13, Table 5.15, and Table 5.18, respectively. The range “Model Solution” contains the solution of the model in Table 5.10 according to the solution presented in Table 5.16, and the range “Cost Analysis” represents the analysis presented in Table 5.17. Finally, the button “optimize” performs an optimization, i.e., it finds the (optimal) values of the design variables (Y , T , V , D), which minimize the objective function (TAC). Figure 5.5 constitutes a simple but accurate belt dryer design simulator. Different problems (different material, financial environment, process specifications) can be solved instantaneously.

Step 4 of Section 5.2 is applied, as an example, (a) to construct a dynamic process flow sheet (Figure 5.6);

TABLE 5.11
Process Variables

<i>Drying air</i>		
F_a	ton/h	Fresh airflow rate
F_f	ton/h	Recycle airflow rate
T	°C	Drying air temperature
Y	kg/kg db	Drying air humidity
V	m/s	Drying air velocity
P	bar	Drying pressure
T_0	°C	Ambient temperature
Y_0	kg/kg db	Ambient humidity
P_s	bar	Vapor pressure at drying conditions
a_w	—	Water activity at drying conditions
<i>Material</i>		
F	ton/h	Material flow rate
X_0	kg/kg db	Initial moisture content
X	kg/kg db	Final moisture content
X_e	kg/kg db	Equilibrium moisture content at drying conditions
d	m	Particle size
t_c	h	Drying time constant at drying conditions
t	h	Drying time
<i>Dryer</i>		
W	ton/h	Drying rate
L	m	Dryer length
D	m	Dryer width
M	ton	Dryer mass holdup
H	m ³	Dryer volume holdup
A_b	m ²	Belt area
A_s	m ²	Air heater transfer area
u_b	m/s	Belt velocity
Z_0	m	Loading depth
ΔP	bar	Pressure loss of air flowing through belt
<i>Thermal load</i>		
Q_{we}	kW	Water vaporization
Q_{sh}	kW	Solid heating
Q_{ah}	kW	Air heating
Q	kW	Total thermal load
T_s	°C	Steam temperature
<i>Electrical load</i>		
E_b	kW	Belt driver
E_f	kW	Fan
E	kW	Total power requirement
<i>Performance</i>		
n	—	Thermal efficiency
r	kg/h m ²	Specific rate of evaporation

TABLE 5.12
Technical Data

<i>Density (kg/m³)</i>	
ρ_w	Water
ρ_a	Air
ρ_s	Dry material
<i>Specific heat (kJ/kg K)</i>	
C_{PL}	Water
C_{PV}	Water vapor
C_{PA}	Air
C_{PS}	Dry material
<i>Latent heat (kJ/kg)</i>	
ΔH_0	Steam condensation at 0°C
<i>Other</i>	
U_s	Heat transfer coefficient at air heater (kW/m ² K)
ε	Void (empty) fraction of loading
<i>Empirical constants</i>	
a_1, a_2, a_3	Antoine equation for vapor pressure of water
b_1, b_2, b_3	Oswin equation for material isotherms
c_0, c_1, c_2, c_3, c_4	Drying kinetics equation
e_1	Belt driver power equation
f_1	Pressure loss equation

TABLE 5.13
Process Specifications

F	ton/h db	Feed flow rate
X_0	kg/kg db	Initial material moisture content
X	kg/kg db	Final material moisture content
d	m	Material characteristic size
T_0	°C	Ambient temperature
Y_0	kg/kg db	Ambient humidity
Z_0	m	Loading depth
P	bar	Ambient pressure
T_s	°C	Heating steam temperature

TABLE 5.14
Degrees-of-Freedom Analysis

Process variables	37	Degrees of freedom	13
Process equations	24	Specifications	9
Degrees of freedom	13	Design variables	4

TABLE 5.15
Design Variables

Y	kg/kg db	Drying air humidity
T	°C	Drying air temperature
V	m/s	Drying air velocity
D	m	Belt width

(b) to investigate the effect of one design variable on an economic variable (Figure 5.7); (c) to analyze the effect of two design variables on a technical variable (Figure 5.8); (d) to summarize the results of the design on a synoptic report (Figure 5.9). Any other analysis

TABLE 5.16
Model Solution Algorithm

Equation		
T10.1	→	P_s
T10.2	→	a_w
T10.3	→	X_c
T10.4	→	t_c
T10.5	→	t
T10.6	→	W
T10.7	→	F_a
T10.8	→	Q_{we}
T10.9	→	Q_{sh}
T10.10	→	Q_{ah}
T10.11	→	Q
T10.12	→	A_s
T10.13	→	M
T10.14	→	H
T10.15	→	L
T10.16	→	A_b
T10.17	→	u_b
T10.18	→	ΔP
T10.19	→	F_f
T10.20	→	E_f
T10.21	→	E_b
T10.22	→	E
T10.23	→	n
T10.24	→	r

can be performed in a similar way according to the scope of the designer.

Step 5 of [Section 5.2](#) is also applied exemplarily to insert a “scroll bar” for each of the design variables and a “dialog box” to modify the process specifications. The results are shown in [Figure 5.10](#). The resulting graphics interface of [Figure 5.10](#) could be further improved by introducing more tools, tables, and graphs. It can also become more professional using appropriate programming in Visual Basic.

TABLE 5.17
Cost Analysis

Equipment cost	
$C_{eq} = C_{bel}A^{n_{bel}} + C_{exc}A_s^{n_{exc}} + C_{fan}E_f^{n_{fan}}$	(T17.1)
Annual operating cost	
$C_{op} = (C_sQ + C_eE)t_y$	(T17.2)
Total annual cost (objective function)	
$TAC = eC_{eq} + C_{op}$	(T17.3)
where the Capital Recovery Factor is calculated from the equation	
$e = \frac{i_r(1 + i_r)^L}{(1 + i_r)^L - 1}$	(T17.4)

TABLE 5.18
Cost Data

<i>Utility cost</i>		
C_e	\$/kW h	Cost of electricity
C_s	\$/kW h	Cost of heating steam
<i>Equipment unit cost</i>		
C_{bel}	\$/m ²	Belt dryer
C_{exc}	\$/m ²	Heat exchanger
C_{fan}	\$/kW	Fan
<i>Equipment size scaling factor</i>		
n_{bel}	—	Belt dryer
n_{exc}	—	Heat exchanger
n_{fan}	—	Fan
<i>Other</i>		
t_y	h/yr	Annual operating time
i_r	—	Interest rate
L	yr	Lifetime

NOMENCLATURE

a_i	Antoine equation constants
A_b	belt area, m ²
A_s	air heater transfer area, m ²
a_w	water activity
b_i	Oswin equation constants
c_i	drying kinetics equation constants
C_{bel}	belt dryer unit cost, \$/m ²
C_e	cost of electricity, \$/kW h
C_{exc}	heat exchanger unit cost, \$/m ²
C_{fan}	fan unit cost, \$/kW
C_{PA}	specific heat of air, kJ/kg K
C_{PL}	specific heat of water, kJ/kg K
C_{PS}	specific heat of dry material, kJ/kg K
C_{PV}	specific heat of water vapor, kJ/kg K
C_{PW}	specific heat of liquid water, kJ/kg K
C_s	cost of heating steam, \$/kW h
D	dryer width, m
d	particle size, m
db	dry basis
E	total power requirement, kW
e	capital recovery factor
e_1	belt driver power equation constant
E_b	belt driver power, kW
E_f	fan power, kW
E_r	rotating driver power, kW
F	material flow rate, ton/h db
f_1	pressure loss equation power
F_a	fresh airflow rate, ton/h
F_f	recycle airflow rate, ton/h
H	dryer volume holdup, m ³
i_r	interest rate
L	dryer length, m

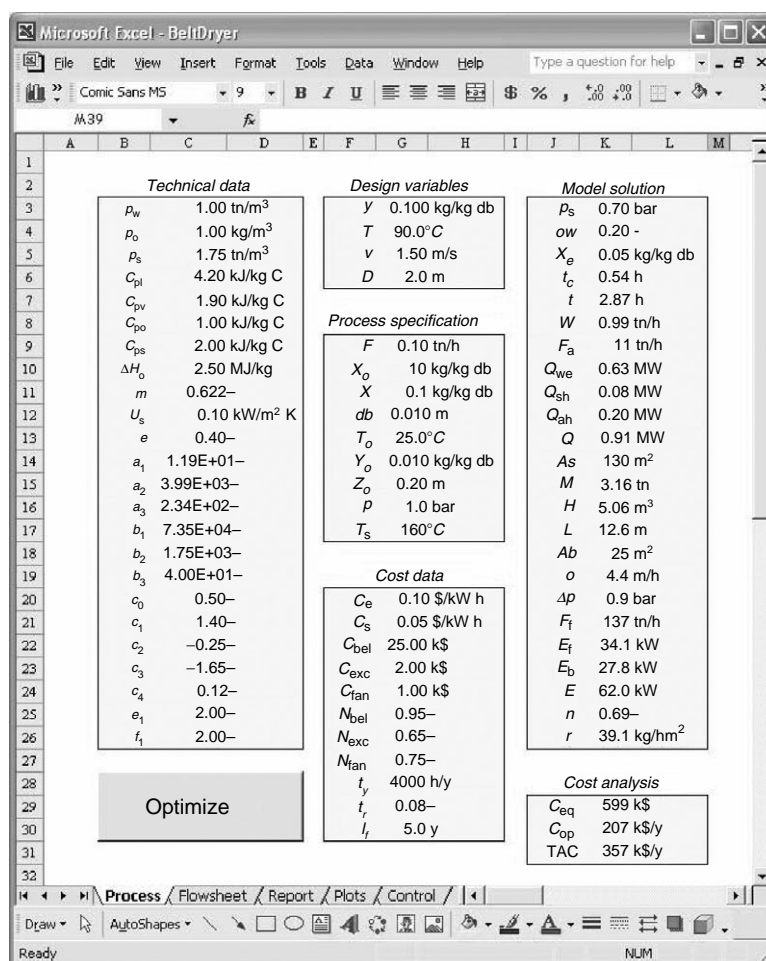


FIGURE 5.5 Dryer model implementation in the “Process” spreadsheet.

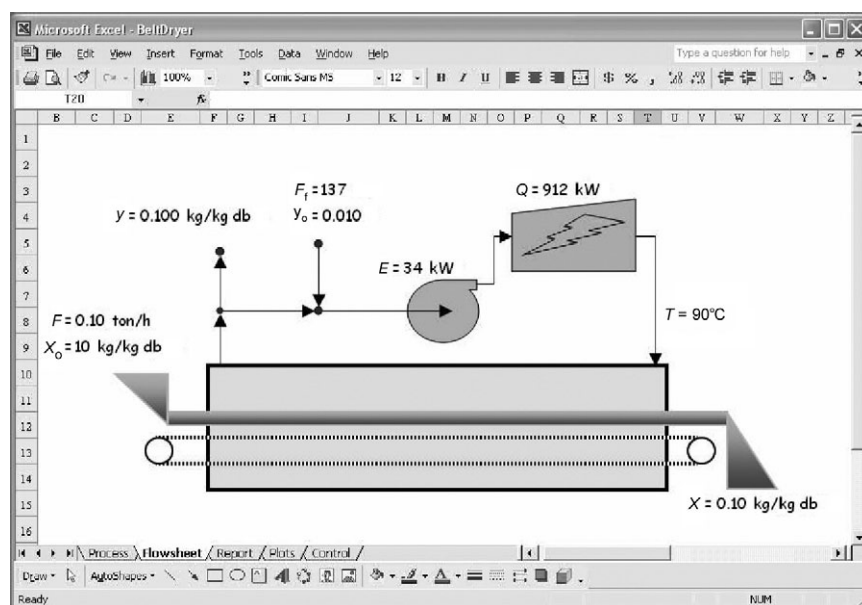


FIGURE 5.6 Process flow sheet implemented in the spreadsheet “Flow sheet.”

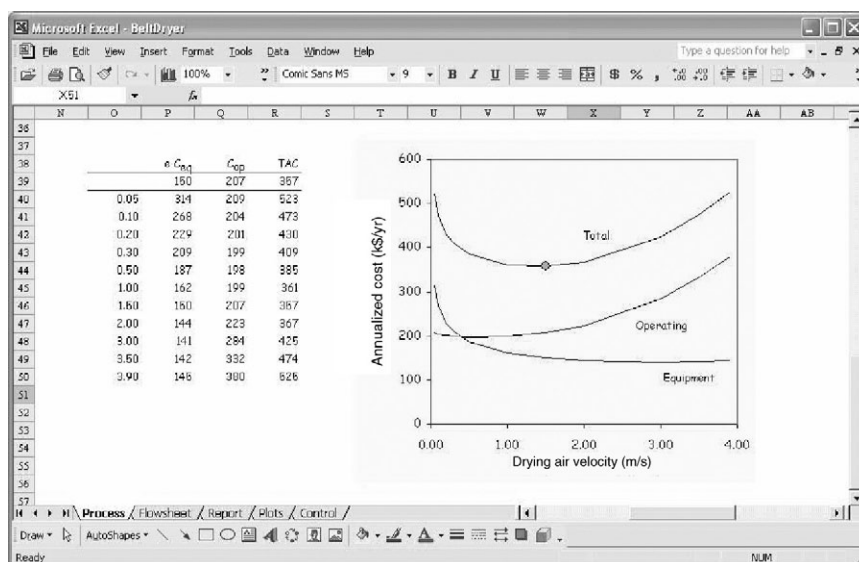


FIGURE 5.7 Analyze the effect of one design variable on some economic variables, using the “One-Dimensional Table” and “Chart” tools, supported by Excel.

l_f	lifetime, yr	Q	total thermal load, kW
m	air–water molecular weight ratio	Q_{ah}	air-heating thermal load, kW
M	dryer mass holdup, ton	Q_{sh}	solid-heating thermal load, kW
n	thermal efficiency	Q_{we}	water vaporization thermal load, kW
n_{bel}	belt dryer scaling factor	r	specific rate of evaporation, kg/h m ²
n_{exc}	heat exchanger	t	drying time, h
n_f	number of flights	T	drying air temperature, °C
n_{fan}	fan scaling factor	t_c	drying time constant, h
P	pressure, bar	T_0	ambient temperature, °C
P_s	vapor pressure at temperature T , bar	T_s	steam temperature, °C

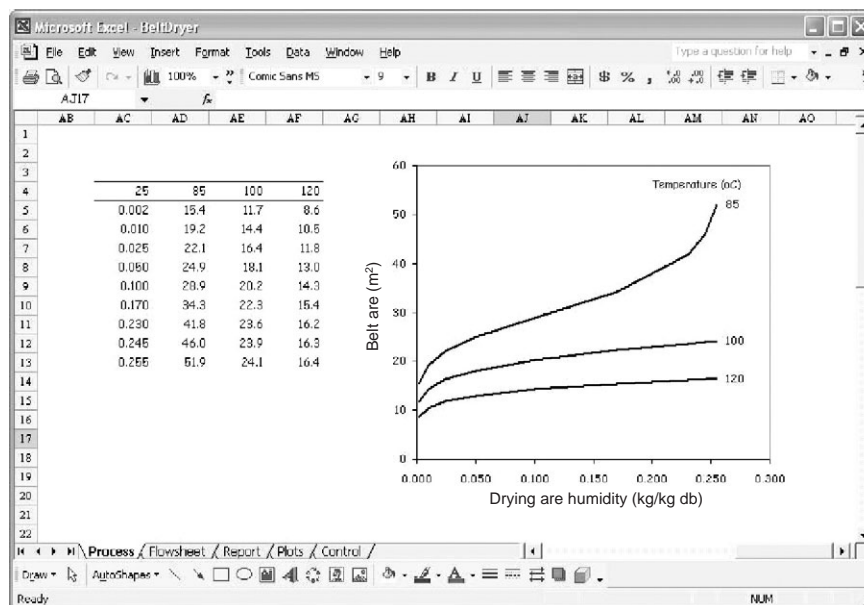


FIGURE 5.8 Analyze the effect of two design variables on a technical variable, using the “Two-Dimensional Table” and “Chart” tools, supported by Excel.

Belt Dryer Design Report		page 1/5
Problem Formulation		
Material	xxxxxxx	
Process Specification		
Feed Flow Rate	$F = 0.1 \text{ ton/h}$	
Initial Material Moisture Content	$X_o = 10 \text{ kg/kg db}$	
Final Material Moisture Content	$X = 0.1 \text{ kg/kg db}$	
Material Characteristic Size	$d = 0.01 \text{ m}$	
Ambient Temperature	$T_o = 25^\circ\text{C}$	
Ambient Humidity	$Y_o = 0.01 \text{ kg/kg db}$	
Loading Depth	$Z_o = 0.2 \text{ m}$	
Ambient Pressure	$P = 1 \text{ bar}$	
Heating Steam Temperature	$T_s = 160^\circ\text{C}$	
Design Variables		
Drying Air Humidity	$Y = 0.1 \text{ kg/kg db}$	
Drying Air Temperature	$T = 90^\circ\text{C}$	
Drying Air Velocity	$V = 1.5 \text{ m/s}$	
Belt Width	$D = 2 \text{ m}$	

Belt Dryer Design Report		page 3/5
Cost Data		
Utility Cost		
Electricity	$C_e = 0.10 \text{ \$/kWh}$	
Heating Steam	$C_s = 0.05 \text{ \$/kWh}$	
Equipment Unit Cost		
Belt Dryer	$C_{bel} = 25.0 \text{ k\$/m}^2$	
Heat Exchanger	$C_{exh} = 2.00 \text{ k\$/m}^2$	
Fan	$C_{fan} = 1.00 \text{ k\$/kW}$	
Equipment Size Scaling Factor		
Belt Dryer	$n_{bel} = 0.95$	
Heat Exchanger	$n_{exh} = 0.65$	
Fan	$n_{fan} = 0.75$	
Other		
Annual Operating Time	$t_y = 4000 \text{ h/y}$	
Interest Rate	$i_r = 0.08$	
Lifetime	$l_f = 5.00 \text{ y}$	

Belt Dryer Design Report		page 2/5
Technical Data		
Density		
Water	$\rho_w = 1.00 \text{ ton/m}^3$	
Air	$\rho_a = 1.00 \text{ kg/m}^3$	
Dry Material	$\rho_s = 1.75 \text{ ton/m}^3$	
Specific Heat		
Water	$C_{pw} = 4.20 \text{ kJ/kgK}$	
Water Vapor	$C_{pv} = 1.90 \text{ kJ/kgK}$	
Air	$C_{pa} = 1.00 \text{ kJ/kgK}$	
Dry Material	$C_{ps} = 2.00 \text{ kJ/kgK}$	
Latent Heat		
Steam Condensation	$\Delta H_o = 2.50 \text{ MJ/kg}$	
Other		
Heat Transfer Coefficient	$U_s = 0.10 \text{ kW/m}^2\text{K}$	
Void Fraction of Loading	$\varepsilon = 0.40$	
Antoine Equation for Vapor Pressure of Water		
	$a_1 = 1.19\text{E}+01$	
	$a_2 = 3.99\text{E}+03$	
Oswin Equation for Material Isotherms	$a_3 = 2.34\text{E}+02$	
	$b_1 = 7.35\text{E}-04$	
	$b_2 = 1.75\text{E}+03$	
Drying Kinetics Equations	$b_3 = 4.00\text{E}-01$	
	$c_0 = 5.00\text{E}-01$	
	$c_1 = 1.40\text{E}+00$	
	$c_2 = -2.50\text{E}-01$	
	$c_3 = -1.65\text{E}+00$	
Belt Driver Power Equation	$c_4 = 1.20\text{E}-01$	
Pressure Loss Equation	$e_1 = 2.00\text{E}+00$	
	$f_1 = 2.00\text{E}+00$	

Belt Dryer Design Report		page 4/5
Results		
Drying Air		
Fresh Air Flowrate	$F_a = 11 \text{ ton/h}$	
Recycle Air Flowrate	$F_r = 137 \text{ ton/h}$	
Drying Air Temperature	$T = 90^\circ\text{C}$	
Drying Air Humidity	$y = 0.1 \text{ kg/kg db}$	
Drying Air Velocity	$V = 1.5 \text{ m/s}$	
Drying Pressure	$P = 1 \text{ bar}$	
Ambient Temperature	$T_o = 25^\circ\text{C}$	
Ambient Humidity	$Y_o = 0.01 \text{ kg/kg db}$	
Water Activity at Drying Conditions	$a_w = 0.20$	
Material		
Material Flow Rate	$F = 0.1 \text{ ton/h}$	
Initial Moisture Content	$X_o = 10 \text{ kg/kg db}$	
Target Moisture Content	$X = 0.10 \text{ kg/kg db}$	
Equilibrium Moisture Content	$X_e = 0.05 \text{ kg/kg db}$	
Particle Size	$d = 0.01 \text{ m}$	
Drying Time Constant	$t_c = 0.54 \text{ h}$	
Drying Time	$t = 2.87 \text{ h}$	
Dryer		
Drying Rate	$W = 0.99 \text{ ton/h}$	
Dryer Length	$L = 12.6 \text{ m}$	
Dryer Width	$D = 2 \text{ m}$	
Dryer Mass Holdup	$M = 3.16 \text{ ton}$	
Dryer Volume Holdup	$H = 5.06 \text{ m}^3$	
Belt Area	$A_b = 25.3 \text{ m}^2$	
Air Heater Transfer Area	$A_s = 130 \text{ m}^2$	
Belt Velocity	$u = 4.40 \text{ m/s}$	
Loading Depth	$Z_o = 0.20 \text{ m}$	
Thermal Load		
Water Vaporization	$Q_{wb} = 0.63 \text{ MW}$	
Solid Heating	$Q_{sh} = 0.08 \text{ MW}$	
Air Heating	$Q_{ah} = 0.20 \text{ MW}$	
Total Thermal Load	$Q = 0.91 \text{ MW}$	
Electrical Load		
Belt Drive	$E_b = 27.8 \text{ kW}$	
Fan	$E_f = 34.1 \text{ kW}$	
Total Power Requirement	$E = 62.0 \text{ kW}$	
Performance		
Thermal Efficiency	$n = 0.69$	
Specific Rate of Evaporation	$r = 39.1 \text{ kg/hm}^2$	

FIGURE 5.9 A summary report of the process design results.

Belt Dryer Design Report		page 5/5
Cost Analysis Results		
<u>Equipment Cost</u>		
Belt Dryer	538	
Heat Exchanger	47	
Fan	14	
Total	$C_{eq} = 599$ k\$	
<u>Operating Cost</u>		
Electricity	25	
Heating Steam	182	
Total	$C_{op} = 207$ k\$	
<u>Annualized</u>		
Equipment	150	
Operating	207	
Total	$TAC = 357$ k\$/y	

FIGURE 5.9 (continued)

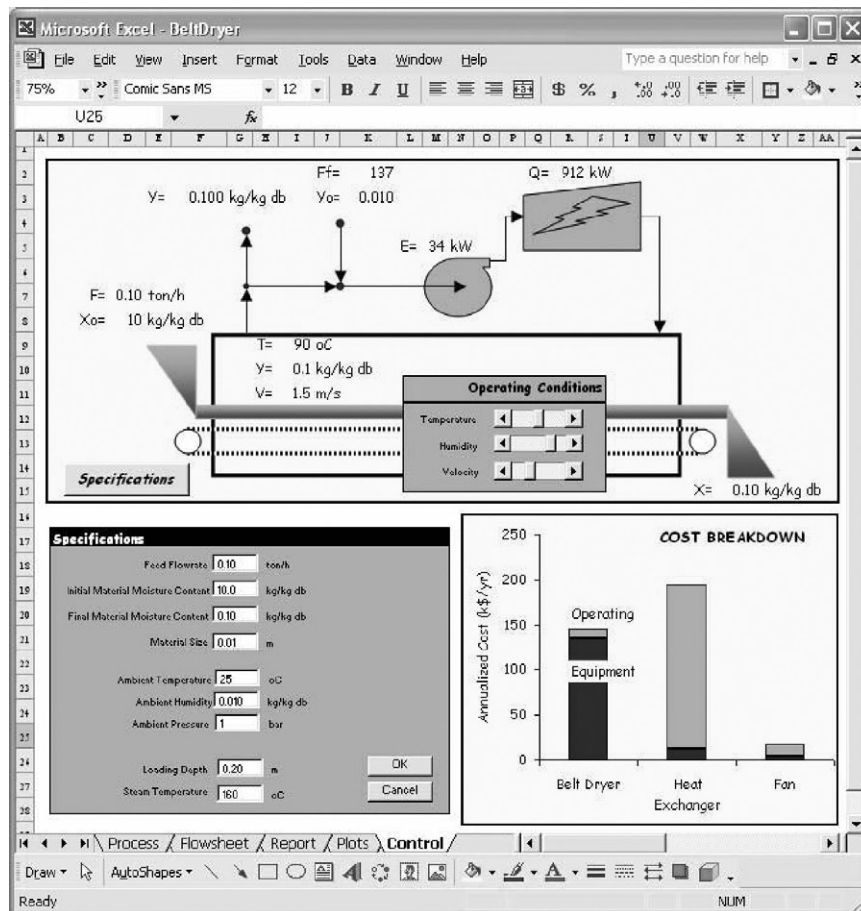


FIGURE 5.10 Toward an integrated graphics interface.

t_y	annual operating time, h/yr
u_b	belt velocity, m/s
U_s	heat transfer coefficient at air heater, kW/m ² K
V	drying air velocity, m/s
W	evaporating capacity, ton/h
X	final moisture content, kg/kg db
X_e	equilibrium moisture content, kg/kg db
X_0	initial moisture content, kg/kg db
Y	drying air humidity, kg/kg db
Y_0	ambient air humidity, kg/kg db
Z_0	loading depth, m
ΔH_0	latent heat of water evaporation at 0°C, kJ/kg
ΔP	pressure loss of air, bar
ε	void (empty) fraction of loading

ρ_a	air density, kg/m ³
ρ_m	construction material density, kg/m ³
ρ_s	dry material density, kg/m ³
ρ_w	water density, kg/m ³

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Part II

Description of Various Dryer Types

6 Indirect Dryers

Sakamon Devahastin and Arun S. Mujumdar

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6.1 INTRODUCTION

Indirect (or contact or conductive) dryers, according to the definition given by Hall (1980), are dryers in which the heating medium (e.g., steam, hot gas, thermal fluids) does not come into contact with the product being dried. Instead, wet material is dried by contact with a heated surface; heat transfer to the wet material is mainly by conduction from this surface. The temperature of heat transfer surfaces may range from -40°C (as in freeze drying) to about 300°C (as in the case of indirect dryers heated by direct combustion of products such as waste sludges) whereas the temperature of the solids in the dryer will be close to the boiling temperature of the moisture component being evaporated at the dryer operating pressure. As no hot gas is required as a source of heat in indirect dryers, only low gas flow (which is usually heated before entering the drying chamber to minimize localized condensation) or vacuum is needed to carry away the moisture evaporated from the wet material so that the drying chamber does not become saturated with vapor. Comparing to their direct (convective) counterparts, indirect dryers have a number of advantages and are summarized below (Walsh, 1992; Oakley, 1997; Vetere and Morris, 1997; Mujumdar, 2000).

Generally, indirect dryers have higher energy efficiency than the direct dryers, because the energy lost through the exhaust gas stream is greatly reduced; the

heat load of the dryers is only from the material as it dries. Waste heat source can also be used to cut drying energy costs. The vapor produced from the drying material can also be used as a heating medium in a subsequent stage of drying, hence reducing the consumption of the primary heat transfer medium. If the vaporization rate is high, e.g., when drying pulp, a two- or three-stage dryer may be economic.

Indirect dryers also need minimal cleaning of the exhaust gas as the exhaust flow is low. The sizes of the blower and interconnecting ductwork are also smaller compared to those required for the direct dryers. Additional relevant benefits are very low emission of fines and particles from the dryer; this is especially beneficial when drying toxic, explosive, flammable, or dusty products. Any vapors released can be condensed easily, thus alleviating serious environmental problems.

Vacuum drying can also be accommodated easily with the use of indirect dryers. This is ideal for heat-sensitive materials such as foods, pharmaceuticals, and other biomaterials as drying can take place at much lower temperatures (due to the reduced boiling points of solvents being removed). Oxidative reactions are minimized or eliminated due to low (or no) oxygen content. The final achievable moisture content of the product is also not limited by the state of the heating medium as in the case of direct dryers. Risk of fire and explosion hazards is eliminated as drying can be conducted in a vacuum or a modified atmosphere (e.g., inerting with nitrogen) having low oxygen

content. It should be noted, however, that it is quite difficult to operate an indirect dryer in a continuous mode under vacuum.

Another noted advantage of indirect dryers is the higher product quality attainable. This is because the flows into and out of the dryer are relatively small; when operating in a batch mode an indirect dryer is practically a closed system. As a result, this type of dryer is suitable for situations where hygienic processing conditions are required.

Accompanying the above advantages of indirect dryers are several limitations such as their limited ability to enhance the drying rates due to their limited availability of the heat transfer areas, especially in large-scale equipment. Indirect dryers also possess lower maximum drying temperature and maximum throughput when compared to direct dryers. Many types of indirect dryers can only be operated in batch mode; hence they have lower production capabilities than direct dryers. In addition, indirect dryers have typically higher capital costs than equivalent direct dryers due to the need for heat exchange surfaces.

A typical indirect dryer is a metal-walled, heat-jacketed chamber that is either stationary or rotating and in contact with the wet material being dried. In some cases, the drying chamber is equipped with an agitator. Condensing steam (most popular heating medium), hot water, combustion gases, molten salt, or electricity (less popular due to high cost involved) can be used to heat the jacket, which in turn transfers heat to the drying surface (by conduction or, in some cases, conduction and radiation). The heating medium and the product to be dried are separated by a drying surface as shown schematically in Figure 6.1. The agitator or the rotation of the drying chamber

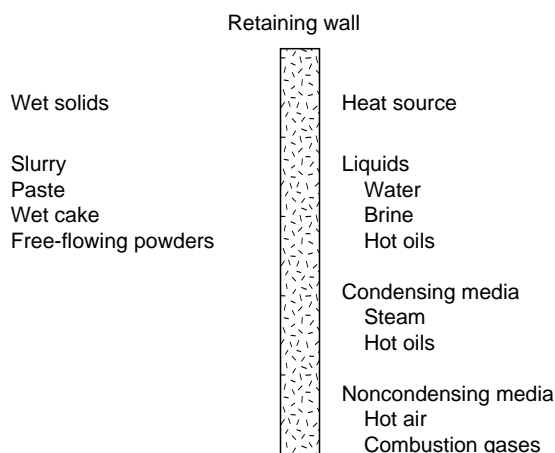


FIGURE 6.1 Schematic diagram of a typical indirect dryer; material to be dried is separated from heat source by a conducting wall.

moves and mixes the bed of drying material, thus eliminating moisture gradients within the bulk of the bed and increasing the drying rate. Low gas flow or vacuum is used to carry away the moisture evaporated from the wet material as mentioned earlier. The typical energy consumption is in the range of 2800–3600 kJ/kg of water evaporated compared to direct dryers, which consume 4000–6000 kJ/kg of water evaporated. The labor for charging and discharging the drying material and the subsequent cleanup constitute a major operating cost of this type of dryers (Vetere and Morris, 1997). Indirect dryers have been used successfully to dry a wide variety of products ranging from food and dairy products such as baby food, potato flakes, buttermilk powder, and coffee grounds sludge to chemical and other miscellaneous products such as carbon black, colloidal clay, pigments, peat, and many kinds of sludge (e.g., pulp mill sludge, metal oxide sludge). In addition to drying, indirect dryers can also be used to cook, hydrolyze, mix, react, stabilize (physically, chemically, and biochemically) as well as pasteurize the material being processed. In some units combination of these operations can be performed (by, for example, zoning of the dryer housing jacket).

More recently, combined direct–indirect dryers, e.g., fluidized-bed dryer with immersed heat exchangers, are becoming popular as they combine advantages of both dryer types. Such types of dryers are presented elsewhere in the handbook.

6.2 CLASSIFICATION AND SELECTION CRITERIA

The choice of an indirect dryer is generally determined by the material to be dried. Some common indirect dryers and their applications to various feedstock forms are listed in Table 6.1. It should be noted that some additional operations, e.g., product back-mixing, could be employed to change the material characteristics, which permits the use of additional dryer types than those listed here. Table 6.2 lists the types of indirect dryers that are suitable for various requirements, both in terms of the operating conditions and the types of feedstock.

Product quality is always of great concern when selecting the type of dryer to be used. Controlling the level of product volatiles is also, in some cases, important. To satisfy these demands the selected dryer should provide the optimum conditions for the time–temperature exposure of the drying material (Kimball, 2001). An indirect dryer that uses high heat-source temperature should be operated in such a way that the residence time of the drying material within it is

TABLE 6.1
Indirect Dryer Selection Based on Feedstock Form

Nature of Feed	Liquids		Pastes	Cakes		Powder	Granules	Free-Flowing Solids		Formed-Shaped Solids	
	Solutions	Slurry		Centrifuge	Filter			Fragile			Fiber
								Crystal	Pellet		
Dryer types											
Thin-film contact		×	×	×	×	×	×		×	×	
Cone, batch				×	×	×	×		×		
Drum	×	×	×								
Steam-jacket rotary				×	×	×	×		×	×	
Tray, batch				×	×	×	×	×	×	×	
Tray, continuous				×	×	×	×	×	×	×	

Source: Adapted from Kimball, G., *Chemical Engineering*, May 2001, pp. 74–81.

minimized; the dryer that operates at lower temperature can be operated with a longer residence time of material within it. In applications that require low product volatile levels a dryer with long residence-time capabilities and gas counterflow may be required. Table 6.3 lists the solids exposure time of some selected indirect dryers. It can be seen that a wide range of residence time, i.e., from a few seconds to few hours, can be accommodated by using different types of indirect dryers.

The guidelines for the selection of some specific indirect dryers, e.g., agitated vacuum dryers, are available in the literature (Fuller, 1999).

6.3 TYPES OF INDIRECT DRYERS

6.3.1 BATCH TRAY DRYERS

In this basic type of indirect dryers the material to be dried is placed in pans or trays on the hollow shelves,

TABLE 6.2
Indirect Dryer Selection Criteria

Dryer Type	Plate	Drum	Tumbling	Vibrating	Conical	Thin Film	Paddle	Mixer-Kneader
<i>Requirement</i>								
Continuous	×	×	—	×	—	×	×	×
Discontinuous	—	—	×	—	×	—	×	×
Vacuum	×		+		×	×	×	×
Large surface area/volume	+	○	○	○	+	+	+	×
High specific capacity	+	×	○	×	+	×	+	×
<i>Materials</i>								
Friable	×	—	×	×	×	+	×	×
Fluid	—	×	—	—	—	×	○	×
Viscous/pasty	—	×	—	—	—	—	○	×
Crusty	—	×	—	—	—	○	+	×
<i>Processing</i>								
Mechanical	×	×	×	○	×	+	+	+
Thermal	+	○	○	+	+	×	○	○

—, not suitable; ○, sometimes suitable; +, good; ×, ideal.

Source: Thurner, F., *Chemical Engineering*, October 1993, pp. 20–22.

TABLE 6.3
Solids Exposure Time of Some Indirect Dryers

Typical Residence Time of Dryer	0–10 s	10–30 s	5–10 min	10–60 min	1–6 h
Thin-film contact	×	×	×		
Cone, batch				×	
Drum		×			
Steam-jacket rotary				×	
Tray, batch					×
Tray, continuous				×	

Source: Adapted from Kimball, G., *Chemical Engineering*, May 2001, pp. 74–81.

which are heated by the heating medium, which can range from high-pressure steam for moderate-to-high temperature operation to subatmospheric steam for low-temperature operation to hot oil, or even by an electric heater in the case of smaller units. The trays are generally metal to ensure good heat transfer between the trays and the shelves. The number of shelves ranges from 1 to more than 20 shelves in a large unit.

For heat-sensitive materials drying can take place at low temperature with the application of vacuum to reduce the boiling point of the liquid to be removed, which can be toxic or valuable solvents that must be recovered. Vacuum is applied to the drying chamber and vapor is removed through an exhaust pipe connected to the chamber; this line is also connected to a condenser where vapor is condensed and, possibly, recovered. The noncondensable gas exits through the vacuum source, which can be a vacuum pump or a steam-jet ejector.

In a standard design all shelves are contained in a single drying chamber, whereas in an alternative design the dryer may be divided into a number of

separate chambers to allow more flexible operation as individual trays can be loaded or unloaded separately. Typically, each tray may be loaded up to a depth of about 40 mm giving a loading of approximately 40 kg of wet material/m² of tray area (Oakley, 1997).

This type of dryers, along with the use of vacuum, is used extensively for drying heat-sensitive or easily oxidized materials. They are particularly well suited for materials, which require gentle handling and where material loss must be minimized. Hygroscopic materials may also be dried completely at temperatures lower than the maximum permissible temperatures of those materials. The major disadvantages of this type of dryers are the high labor cost involved during the loading and unloading of the drying materials and the low capacities of the units.

6.3.2 INDIRECT-CONTACT ROTARY DRYERS

The most common type of indirect-contact rotary dryers is the steam-tube dryer (Figure 6.2). This type of dryer basically consists of a cylindrical shell, typically inclined slightly (1° to 5°) to the horizontal to

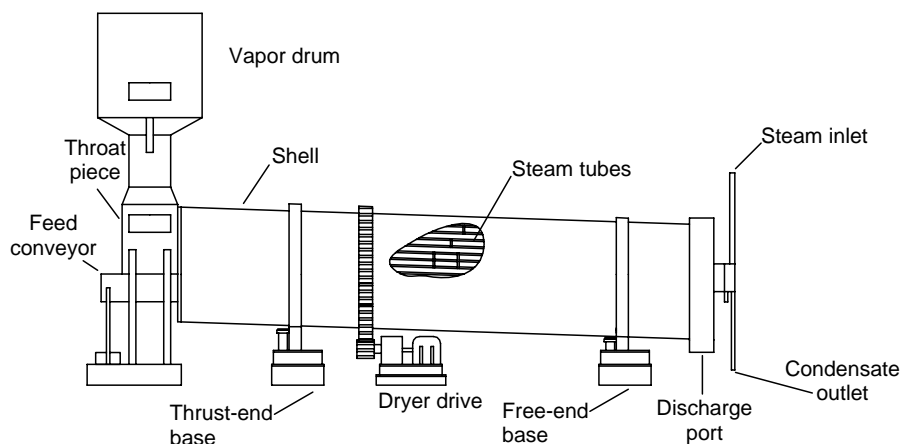


FIGURE 6.2 Indirect-contact rotary dryer.

facilitate the transportation of the wet feedstock through its body, and in which a number of steam-heated tubes are placed symmetrically (in one, two, or three concentric rows) around its perimeter and rotate with it. The length to diameter ratio of the rotating shell can vary from 4:1 to 10:1. Steam-heated tubes may either be simple pipes with condensate draining by gravity into the discharge manifold or bayonet-type; typical steam pressure used may be in the range of 4–10 bar (van't Land, 1991). When handling sticky materials one row of tubes is preferred. Lifting flights are usually inserted behind the tubes to aid agitation. Typically, indirect rotary dryers would be used when direct-contact rotary dryers are not suitable either because direct contact with hot gases is not permissible for product quality reasons or because of the presence of fine particles, which can be entrained along with the drying gas flows (Oakley, 1997).

Wet feed enters the dryer at the upper end of the shell and tumbles around the steam tubes and dries. Gentle airflow (generally countercurrent to solids flow) is normally applied to the drying chamber to carry away the evaporated moisture. This type of dryer is used for continuous drying or heating of granular or powdery solids, which cannot be exposed to ordinary atmospheric or combustion gases. It is especially suitable for fine dusty particles as only low gas velocities are needed to purge the drying chamber.

Comparing to their direct-contact counterparts indirect-contact dryers possess some advantages such as the reduced deterioration of the drying material due to an excessive exposure of the material to the high-temperature drying medium. In addition, adsorption of such toxic substances in the drying medium (in the case where direct combustion gases are used) as NO_x is eliminated.

The thermal efficiency of steam-tube dryers is in the range of 70–90%, if well insulated. This number does not allow for boiler efficiency, so it cannot be directly compared with that of direct-heat units, however. Heat transfer coefficients in steam-tube dryers may range from 30 to 85 $\text{W}/(\text{m}^2 \text{ K})$ and these values increase with increasing steam temperature due to an increased effect of radiation. The heat flux of these dryers, carrying saturated steam at 140–170°C, may range from 6300 W/m^2 for difficult-to-dry and organic products to between 1900 and 3800 W/m^2 for fine inorganic materials. Typical performance data of steam-tube dryers on some selected applications are shown in Table 6.4 (Moyers and Baldwin, 1997).

Mathematical models that specifically enable prediction of the moisture content and temperature profiles of the material undergoing drying in an indirect-contact steam-tube dryer are available in the literature (e.g., Shene and Bravo, 1998; Canales et al., 2001).

TABLE 6.4
Indirect Steam-Tube Dryer Performance Data

	Type 1	Type 2	Type 3
Type of materials processed	High-moisture organic, distillers' grains, citrus pulp	Pigment filter cakes, precipitated chalk	Finely divided inorganic solids, floatation concentrates
Feed description	Wet feed is granular solid and damp but not sticky and dries to granular solid	Wet feed is pasty and dries to hard pellets	Wet feed is crumbly and friable and dries to powdery product of very few lumps
Moisture content of wet feed (% d.b.)	223	100	54
Moisture content of product (% d.b.)	11	0.15	0.5
Normal temperature of wet feed (K)	310–320	280–290	280–290
Normal temperature of product (K)	350–355	380–410	365–375
Evaporation per kg of product (kg)	2	1	0.53
Heat load per kg of product (kJ)	4950	2620	1375
Heating surface required per kg of product (m^2)	0.34	0.4	0.072
Steam (at 860 kPa gauge) consumption per kg of product (kg)	3.33	1.72	0.85

Source: Adapted from Moyers C.G., Baldwin, G.W., in R.H. Perry and D.W. Green, eds. *Perry's Chemical Engineers' Handbook*, 7th ed., McGraw-Hill, New York, 1997, pp. 12-1–12-90.

6.3.3 ROTATING BATCH VACUUM DRYERS

The rotating (or double cone) batch vacuum dryer is essentially a rotating vessel containing materials to be dried. This type of dryers is widely used to dry such materials as free-flowing powders, granules, and crystals. Due to the gentle tumbling motion this type of dryers, however, is not suitable for processing sticky materials, which may stick to the dryer walls or form lumps. Vacuum is used in conjunction with drying when low solid temperatures must be maintained due to heat-sensitive nature of the products or to avoid oxidation or fire or explosion.

Rotating batch vacuum dryers (Figure 6.3) consist of a heated vacuum chamber, which rotates about a horizontal axis. Wet material is loaded through the charge opening and the vessel is closed and evacuated down to the desired operating pressure. Heat for drying is supplied by a surrounding heating jacket. During the drying cycle the vessel rotates and imparts a tumbling motion to the drying material, aiding heat transfer, mixing, and vapor release. The speed of rotation may vary from 5 rpm for a large unit to 30 rpm for a smaller unit. The temperature of the heating jacket may be varied through the drying cycle; it is also common to replace the heating medium by cooling water in the final stage of drying to cool down the product.

The major disadvantages of this type of dryers are the limited heat transfer areas, especially in larger units. One possible solution to this problem is the addition of internal heating panels; however, this alternative is prone to fouling. The stickiness problem can be alleviated by rotating the vessel slowly or intermittently when the drying material is at high moisture content. However, truly sticky materials may not still be processed in this type of dryers.

The overall heat transfer coefficients of this type of dryer, which may vary near $35 \text{ W}/(\text{m}^2 \text{ K})$, depend

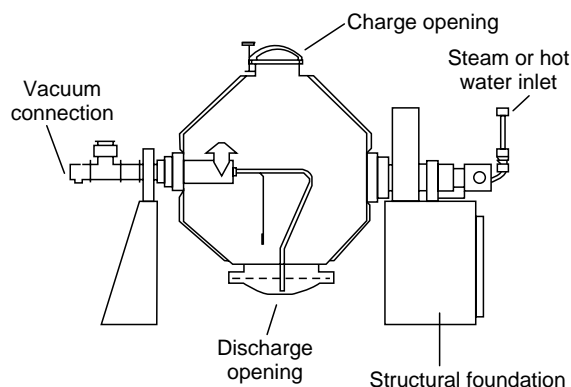


FIGURE 6.3 Rotating batch vacuum dryer.

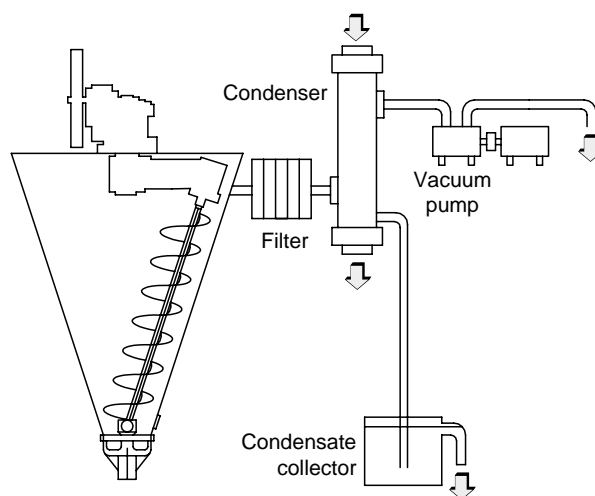


FIGURE 6.4 Cone dryer.

largely on the resistance between the inner jacket wall and the drying solids, which depends to a large extent on the solid characteristics. The overall heat transfer coefficients may drop considerably if the dryer walls are fouled; the values in the range of $5\text{--}10 \text{ W}/(\text{m}^2 \text{ K})$ are not uncommon.

6.3.4 AGITATED DRYERS

The first type of agitated dryers is the vertically agitated dryer (shown in Figure 6.4 as a cone dryer). Batch cone dryer (or conical mixer dryer) is a conically shaped vessel, with capacities ranging from 50 l to 25 m^3 , and is commonly used for drying solvent or water-wet, free-flowing pharmaceuticals, and fine chemical products, so the dryer is normally operated under vacuum. Agitating action is provided by an internally mounted screw, which can be heated to provide additional heating (possibly in the range of 10–30%) to that available through the vessel wall, which rotates about its own axis and also moves around the dryer on an orbiting arm. The distance between the screw and the vessel wall is kept small so that the layer of material on the heating surface is continually renewed. As the screw rotates around the full circumference of the vessel it also provides self-cleaning for the vessel walls as well. The cover is also normally heated to avoid vapor condensation in the process area. To avoid contamination of the product good sealing of bearings is an important consideration (Oakley, 1997).

The heating medium for this type of dryer is either hot water, steam, or hot oil in the temperature range of $50\text{--}150^\circ\text{C}$ and pressure in the range of 3–30 kPa absolute. The vapor generated during drying is evacuated by a vacuum pump and, if necessary, passed

through a condenser for recovery of solvent. For design purposes the values of the heat transfer coefficients can be assumed to be around $60 \text{ W/(m}^2 \text{ K)}$ (Moyers and Baldwin, 1997).

Another type of agitated dryers is the horizontally agitated dryer, which consists of a stationary horizontal cylindrical shell in which a set of agitator blades mounted on a central shaft mixes and conveys the wet material to be dried. Heat is supplied to the dryer, and indirectly to the drying material, by circulation of hot water, steam, or heat transfer fluid through a jacket surrounding the shell and, in larger units, through the hollow central shaft. This class of dryer can indeed be subdivided according to the agitator type viz., paddle dryers and screw conveyor dryers. A variety of batch and continuous units are available and are employed to dry different types of sticky and free-flowing solids, pastes, and suspensions. To use this type of dryers to dry heat-sensitive material, again, the dryer may be operated in a vacuum mode. Materials such as chocolate crumb, cornmeal, and confectionary materials have successfully been dried in this type of dryers (Oakley, 1997).

Figure 6.5 shows a typical setup of a paddle dryer. The agitator consists of a rotating shaft to which paddles are attached; discs or coils may be used instead of paddles in other types of designs. Typical rotational speeds of the agitators are between 10 and 30 rpm. The selection of the type of agitators depends on the nature and handling behavior of the

material being dried. In some modern design spring-loaded shell scrapers or breaker bars may also be installed to aid handling of sticky materials. To minimize the build-up of solids on the inner surface of the jacket in order to maintain a high heat transfer rate the gap between the vessel wall and the paddle must be again kept very small, generally in a few millimeters range. Solids mixing and transport are accomplished by angling the paddles relative to the rotor arm.

The wet feedstock enters, via the application of a pump, a screw conveyor or a rotary valve, at one end of the vessel and the discharge port is generally located on the opposite side. As shown in Figure 6.5 for a batch unit the discharge port is normally located at the center of the housing, whereas for a continuous unit the discharge port may be located at the end of the dryer. The discharge of material from the dryer also occurs by means of a screw conveyor, a rotary valve, or a double butterfly valve. Exhausted vapor or moisture is withdrawn from the dryer through the exhaust situated on the top of the vessel, which is connected to gas cleaning equipment, condenser, vacuum pump, or exhaust fan. Inert gas such as nitrogen can be used to purge the drying chamber to aid removal of vapor as well.

For design purposes the overall heat transfer coefficients of this type of dryer, which depend largely on the film coefficient between the inner jacket wall and the solids, which in turn depend largely on the solid characteristics, can be assumed to be around $50 \text{ W/(m}^2 \text{ K)}$ (Moyers and Baldwin, 1997). However, higher values are expected for products that have surface moisture (van't Land, 1991).

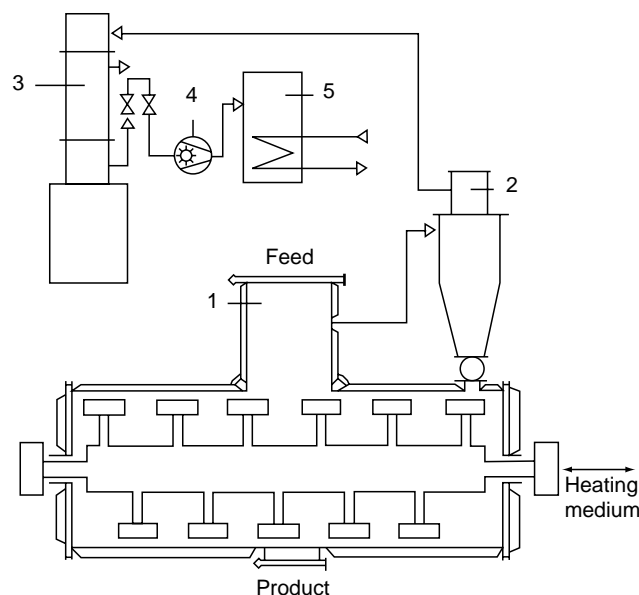


FIGURE 6.5 Horizontal agitated dryer (1, horizontal agitated dryer; 2, dust collector; 3, surface condenser; 4, liquid ring vacuum pump; 5, water circulation pump).

6.3.5 OTHER TYPES OF INDIRECT DRYERS

Although not common it is possible to use superheated steam as the drying medium for certain dryers, which are heated indirectly with high pressure condensing steam in the jacket, e.g., a flash dryer for pulp or hog fuel. Relative motion between the heated surface and the material being dried always helps increasing both heat and mass transfer rates. Such motion can be achieved by mechanical stirring, vibration, rotation or rolling of the material, fluidization, or spouting. Generalized correlations for estimation of the heat transfer coefficients under a wide assortment of operating conditions are not available in the open literature, however. It is important to note that such data can be highly dependent on the physical characteristics of the material, e.g., size and shape of particles, moisture, stickiness, and tendency to adhere to the heat transfer surface. Materials that may undergo glass transition over temperature ranges

encountered may pose unexpected problems. In addition, significant differences between the heat transfer coefficients as well as thermal contact resistance between the wall and the material being heated may occur for pasty materials and materials that have tendency to form large lumps.

Under a wider definition of indirect dryers one may technically include radiant and microwave/radio frequency (RF) dryers as there is no direct contact between the drying solid and the heat source (or medium). However, these dryer types are covered elsewhere in the handbook. Drum drying of thin pastes also involves indirect drying but it is covered under a different chapter in this handbook.

6.4 DESIGN AND MODELING OF INDIRECT DRYERS

Generally, it is possible to calculate the rate of heat transfer from the surface of an immersing body to either packed or stirred bed of granular material (or from the heating surface to the drying material in other types of indirect dryers such as drum dryers) using the following equation (Kimball, 2001):

$$Q = UA(T_1 - T_2) \quad (6.1)$$

where Q is the heat transfer rate, U is the overall heat transfer coefficient, A is the heat transfer area, T_1 is the temperature of the heat transfer medium, and T_2 is the temperature of the drying particles. The overall heat transfer resistance is the sum of a number of film and thermal resistances in the system as follows (Walsh, 1992):

$$\frac{1}{U} = \frac{1}{h_h} + \frac{1}{h_w} + \frac{1}{h_f} + \frac{1}{h_m} + \frac{1}{h_v} \quad (6.2)$$

where $1/h_h$ is the resistance between the heating medium and the conducting wall, $1/h_w$ the resistance across the conducting wall, $1/h_f$ the resistance between the wall and the surface of the bed of drying material, $1/h_m$ the resistance across the bed of drying material, and $1/h_v$ the resistance at the evaporating surface.

Although there are five resistances in series, there are indeed only two main resistances, i.e., the resistance between the wall and the surface of the bed of drying material (or contact resistance) as well as the resistance across the bed of drying material (or bulk penetration resistance). In some special cases h_m may be much larger than h_f . In general, h_m is infinitely large as the bed is isothermal such as during the initial start-up of the drying process when the bed is at its initial temperature, when the heat capacity of the bed

approaches infinity as a result of chemical reactions or evaporation, or when there is a perfect mixing of the bed by stirring (Schlünder, 1984).

Schlünder (1984) gave the formula for the prediction of the contact resistance and bulk penetration resistance for packed beds and stirred beds of materials in contact with the surface of immersed bodies. For the typical engineering applications the time-averaged heat penetration coefficient for a packed bed (or a moving bed with plug flow pattern) can be predicted by

$$h_m = \frac{2}{\pi} \frac{\sqrt{(\rho c \lambda)_{\text{bed}}}}{\sqrt{t}} \quad (6.3)$$

and

$$\frac{U}{h_f} = \frac{1}{1 + \frac{\sqrt{\pi}}{2} \sqrt{\tau}} \quad (6.4)$$

where ρ , c , and λ are respectively the density (kg/m^3), heat capacity at constant pressure, (J/(kg K)), and thermal conductivity, (W/(m K)) of the bed. $\tau = (h_f^2/(\rho c \lambda)_{\text{bed}})t$ is the dimensionless residence time of the packed bed (or moving bed with plug flow pattern) at the heated wall. By letting $h_m = h_f$ it is also possible to derive a formula for the critical residence time t_c ; if the residence time of the bed at the heated wall is less than t_c there is no difference in the value of U whether the bed is stirred or not:

$$t_c = \frac{4}{\pi} \frac{(\rho c \lambda)_{\text{bed}}}{h_f^2} \quad (6.5)$$

To obtain the similar expressions for a stirred bed it is first assumed that there exists some fictitious time period t_R during which the bed is assumed to be a packed bed and the heat transfer to the bed is governed by Equation 6.4. After t_R a perfect mixing of the bed is assumed. This assumption yields oscillating instantaneous heat transfer coefficient and the time average of these values yields the overall heat transfer coefficient of the stirred bed U_∞ . The value of t_R is calculated by

$$t_R = N_{\text{mix}} t_{\text{mix}} \quad (6.6)$$

where N_{mix} is the so-called mixing number, which is an experimentally determined parameter and is dependent only on the mechanical property of the system as well as on the Froude number. The mixing number lies roughly in the order of 2–25 and is rather independent of the hot surface temperature, operating pressure, or the moisture content of the drying

material. It can be simply thought of as the number of revolutions of the stirrer required to perfectly mix the bed of particles once. As a rough guess, Schlünder and Mollekopf (1984) recommended the following correlation for the prediction of the mixing number

$$N_{\text{mix}} = C Fr^x \quad (6.7)$$

where $Fr = ((2\pi n)^2 D/2g)$, D is the diameter of the disc or the rotary drum and n is the number of revolutions of the mixing device. The values of C and x for various types of dryers are listed in Table 6.5.

Finally, t_{mix} is the time constant of the stirrer, which may be taken as the time required for one revolution of the stirrer. The dimensionless residence time in Equation 6.4 is replaced by

$$\tau_R = N_{\text{therm}} N_{\text{mix}} \quad (6.8)$$

where $N_{\text{therm}} = (h_f^2 t_{\text{mix}} / (\rho c \lambda)_{\text{bed}})$. The formula for the prediction of the overall heat transfer coefficient for a stirred bed can then be written as

$$\frac{U_{\infty}}{h_f} = \frac{1}{1 + \frac{\sqrt{\pi}}{2} \sqrt{N_{\text{therm}} N_{\text{mix}}}} \quad (6.9)$$

To estimate the value of the contact heat transfer coefficient, h_f , the following simplified equation is used (Schlünder, 1984):

$$h_f = \phi_A h_{\text{fp}} + (1 - \phi_A) \frac{2\lambda_G/d}{\sqrt{2} + (2l + 2\delta)/d} + h_{\text{rad}} \quad (6.10)$$

where h_{fp} is the heat transfer coefficient for a single particle and is calculated by

$$h_{\text{fp}} = \frac{4\lambda_G}{d} \left[\left(1 + \frac{2l + 2\delta}{d} \right) \ln \left(1 + \frac{d}{2l + 2\delta} \right) - 1 \right] \quad (6.11)$$

where λ_G is the thermal conductivity of the gas, ϕ_A is a plate surface coverage factor, which has the value of

around 0.8, d is the particle diameter, δ is the roughness of the particle surface (can be assumed to be zero in most cases or if at all necessary can be set at 1 μm), and l is the modified mean free path of the gas molecules, which can be calculated by

$$l = 2\Lambda \frac{2 - \gamma}{\gamma} \quad (6.12)$$

where Λ is the mean free path of the gas molecules and γ is the so-called accommodation coefficient, which lies in the range of 0.8–1 for normal gases at moderate temperatures. The mean free path of the gas molecules is calculated from the following equation:

$$\Lambda = \frac{16}{5} \sqrt{\frac{RT}{2\pi M}} \frac{\mu}{p} \quad (6.13)$$

where μ is the gas dynamic viscosity, p and T are the gas pressure and temperature, respectively, M is the gas molecular weight, and R is the universal gas constant.

Finally, h_{rad} , which takes into account the effect of radiation, is calculated by the following equation:

$$h_{\text{rad}} = 4C_{12} T_m^3 \quad (6.14)$$

where the overall radiation exchange coefficient, C_{12} , is calculated by

$$C_{12} = \sigma \frac{1}{\frac{1}{\varepsilon_{\text{wall}}} + \frac{1}{\varepsilon_{\text{bed}}} - 1} \quad (6.15)$$

in which σ is the blackbody radiation coefficient of $5.67 \times 10^{-8} \text{ W/(m}^2 \text{ K}^4)$ and $\varepsilon_{\text{wall}}$ and ε_{bed} are respectively the emissivities of the wall and of the bed surface. In most indirect dryer calculations, however, the effect of radiation is small as the temperatures used are often low.

To estimate the values of the thermal conductivity of packed beds of either monodispersed or polydispersed spherical or nonspherical particles within the temperature range of 100–1500 K and the pressure range of 10^{-3} to 100 bar the following correlations may be used (Schlünder, 1984):

$$\frac{\lambda_{\text{bed}}}{\lambda_G} = (1 - \sqrt{1 - \psi}) \left(\frac{\psi}{\psi - 1 + \lambda_G/\lambda_D} + \psi \frac{\lambda_R}{\lambda_G} \right) + \sqrt{1 - \psi} \left[\phi_K \frac{\lambda_s}{\lambda_G} + (1 - \phi_K) \frac{\lambda'_{\text{bed}}}{\lambda_G} \right] \quad (6.16)$$

TABLE 6.5
Typical Values of C and x in Equation 6.7

Dryer Type	C	x
Disc dryers	25	0.20
Rotary drum dryers	16	0.20
Paddle dryers	9	0.05

and

$$\frac{\lambda'_{\text{bed}}}{\lambda_G} = \frac{2}{K} \left(\frac{B(\lambda_s/\lambda_G + \lambda_R/\lambda_G - 1)(\lambda_G/\lambda_D)(\lambda_G/\lambda_S)}{K^2} \right. \\ \left. \ln \frac{(\lambda_s/\lambda_G + \lambda_R/\lambda_G)\lambda_G/\lambda_D}{B[1 + (\lambda_G/\lambda_D - 1)(\lambda_s/\lambda_G + \lambda_R/\lambda_G)]} \right. \\ \left. + \frac{B+1}{2B} \left\{ \frac{\lambda_R}{\lambda_G} \frac{\lambda_G}{\lambda_D} - B \left[1 + \left(\frac{\lambda_G}{\lambda_D} - 1 \right) \frac{\lambda_R}{\lambda_G} \right] \right\} \right. \\ \left. - \frac{B-1}{K} \frac{\lambda_G}{\lambda_D} \right)$$

where

$$K = \frac{\lambda_G}{\lambda_D} \left[1 + \left(\frac{\lambda_R}{\lambda_G} - B \frac{\lambda_D}{\lambda_G} \right) \frac{\lambda_G}{\lambda_S} \right] \\ - B \left(\frac{\lambda_G}{\lambda_D} - 1 \right) \left(1 + \frac{\lambda_R}{\lambda_G} \frac{\lambda_G}{\lambda_S} \right) \quad (6.17)$$

$$B = C_{\text{shape}} \left(\frac{1 - \psi}{\psi} \right)^{0.9} f(\xi_r) \quad (6.18)$$

$$\frac{\lambda_R}{\lambda_G} = \frac{4c_S}{2/(\varepsilon - 1)} T_m^3 \frac{x_R}{\lambda_G} \quad (6.19)$$

$$\frac{\lambda_G}{\lambda_D} = 1 + \frac{2A}{x_D} \left(\frac{2}{\gamma} - 1 \right) \quad (6.20)$$

where $x_R = R_{\text{shape}}d$ and $x_D = D_{\text{shape}}d$. In this case, d is the equivalent particle diameter, which equals $d = \sqrt[3]{6V/\pi}$ when V is the particle volume. R_{shape} and D_{shape} are the shape factors for the interstitial energy transport by radiation and molecular flow, respectively. The values of the relative particle to particle contact surface area, ϕ_K and the three shape factors, C_{shape} , R_{shape} , and D_{shape} , must be evaluated from the experiments. Typical values of all these parameters for spherical particles are: $C_{\text{shape}} = 1.25$, $R_{\text{shape}} = 1$, $D_{\text{shape}} = 1$, and $\phi_K = 0.0077$ (ceramic), 0.0013 (steel), and 0.0253 (copper).

If the packed bed consists of particles of various sizes of mass fractions Δz_i the values of x_R and x_D must be calculated by the following equations:

$$\frac{1}{x_R} = \sum_{i=1}^n \frac{\Delta z_i}{R_{\text{shape},i}d_i} \quad (6.21)$$

$$\frac{1}{x_D} = \sum_{i=1}^n \frac{\Delta z_i}{D_{\text{shape},i}d_i} \quad (6.22)$$

Finally, the particle size distribution function, $f(\xi_r)$, is calculated by the following equations

$$f(\xi_r) = 1 + 3\xi_1 \quad (6.23)$$

where the distribution coefficient, ξ_1 , is calculated by

$$\xi_1 = \left[\frac{\sum_{i=1}^n \frac{\Delta z_i}{d_i^2}}{\left(\sum_{i=1}^n \frac{\Delta z_i}{d_i} \right)^2} - 1 \right]^{1/2} \quad (6.24)$$

Predicted and experimental heat transfer coefficients are found to agree well (within $\pm 25\%$) within the following ranges of parameters: Particle diameter: $4 \mu\text{m} < d < 3100 \mu\text{m}$; pressure: $10^{-3} \text{ Torr} < p < 760 \text{ Torr}$. The experiments were performed with such a wide range of materials as polystyrene, glass, sand, copper, aluminum, bronze, and celite under the atmospheres of air, helium, and Freon.

As mentioned earlier, for other types of indirect dryers than agitated beds (e.g., drum dryers), [Equation 6.1](#) can be used to calculate the rate of heat transfer from the heating surface to the drying material. The values of the overall heat transfer coefficient reported vary rather widely. For example, the values between 105 and 345 W/(m²K) are reported for drum dryers. The layer of drying material on the surface of the drum also affects significantly the value of the heat transfer coefficient due to the presence of the moisture evaporation from the surface of the drying material; the value of the heat transfer coefficient may increase by four times when there is a layer of the drying material on the drum surface (Tolmac and Lambic, 1997). Walsh (1992) also reported an increase in the value of the overall heat transfer coefficients of various types of indirect dryers by 2.5–3 times when feed moisture content increases from 10 to 65% (w.b.). Lecomte et al. (2004) indeed provided a simple method for the design of a contact dryer for sludge drying involving thin film boiling using a simple experimental device to measure the heat transfer rate between a heated metal plate and a thin layer of the sludge. Indirect drying of sludge offers advantages such as high thermal efficiency, volatile organic compound (VOC) concentration and volume reduction, and odor control. Their study showed that a strong boiling period existed initially with very high heat fluxes, which was followed by a transient phase and finally a phase with low heat fluxes. This technique is useful for the design of indirectly heated drum dryers.

For indirect-heat rotary steam-tube dryers the values of heat transfer coefficients may range from 30 to 90 W/(m² K) depending on the dryer rotational speed, flow rate of air (used to carry away evaporated vapor), and steam pressure used (Moyers and Baldwin, 1997; Vega et al., 2000). In units carrying

saturated steam at 145–175°C the heat flux ($U\Delta T$) may range from 6300 W/m² for difficult-to-dry and organic solids to 1900–3900 W/m² for finely divided inorganic materials. Typical values of various resistances (in terms of various heat transfer coefficients) are given in Table 6.6.

For agitated-bed dryers, in the case when all the liquid evaporates at the particle surface, i.e., in the case of drying nonhygroscopic materials, the resistances to heat and mass transfer are the aforementioned contact resistance and the bulk heat penetration resistance (the bulk permeation resistance to mass transfer is generally very small and normally neglected). In addition, with decreasing average moisture content of the bed, an increasing number of already dried particles would prevent the still wet particles from contacting the heated surface and hence reduces the actual drying rate further. In such case the drying rate $\dot{m}(X)$ can be predicted together with the bulk temperature T_b following a stepwise calculation of moisture content (from the initial moisture content) and bulk temperature (from the initial bulk temperature, which equals the saturation temperature) as follows (Schlünder and Mollekopf, 1984; Tsotsas and Schlünder, 1987):

$$\dot{m}(X) = \frac{U_{\text{wet}}(T_w - T_s) \exp(-\xi^2)}{\lambda} \quad (6.25)$$

The decrease in moisture content of the packing during the contact period is predicted by

$$\Delta X = \frac{[\dot{m}(X)t_R A]}{M_{\text{dry}}} \quad (6.26)$$

TABLE 6.6
Typical Values of Various Resistances

Heat Transfer Coefficient (Resistance) ⁻¹	Upper Limit (W/(m ² K))	Lower Limit (W/(m ² K))
h_h	10,500 (steam)	570 (hot liquid)
h_w	13,650 (3-mm steel)	1050 (corrosion-resistant lining)
h_f	280 (50% solid paste)	2.8 (dry solid)
h_m	1050 (3-mm thick layer of medium conductivity material)	4.5 (25-mm thick layer of low conductivity material)
h_v	1050	1050

Source: Adapted from Walsh, J.J., *Indirect Drying of Solids Particles*, Bepex Corporation, Minneapolis, 1992.

and the change in average bulk temperature is calculated from

$$\Delta T_b = \frac{\lambda}{c_{p,\text{bed}} + Xc_{p,L}} \frac{1 - \exp(-\xi^2)}{\exp(-\xi^2)} \Delta X \quad (6.27)$$

where $c_{p,\text{bed}}$ and $c_{p,L}$ are the heat capacity of the dry bed and of the evaporating liquid, respectively, T_w is the contact surface temperature, T_s is the saturation temperature, and λ is the latent heat of evaporation of water. The ratio of the overall heat transfer coefficient of the wet bed to the contact heat transfer coefficient is determined from

$$\frac{U_{\text{wet}}}{h_f} = \frac{1}{1 + (h_f/U_{\text{dry}} - 1)\text{erf}(\xi)} \quad (6.28)$$

where

$$\sqrt{\pi} \exp(\xi^2) \left[1 + \left(\frac{h_f}{U_{\text{dry}}} - 1 \right) \text{erf}(\xi) \right] = \left(\frac{h_f}{U_{\text{dry}} - 1} \right) \frac{1}{\xi} \quad (6.29)$$

$$\xi = \frac{X\lambda}{c_{p,\text{bed}}(T_w - T_b)} \quad (6.30)$$

Finally, the ratio of the overall heat transfer coefficient for dry bed to the contact heat transfer coefficient is determined from

$$\frac{U_{\text{dry}}}{h_f} = \frac{1}{1 + (\sqrt{\pi}/2)\sqrt{N_{\text{therm}}N_{\text{mix}}}} \quad (6.31)$$

For hygroscopic materials the effect of bound moisture can be taken into account, for example in the case of vacuum drying, by the use of an effective heat capacity of the hygroscopic bed as (Tsotsas and Schlünder, 1987)

$$c_{p,\text{bed},h} = c_{p,\text{bed}} + c_{p,L}X_h - \lambda \left(\frac{\partial X_h}{\partial T} \right)_p \quad (6.32)$$

where $(\partial X_h/\partial T)_p$ is the slope of the sorption isobar and X_h is the bound moisture. If the sorption isobar is assumed to be linear, Equation 6.32 can then be written as

$$c_{p,\text{bed},h} = c_{p,\text{bed}} + c_{p,L}X_h - C_1\lambda \quad (6.33)$$

where C_1 is a constant.

Finally, the decrease of bound moisture content can be obtained from

$$\Delta X_h = C_1 \Delta T \quad (6.34)$$

and the sum of ΔX_f and ΔX_h gives the total decrease of moisture content of the bed and can be used together with Equation 6.26 to calculate the overall drying rate.

Once the drying rate is obtained it is possible to determine the size of a contact dryer by finding the value of the following integral:

$$I = \int_{X_{out}}^{X_{in}} \frac{dX}{\dot{m}(X)} \quad (6.35)$$

In the case of a batch dryer this integral yields the product of the required contact surface area A and the required residence time t_{dry}

$$At_{dry} = M_{dry} I \quad (6.36)$$

where M_{dry} is the bone dry mass of the drying particles. For a continuous dryer operating at a through-put \dot{M}_{dry} the required contact area is

$$A = \dot{M}_{dry} I \quad (6.37)$$

For the calculation of the drying rates of granular beds wetted with a binary mixture, the reader is referred to, for example, Heimann and Schlünder (1988).

During the constant rate period, when heat transfer dominates the drying process, a dryer with a large heating surface is preferred. On the other hand, during the falling rate period when internal resistances control the rate of drying, a dryer that allows a longer residence time of the drying material within it is recommended.

Except for the dryers operated under vacuum some air leakage is not economically avoidable and up to 6 m³/min of air leakage is considered normal for large-scale indirect dryers. When gentle airflow is used to carry away evaporated vapor from the dryer its velocity is generally in the range of 3 to 6 m/min and its dew point in the range of 60 to 90°C (Cook and DuMont, 1991).

For sample calculations of some types of indirect dryers, e.g., paddle dryers and indirect-contact rotary dryers, the reader is referred to van't Land (1991).

6.5 CONCLUSION

Indirect dryers, when they can be used, are thermally more efficient because less or no convective heating

medium is required to provide the needed thermal energy for vaporization of moisture. Under vacuum conditions there is no mixing of the condensable (typically water) with noncondensable gas (e.g., air). Thus, heat recovery from the dryer exhaust is more efficient and cost-effective. Combined mode dryers, for example, those utilizing convection along with conduction or radiation or dielectric heating are becoming more popular as they tend to combine the respective advantages of the different modes of heat delivery. It is noteworthy that immersed surface heat transfer between surface-wet solid particles and surface-dry particles in fluidized beds or vibrated beds can differ by factors of 4–8 depending on the particle size, surface temperature, and texture; a much higher rate is observed under the wet condition. This, however, lasts only for a short time if much of the moisture to be removed is internal. For design purposes, to obtain a conservative estimate of the immersed surface heat transfer rate, it is recommended that the abundant dry particle data and correlations be used.

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7 Rotary Drying

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7.1 INTRODUCTION

Rotary drying is one of the many drying methods existing in unit operations of chemical engineering. The drying takes place in rotary dryers, which consist of a cylindrical shell rotated upon bearings and usually slightly inclined to the horizontal. Wet feed is introduced into the upper end of the dryer and the feed progresses through it by virtue of rotation, head effect, and slope of the shell and dried product withdrawn at the lower end. A simplified diagram of a direct-heat rotary dryer is presented in [Figure 7.1](#). The direction of gas flow through the cylinder relative to the solids is dictated mainly by the properties of the processed material. Cocurrent flow is used for heat-sensitive materials even for high inlet gas temperature due to the rapid cooling of the gas during initial evaporation of surface moisture, whereas for other materials countercurrent flow is desirable in order to

take advantage of the higher thermal efficiency that can be achieved in this way. In the first case, gas flow increases the rate of solids flow, whereas it retards it in the second case [3,19,20,35].

7.2 TYPES OF ROTARY DRYERS

Rotary dryers are classified as direct, indirect-direct, indirect, and special types. This classification is based upon the method of heat transfer being direct when heat is added to or removed from the solids by direct exchange between gas and solids, and being indirect when the heating medium is separated from contact with the solids by a metal wall or tube. There is an infinite number of variations, which present operating characteristics suitable for drying, chemical reactions, mixing, solvent recovery, thermal decompositions, sintering, and agglomeration of solids [35].

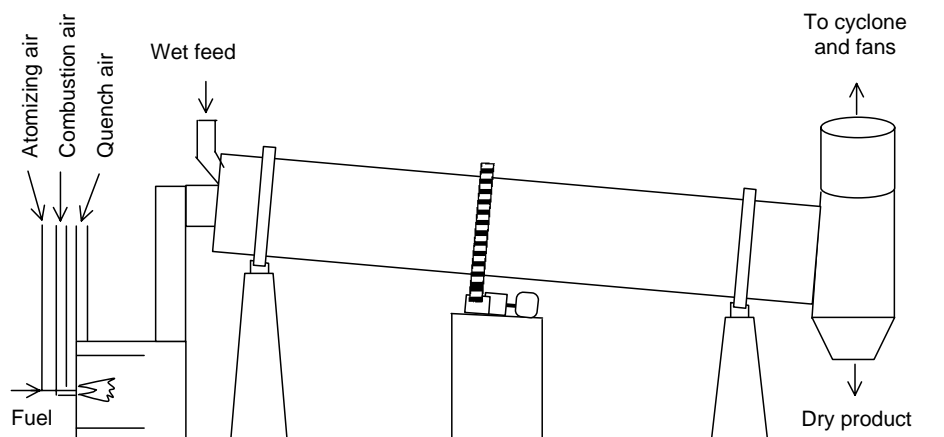


FIGURE 7.1 Simplified diagram of direct-heat rotary dryer.

The main types of rotary dryers include the following:

- *Direct rotary dryer.* It consists of a bare metal cylinder with or without flights, and it is suitable for low- and medium-temperature operations, which are limited by the strength characteristics of the metal.
- *Direct rotary kiln.* It consists of a metal cylinder lined in the interior with insulating block or refractory brick, in order to be suitable for operation at high temperatures.
- *Indirect steam-tube dryer.* It consists of a bare metal cylindrical shell with one or more rows of metal tubes installed longitudinally in its interior. It is suitable for operation up to the available steam temperature or in processes requiring water-cooling of the tubes.
- *Indirect rotary calciner.* It consists of a bare metal cylinder surrounded by a fired or electrically heated furnace and it is suitable for operation at temperatures up to the maximum that can be tolerated by the metal of the cylinder, usually 800–1025 K for stainless steel and 650–700 K for carbon steel.
- *Direct Roto-Louvre dryer.* It is, perhaps, the most important of the special types, as the solids progress in a crosscurrent motion to the gas, and it is suitable for low- and medium-temperature operations.

The rotary dryers can perform batch or continuous processing of the wet feed, and the discharged product should be solids relatively free flowing and granular. If the material is not completely free flowing in its feed condition, a special operation is necessary, which includes recycling a portion of the final product, a premixing with the feed or maintaining a bed of

free-flowing product in the cylinder at the feed end. The direct-heat dryers are the simplest and most economical and are used when the contact between the solids and gases or air is not harmful. However, if the solids contain extremely fine particles, excessive entrainment losses in the exit gas stream is possible, due to the large gas volumes and high gas velocities that are, usually, required.

The indirect types require only sufficient gas flow through the cylinder to remove vapors, and have the advantage to be suitable for processes requiring special gas atmospheres and exclusion of outside air.

The auxiliary equipment of a direct-heated rotary dryer includes a combustion chamber for operation at high temperatures, while steam coils are used for low temperatures. Gases are forced through the cylinder by either an exhaustor (especially when a low-pressure drop heater is employed) or an exhaustor–blower combination, which is suitable for maintaining precise control of internal pressure even in the case of high-pressure drop in the system. The material characteristics determine the method of feeding of the rotary dryer, which can be done by a chute extending into the cylindrical shell or by a screw feeder for sealing purposes or if gravity feed is not convenient. The feedrate should be controlled and uniform in quality and quantity. In the exit end of the dryer, cyclone collectors are usually installed for the removal of the dust entrained in the exit gas stream. Bag collectors in case of expensive materials or extremely fine product may follow cyclone collectors. Wet scrubbers may be used when toxic solids or gases are processed, the temperature of the exit gas is high, the gas is close to saturation or there is recirculation of the gas. Insulation and steam tracing usually required for cyclones and bag collectors, and an exhaust fan should be used downstream from the collection system.

For the reduction of heat losses the dryer (especially cocurrent direct-heat dryers) and its equipment should be insulated, except when brick-lined vessels or direct-heat dryers operating at high temperatures are employed. In the last case, heat losses from the shell cause a cooling of its material and prevent overheating.

The rotary dryers (direct-heat dryers and kilns) are controlled by indirect means, e.g., by measuring and controlling the gas temperatures in their two ends, whereas shell temperature is measured on indirect calciners, and steam pressure and temperature as well exit gas temperature and humidity are controlled on steam-tube dryers. It is not possible to achieve control by measuring the product temperature because not only this is difficult but also its changes are slowly detected, although the product temperature is used for secondary controls.

External shell knockers are often used for removing solids sticking on flights and walls. In case of large cross section, internal elements or partitions can be used to increase the effectiveness of material distribution and reduce dusting.

For systems operating at temperature higher than 425 K and are electrically driven, the existence of auxiliary power sources and drivers is necessary, as loss of rotation will cause sagging of the cylinder.

Representative materials dried in direct-heat rotary dryers are sand, stone, ilmenite ore, sodium sulfate, sodium chloride, and fluorspar, for which high temperatures are used, cellulose acetate, sodium chloride, styrene, copperas, cast-iron borings, and ammonium sulfate, for which medium temperatures are required, and urea prills, vinyl resins, oxalic acid,

urea crystals, and ammonium nitrate prills, that are dried at low temperatures [35].

7.3 FLIGHT DESIGN

Of all types of rotary dryers the ones that have been studied more extensively are the direct-heat rotary dryers equipped with peripheral flights, while very little scientific work has been published for the other types. Their purpose is to lift and shower the solid particles through the gas stream promoting intimate contact between wet solids and hot gases. The flights are usually offset every 0.6–2 m and their shape depends upon the characteristics of the solids. Radial flights with a 90° lip are used for free-flowing materials and flat radial flights without lip for sticky ones. It is a common practice to employ different flight designs along the dryer length to accommodate with the changing characteristics of the material during drying. In the first meter or so at the feed end spiral flights are used for better distribution of the material under the feed chute or conveyor. The flights most commonly used are presented in Figure 7.2 [35].

Flights a, b, c, and d of Figure 7.2 are frequently used in cascading rotary dryers; the first one is suitable for sticky solids in the wet end of the dryer, while the fourth one, which has a semicircular shape, has been proposed by Purcell [45], because it is supposed to be formed easier in comparison with types b and c. The last two designs have been proposed on the basis of theory for improving dryer's performance, but their profile is rather complex. They have been studied by Kelly [19] and include the equal angular distribution (EAD) flight and the centrally biased distribution

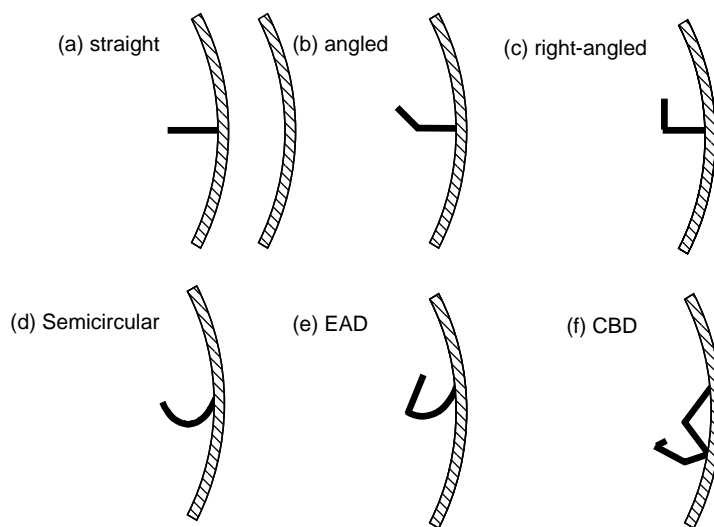
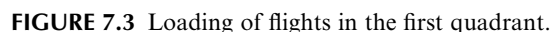


FIGURE 7.2 Common flight profiles.

Schofield and Glikin [49] determine this angle from an equilibrium balance of the forces acting on a particle, which is about to fall from a flight. Gravitational force φ_g , centrifugal force φ_c , and frictional force φ_f act on the particle, which is the product of the dynamic coefficient of friction γ as it slides down the surface of like particles by the normal reaction of this surface on the particle φ_n . The force balance yields the following equation:

where θ is the angle subtended by the flight lip at the center of the drum, and $\nu = r_e \omega^2/g$ is the ratio of the centrifugal to the gravitational forces acting on the particle.

Rotary dryers are usually operated in the range $0.0025 \leq \nu \leq 0.04$, therefore the above equation gives accurate results over the range of practical importance, considering that Kelly [19] and Purcell [45] found that it is valid for values of ν up to about 0.4. It has to be mentioned that this equation was tested for free-flowing solids having a constant moisture content. In practice the moisture content decreases



The flight becomes empty for $\theta - \varphi - \beta = 90^\circ$ where β is the angle subtended by the flight at the center of the drum and is given by the relationship

The maximum loading occurs at $\theta = 0^\circ$ and is equal to

where $\tan \varphi_0 = ((\gamma + \nu) / (1 - \nu\gamma))$.

Analogous analysis has been done for other common flights, for example, Baker [3] described in the same way for the angled- and extended-circular flights.

The total amount of solids contained in the drum is about 10–15% of its volume. It has been proved empirically that this loading gives the most efficient performance, therefore a sufficient number of flights must be provided to contain and distribute these solids. Assuming that there are n_f flights in the shell, the spacing between each will be

$$\theta_i = 360^\circ / n_f \quad (7.8)$$

In the case of right-angled flights, it has been proved by Glikin [11] that the minimum spacing between them must be such as to satisfy the equation

$$r_e \tan(\theta_i - \beta) > l \tan \varphi_0 \quad (7.9)$$

in order for the flights to load completely at $\theta = 0^\circ$.

The holdup of any particular flight, in the upper section of the drum, decreases, as the cylindrical shell rotates, from its maximum value h_0^* to zero at a value of θ equal to or, usually, less than 180° . According to Glikin [11], the loading on any flight in the bottom half of the drum is the mirror image of the flight positioned vertically above it in the upper section, and if the number of flights is even, the total holdup in the flights in a design-loaded drum will be

$$H^* = 2 \sum h^* - h_0^* \quad (7.10)$$

In this equation, the sum includes the holdup of each flight in the upper half of the shell, thus for $0^\circ \leq \theta \leq 180^\circ$.

A revised equation suggested by Kelly and O'Donnell [22], which has the form

$$H^* = \frac{h_0^*(n_f + 1)}{2} \quad (7.11)$$

This relationship is more accurate when the particles cascade across the whole upper region. Nevertheless, in most practical cases cascading ceases for θ much less than 180° [19], and then that equation gives a value of H^* much higher than the correct one. Glikin [11] proved that the discrepancy could get up to 80% or more.

The design of the flights, not only determines the holdup of the dryer, but also the manner in which solids are shed from them. Kelly [19] has published many data about the distribution of cascading solids across the drum for right-angled, semicircular, and angled flights, but did not give detailed information on the geometry of them. It is not easy to determine which flight profile is the most efficient. Of course, particles cascading down the center of the shell will present the longest contact time with the hot gases, but the fact that the cascading is concentrated in a particular area, will cause considerable shielding of the particles by their neighbors, resulting in inefficient heat and mass transfer.

The average length of fall depends on characteristics of the shell, flights, and particles and is given by the equation

$$\bar{Y} = \frac{\int_{h_0}^0 Y dh^*}{\int_{h_0}^0 dh^*} \quad (7.12)$$

where h_0 is the actual holdup in the flight at $\theta = 0^\circ$ and h^* the design holdup at any other value of angle θ . h_0 may be less or equal to the design holdup h_0^* .

Kelly [19] proposed the expression $Y = D_e (\sin \theta / \cos \alpha)$ for underloaded and design-loaded drum; therefore

$$\bar{Y} = \frac{D_e}{\cos \alpha} \frac{\int_{h_0}^0 \sin \theta dh^*}{\int_{h_0}^0 dh^*} \quad (7.13)$$

In general, the solution of the above equation requires numerical integration; only in a few special cases there is an analytical solution, like for EAD flights that was presented by Kelly [19]. Thus, for a design-loaded drum the following simple equation may be used:

$$\bar{Y} = \frac{2D_e}{\pi \cos \alpha} \quad (7.14)$$

In an overloaded drum cascading commences at $\theta = 0^\circ$, while in an underloaded one cascading only starts at some angle between 0° and 180° at which the actual holdup becomes equal to the design one. The following revised expression gives the average distance of fall in an overloaded drum:

$$\bar{Y} = \frac{2D_e}{M\pi \cos \alpha} \quad (7.15)$$

where $M = H/H^* \geq 1$.

The next general expression gives the average distance of fall in cascading rotary dryers:

$$\bar{Y} = \frac{k' D_e}{M \cos \alpha} \quad (7.16)$$

The constant k' depends upon the flight geometry and its value for different design-loaded flights are given in Table 7.1 [19].

7.4 RESIDENCE TIME MODELS

A rotary dryer is a conveyor of solid material and at the same time promotes heat and mass transfer between the drying material and the hot gas. The particles move through the dryer by three distinct and independent mechanisms, and are described as follows.

TABLE 7.1
Values of k' for Different Design-Loaded Flights

Flight Profile	k'
Semicircular	0.570
Equal angular distribution (EAD)	0.637
Right-angled	0.760
Equal horizontal distribution (EHD)	0.784
Centrally biased distribution (CBD)	0.902

7.4.1 CASCADE MOTION

This is the result of the lifting action of the flights and the slope of the dryer. The advance of a particle per cascade is equal to $D_e (\sin \theta / \tan \alpha)$ assuming that the descent path of the particle is vertical when there is no gas flow. With cocurrent gas flow there is increased advance of the particle due to the drag on the cascading solids, while the reverse action occurs with countercurrent flow.

7.4.2 KILN ACTION

It is the motion of the particles as they slide either over the metal surface in the lower half of the shell, or over one another. Due to the slope of the dryer the particles proceed to its exit. This movement can also appear in horizontal drums as a result of the “hydraulic gradient” of the solids. Kiln action is always present, but is of major importance for overloaded dryers.

7.4.3 BOUNCING

This motion occurs when a falling particle rebounds from the shell surface or from the settler layer of particles, instead of come to rest, and results in the particle progress because of the dryer slope.

The average residence time (or, time of passage) $\bar{\tau}$ is defined as holdup H divided by the solids feedrate F , thus

$$\bar{\tau} = \frac{H}{F} \quad (7.17)$$

Theoretically, holdup can be measured directly. Nevertheless, in an industrial dryer this measurement is inconvenient, because the system must shut down and its content has to be discharged and weighed. In order to avoid that, a radioisotope or a small amount (0.5–1.0 kg) of an inert detectable solid may be added to the feed and analyzed in the product. The time required for the maximum concentration to occur represents the average time of passage [35].

Most of the studies referred to particles residence time, consider average holdups and residence time. To determine the distribution of residence times, Miskell and Marshall [31] used closely sized 496- μm sand containing a radioactive tracer in a 0.14-m diameter flighted drum, and found that the residence time is normally distributed. Fan and Ahn [8] showed that an axial dispersion model could describe the above results. Porter and Masson [43] concluded that deviations from plug flow are not large after examination of two cocurrent industrial dryers. However, it is not safe to assume plug flow of the particles in industrial dryers, because only narrowly sized materials were studied; furthermore just two dryers were studied. In practice there is a wider size distribution and a wider range of residence times. Moreover, if the operating criterion is the maximum moisture content of any particle instead of the bulk average value, it is logical to consider that there will be deviations from plug flow.

In order to express residence time as a function of dryer’s characteristics Johnstone and Singh [16] proposed the equation

$$\bar{\tau} = \frac{0.0433(Ln)^{1/2}}{DN \tan \alpha} \quad (7.18)$$

where $\bar{\tau}$ is the residence time (min), L is the length, D is the diameter, N is the rotational speed (r/min), $\tan \alpha$ is the slope of the dryer, and n is the dynamic angle of repose of the solids (degrees). This formula is derived from the equation

$$\bar{\tau} = \frac{0.0310(Ln)^{1/2}}{DN \tan \alpha} \quad (7.19)$$

which is known as the “Bureau of Mines,” proposed by Sullivan et al. [52] and refers to the passage of solids through a rotary kiln not equipped with flights or retaining dams. The modified constant in Equation 7.18 stands for the action of the flights. A much more extensive experimental study on rotary dryer holdup was done by Prutton et al. [44], who correlated their data of a design-loaded shell by the following empirical expression:

$$\bar{\tau} = \frac{kL}{DN \tan \alpha} + \frac{mu}{60} \quad (7.20)$$

where k is a dimensionless constant, depending on the number and design of the flights and varies from 0.275 for 6 flights to 0.375 for 12 flights and m is a factor depending on the size and density of the particles and the direction of the airflow varies (in the range of the particular study) from -177 to $-531 \text{ s}^2/\text{m}$

for cocurrent flow and from 236 to 945 s²/m for countercurrent flow. This equation does not express m as a function of particle properties and, furthermore, is not considered to give accurate results at air velocities much higher than those used in the study, because, although it implies a linear relationship between residence time and gas velocity, it has been proved that there is a curvature in the plots between those two parameters, especially in the case of countercurrent flow at high gas rates. Perry and Chilton [35] proposed the following equation:

$$\bar{\tau} = \frac{0.23L}{DN^{0.9} \tan \alpha} \quad (7.21)$$

based on the experimental data obtained by Friedman and Marshall [9] who present a wide-ranging study on residence times and recognized that the dryer holdup is affected by the number of flights, particularly at low feedrates, even though most of their data refer at values lower than those of industrial dryers.

The following equation:

$$X_a = X_0 \pm KG \quad (7.22)$$

expresses the effect of air velocity for values up to 1 m/s, where X_a is the holdup with airflow, X_0 is the holdup without airflow, G is the gas flow rate (kg/hm²), and $K = 16.9/d_p^{1/2} \rho_b$ is a dimensional constant in which ρ_b is the bulk density (kg/m³) and d_p is the weight average particle size (μm). For cocurrent flow the negative sign stands and for countercurrent the positive. The constant K has not been proven quite sufficient.

Saeman and Mitchell [47] proposed the following expression, based on a theoretical analysis of the material's transport through the dryer taking into account the incremental transport rates associated with individual cascade paths

$$\bar{\tau} = \frac{L}{f(H)DN(\tan \alpha \pm m'u)} \quad (7.23)$$

where $f(H)$ is the cascade factor varied between 2 for lightly loaded dryers and π for heavily loaded ones with small flights. The exact value seems to be affected by the cascading pattern. The positive sign stands for cocurrent flow and the negative sign for countercurrent flow; m' is an empirical constant depended on the material. Saeman [48] developed a model for the estimation of that constant, but concluded that it is easier to measure it, due to the parameters required for the estimation, which are difficult to obtain.

Schofield and Glikin [49] analyzed the fluid mechanics of falling granules and proposed the relationship

$$\bar{\tau} = \frac{L}{\bar{Y}(\sin \alpha - K'u^2/g)} \left(\frac{1}{\sigma N} + t_f \right) \quad (7.24)$$

where \bar{Y} is the average height of fall of the particle given by Equation 7.16, g is the acceleration due to gravity, $K' = 1.5 f \rho_f / d_p \rho_p$ is a constant related to the drag coefficient f , ρ_f is the air density, ρ_p is the particles density, $1/\sigma N$ is the time spent by a particle on the flights, where $\sigma = 180/\bar{\theta}$ and $\bar{\theta} = (1/h_0) \int_0^{h_0} \theta dh$ is the angle that the particle is carried in the flights, and $t_f = (2 \bar{Y}/g)^{1/2}$ is the average time of fall of the particles, assuming that the vertical component of the air drag is negligible as was proved by Kelly [21]. Generally, $t_f \ll 1/\sigma N$

A critical point in the analysis of the above residence-time equations is that residence time is calculated from the velocity of the average particle $L/\bar{\tau}$, whereas the residence time calculated from the average velocity $\bar{L}/\bar{\tau}$ is much higher, because the particles progress through the dryer not by a simple kiln action, but there is a cascade motion of them. This observation was made by Glikin [11] who showed that for EAD flights the following expression stands:

$$(L/\bar{\tau}) \approx 0.69(\bar{L}/\bar{\tau}) \quad (7.25)$$

The average particle velocity is

$$(\bar{L}/\bar{\tau}) = \frac{Z}{h_0^*} \int_0^{h_0^*} \frac{\sin \theta}{\theta} dh^* \quad (7.26)$$

where $Z = \pi ND_c [(\sin \alpha \pm K' u_r^2/g)/\cos \alpha]$ and the relative velocity between the particles and the gas is $u_r = u \pm (1/2) \sin \alpha (2g\bar{Y})^{1/2}$. In these equations the plus sign applies for countercurrent flow and the minus sign for cocurrent flow.

Glikin showed that for cocurrent flow the residence time $\bar{\tau}$ increases with particle size d_p while the reverse relationship seems to exist for countercurrent flow.

In order to explain the discrepancies between his equation of the form

$$\bar{Z} = \frac{L_{\text{eff}}}{\bar{Y} \sin \alpha - f(u)} \left[\frac{1}{N} \left(1 - \frac{1}{2} m_0 \right) + t_f \right] \quad (7.27)$$

which stands for EAD flights and experimental results, Kelly [20] proposed that a rapid forward movement as a result of kiln action, which should be taken into account, follows the cascade motion of the particles.

Therefore, the effective length in that equation is L_{eff} instead of L , where L_{eff} is the length of the shell over which the average granule progresses due to the cascade motion only and is given by the expression

$$L_{\text{eff}} = k_c L \quad (7.28)$$

The constant k_c is a function of the loading and rotational speed, but it is independent of the slope of the drum, as Kelly's experimental procedure proved. He proposed the following empirical expression for that constant:

$$k_c = bM + b' \quad (7.29)$$

in which b and b' are functions of the rotational speed N . The values of these constants are presented in Table 7.2.

Kiln action becomes important in overloaded drums as proved by experimental data and supported by the model of Kelly and O'Donnell [23]. In underloaded drums, particle bouncing, especially on the exposed metal surface of the shell, has an important contribution to their motion.

Kelly and O'Donnell [23] present the most advance study of the particles motion through rotary dryers that have flights. Their work includes an extensive experimental procedure as well as a theoretical analysis of the behavior of the particles. They measure the cycle time and the advance per cycle for a single average particle, which was compared to the predictions of the model that incorporates cascade motion, kiln action, and bouncing. The basic features of their model are the following.

The Schiller and Naumann equation was used for the estimation of the drag coefficient and the pressure drop of the air flowing through the curtain of the falling solids to that of air flowing through the free cross section of the drum for estimation of the effect of the particle shielding.

The movement of the particles after bouncing from the shell, a flight or a bed of particles was taken into consideration. This effect is not important

because, after contact, the particle loses most of its velocity in the direction normal to the surface and, in practice, the advance of the particle after the second bounce is very limited.

There are three varieties of kiln action; the first refers to the sliding motion of the particle inside the shell after its bounce has stopped, the second occurs if the particle moves contrary to the direction of rotation of the drum and slips backward into the flight, and the third, appears only for overloaded drums, for which the holdup ratio $M > 1$, because there is a rolling load of solids in the bottom of the drum and the average particle bypasses one or more flights before being arrested by a flight. The particle advances with each bypass due to the slope.

The above features were included in a computer simulation for the calculation of the advance and time for the average particle in a single cascade, as well as the average residence time. It was proved that cascade motion and bouncing are very important in the pilot dryer; bouncing has a major effect in underloaded kilns. Under these conditions, in a pilot and an industrial dryer, about 50 and 22%, respectively, of the particle advance was due to bouncing. At the same time, kiln action accounted for less than 10% of the advance, while it was becoming important for overloaded drums. The computed values of the residence time given by the model are greater than the measured ones, and the error becomes greater as the air velocity increases. Although the model proposed by Kelly and O'Donnell is quite advanced as long as it concerns the mechanisms of particle transport in rotary dryers, it is quite complex to be used for industrial design purposes.

The kiln equations of Sullivan et al. [52] and of Johnstone and Singh [16], which are experimentally based, predict low values of the residence time in case of zero gas flow. When gas flow is applied, these relationships are inadequate, as there is no term to express that flow and, furthermore, give the same result for both cocurrent and countercurrent flow. So, these equations are unreliable for gas velocity greater than 1 m/s. The equations of Prutton et al. [44], Saeman and Mitchell [47], and Friedman and Marshall [9], which are also experimentally based, give comparable results at zero and low gas velocities, although the first and second seem to predict rather wide ranges of residence time. Their application requires judgment and experience, whereas their theoretical basis is not solid. These expressions have been formed for gas velocity less than 1.5 m/s. The standard deviation between their predictions is about 25% for gas velocities up to 1 m/s, but exceeds 100% at 3 m/s, so the extrapolation seems rather invalid. The disagreement between real and calculated values is

TABLE 7.2
Values of b and b' in Equation 7.29

N (r/min)	$0.4 < M < 1.0$		$1.0 < M < 1.6$	
	b	b'	b	b'
8	0.530	-0.124	-0.280	0.672
24	0.719	-0.178	-0.426	0.932

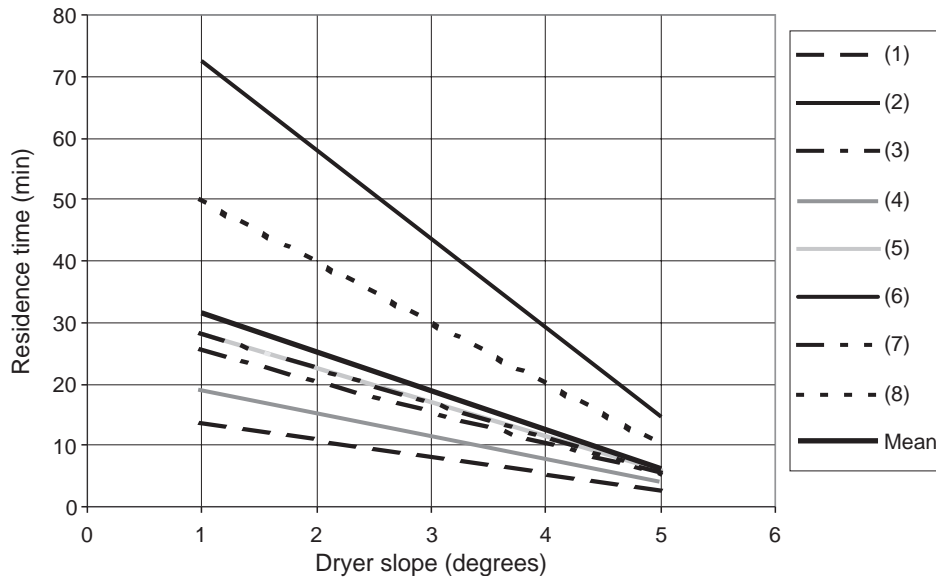


FIGURE 7.4 Residence time prediction versus dryer slope for zero airflow. 1, Sullivan et al. [52]; 2, Johnstone and Singh [16]; 3, Prutton et al. [44]; 4, Friedman and Marshall [9]; 5, Saeman and Mitchell [47]; 6, Schofield and Glikin [49]; 7, Kelly [20]; 8, Glikin [11].

expected to be greater in the range of industrial importance, which is 3–5 m/s.

The equations proposed by Schofield and Glikin [49], Kelly [20], and Glikin [11] have theoretical basis and seem to be the most accurate for zero gas flow. Under these circumstances, Kelly's model presents the best agreement with the experimental data. To some degree this occurs because of the presence of the empirical constant k_c , which has been evaluated by fitting the model to those data. The models of Schofield and Glikin [49], and Glikin [11] predict residence times much higher than the experimental ones. Kelly and Glikin used the equation of Schiller and Naumann for the estimation of the friction factor. However, this expression refers to a single particle and it cannot predict sufficiently the effect of the raining curtain of the solids as they drop from the flights, particularly at high gas velocities. In fact, Kelly [19] rejected his model in favor of an empirical method. Figure 7.4 presents the effect of dryer slope to the residence time for zero airflow according to the above-mentioned equations, while Figure 7.5 shows the residence time versus airflow velocity for cocurrent flow.

7.5 HEAT AND MASS TRANSFER IN ROTARY DRYERS

During drying, heat is supplied to the solids for the evaporation of water or, in a few cases, some other volatile component, and the removal of the corresponding vapor from the dryer.

The heat transferred in direct-heat rotary dryers is expressed by the following equation:

$$Q = U_v a V (\Delta t)_m \quad (7.30)$$

where Q is the rate of heat transfer, J/s, $U_v a$ is the volumetric heat transfer coefficient, J/(sm³K) or W/(m³K), V is the dryer volume, m³, and $(\Delta t)_m$ is the true mean temperature difference between the hot gases and the material. Miller et al. [30], Friedman and Marshall [9], and Seaman and Mitchell [47] have done considerable amount of research for the evaluation of $U_v a$. The volumetric coefficient $U_v a$ is the product of the heat transfer coefficient U_v based on the effective area of contact between the gas and the solids, and the ratio a of this area to the volume of the dryer. When a considerable amount of surface moisture is removed from the solids and their temperature is unknown, a good approximation of $(\Delta t)_m$ is the logarithmic mean between the wet-bulb depressions of the drying air at the inlet and outlet of the dryer [35].

Miller et al. [30] present the first extensive study of heat transfer in rotary dryers and conclude that the total rate of heat transfer is affected by the number of flights. It is given by the following equations:

$$Q = 1.02LD \frac{(n_f - 1)}{2} G^{0.46} \Delta t_{lm} \quad \text{for 6 flights} \quad (7.31)$$

$$Q = 0.228LD \frac{(n_f - 1)}{2} G^{0.60} \Delta t_{lm} \quad \text{for 12 flights} \quad (7.32)$$

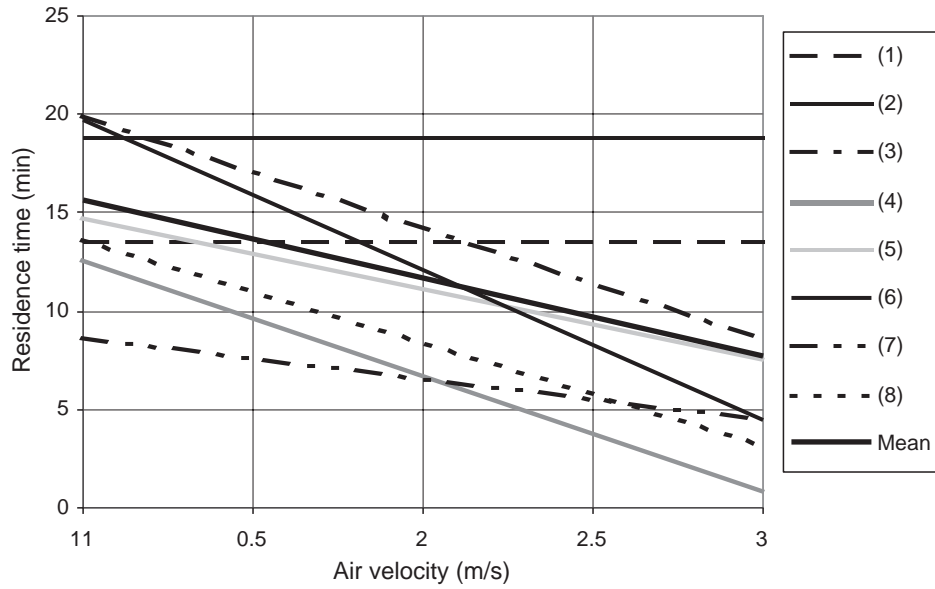


FIGURE 7.5 Residence time prediction versus drying air velocity for cocurrent flow. 1, Sullivan et al. [52]; 2, Johnstone and Singh [16]; 3, Prutton et al. [44]; 4, Friedman and Marshall [9]; 5, Saeman and Mitchell [47]; 6, Schofield and Glikin [49]; 7, Kelly [20]; 8, Glikin [11].

Comparing the above two relationships to the general equation (Equation 7.30), we can express the volumetric coefficient as

$$U_v a = 0.652(n_f - 1)D^{-1}G^{0.46} \quad \text{for 6 flights} \quad (7.33)$$

$$U_v a = 0.145(n_f - 1)D^{-1}G^{0.60} \quad \text{for 12 flights} \quad (7.34)$$

They also note that the rate of heat transfer is independent of the slope and the rotational speed of the shell, and therefore of the residence time, as well as of the flight size. The increase of the gas flowrate increases the efficiency of the dryer. Furthermore, they study a number of dryers having diameters up to 2.13 m and propose that the equations for 6 flights are more representative for the design of industrial dryers. This proposal is in agreement with the study of Prutton et al. [44]. Friedman and Marshall [9] noted that in practice the number of flights is in the range

$$6.56 \leq (n_f/D) \leq 9.84 \quad (7.35)$$

In case that $n_f \gg 1$, the following simple equation can be used:

$$U_v \alpha = K_s G^{0.46} \quad (7.36)$$

where $4.3 \leq K_s \leq 6.4$.

Friedman and Marshall concluded that the above analysis has three major simplifications and cannot predict the heat transfer quite accurately. First, the heat losses from the dryer have not been taken into account, second the use of the logarithmic mean temperature difference Δt_{lm} is not correct, as the temperature of the solids does not vary linearly with the gas temperature, and third, they express doubts for the correlations between the rate of heat transfer and the number of flights, due to the fact that Miller et al. [30], although they proposed the above equations based on experimental data, accepted that the one for 12 flights is not representative for industrial dryers. For their experiments, Friedman and Marshall [9] used an extensively insulated dryer to reduce heat losses to around 15%. They used advanced methods to achieve accurate measurements of the gas, solids, and shell along the length of the dryer, although due to circulation patterns within the dryer they obtain erratic results for the gas temperature, and had to calculate it by heat balances. From the analysis of their data they concluded that the use of coefficients based on the terminal temperature differences does not properly predict the performance of pilot dryers, which have a large shell-area-to-volume factor, although the estimations are a lot better in commercial dryers, which have relatively much smaller heat losses. Therefore, the scale-up of heat transfer data requires caution and experience. They found that $U_v \alpha$ varies proportionally with the solids holdup (as a percentage of drum volume) $X^{0.5}$ and increases with $G^{0.16}$,

therefore is affected by gas rate in two independent ways. The holdup increases with G for countercurrent flow causing an increase in the effective contact area between the solids and the gas, which is expressed by parameter α . Friedman and Marshall [9] also suggested that the heat transfer coefficient U_v increases with gas rate, although this is not generally accepted. The rate of rotation has little effect on $U_v\alpha$, as it has opposing effects on the holdup and the cascade rate. They examined the effect of the number of flights on $U_v\alpha$ and concluded that the major increase occurs.

The simplest and rather conservative equation has the form

$$U_v a = KG^n/D \quad (7.37)$$

where K is a proportional constant, G is the gas mass velocity ($\text{kg/m}^2\text{h}$), D is the dryer's diameter (m), and n is a constant. Based on the data of Friedman and Marshall [9] the constants are: $K = 44$ and $n = 0.16$ [35].

According to McCormick [29], the constant K determines flight geometry and shell speed. These parameters in addition to the number of flights seem to affect the overall balance, although there are no available data for evaluating these variables separately. As long as it concerns the gas velocity it seems that its increase breaks up the showering curtains of solids more effectively and exposes more solids surface, therefore there is an increase of a in Ua rather than U .

Saeman and Mitchell [47] suggested a more advanced approach to the heat transfer mechanism in rotary dryers (and coolers). The heat transfer takes place mainly between the cascading solids and air entrained by them. This mass of air attains thermal equilibrium with the particles surface in a very short period and when it reaches the bottom of the dryer it diffuses into the main horizontal airflow, which confines to the voids between the cascades. To support this theory, they measure the air temperatures in the voids near the bottom of the shell, which were higher than those at the top. Also, the bulk of the heat transfer occurs within about 0.3–0.6 m of the origin of the cascade, as it was proved by temperature measurements in the cascading streams. Therefore, the heat transfer rate is a function of the cascade rate, which depends on the flights number and size, the rate of rotation, and the holdup, and the ratio of air-material entrainment in the cascading streams, which depends mainly on flight size. Other parameters, such as the surface area of the particles, have less influence. They employ a heat transfer coefficient based on unit length of the dryer, thus

$$Q = U_L \alpha L \Delta t_{lm} \quad (7.38)$$

where $U_L \alpha$ is in $\text{W}/(\text{mK})$. The two coefficients $U_v \alpha$ and $U_L \alpha$ are related by the equation

$$U_L \alpha = \frac{\pi D^2}{4} U_v \alpha \quad (7.39)$$

For modern commercial dryers that have a flight count per circle of 2.4 to 3.0 D and operate at shell peripheral speeds of 60–75 ft/min, the following equation has been proposed:

$$Q = (0.5G^{0.67}/D)V\Delta t_{lm} = 0.4LDG^{0.67}\Delta t_{lm} \quad (7.40)$$

where Q is in Btu/h, L is the dryer's length in ft, D is in ft, G is in lb (h ft² of cross section), and Δt_{lm} is the log mean of the drying gas wet-bulb depressions at the inlet and outlet of the dryer.

A different method for the estimation of heat transfer during drying can be obtained by dimensionless equations of the type $Nu = a' Re^{n'} Pr^{n'}$, which can be transformed as

$$j_H = a Re^n \quad (7.41)$$

thus the heat transfer factor correlates to Reynolds number. This analysis can be done for two reasons. First, it is a very simple expression and the Reynolds number is easy to be estimated in most cases, through three parameters that can be measured quite easily, the particles diameter, the velocity of the drying medium, and its temperature. The second reason is that using this equation, we calculate directly the heat transfer factor that is important when we use analogies among momentum, heat, and mass transport. The most common is the well-known Chilton–Colburn analogy (or simply Colburn analogy) that is based on empirical correlations, and not on mechanistic assumptions that are only approximations. Thus, it represents the experimental data extremely well over the range in which the empirical correlations are valid. This analogy stands for both laminar and tubular flows and for Prandtl and Schmidt numbers between 0.6 to 100 and 0.6 and 2500, respectively. The Chilton–Colburn analogy can be expressed as

$$j_H = j_D = \frac{f}{2} \quad (7.42)$$

where f is the friction factor, j_D is the mass transfer factor, and j_H is the heat transfer factor, given by the expression

$$j_H = St Pr^{2/3} = \frac{h}{\rho u_\infty C_p} Pr^{2/3} \quad (7.43)$$

TABLE 7.3
Constant of Equation 7.41 and the Corresponding
Reynolds Number Range for Some Products

Product/Reference	<i>a</i>	<i>n</i>	min <i>Re</i>	max <i>Re</i>
<i>Fish</i>				
Shene et al.	0.00160	−0.258	80	300
<i>Soya</i>				
Alvarez et al.	0.00960	−0.587	10	100
Shene et al.	0.00030	−0.258	20	80
<i>Sugar</i>				
Wang et al.	0.805	−0.528	1 500	17 000
Rotary	0.001	−0.161	10	300

where $St = h/\rho u_\infty C_p$ is the Stanton number, h is the heat transfer coefficient, ρ is the air density, u_∞ is the air velocity, and C_p is the specific heat of the air.

Note that the second equivalence, $f/2$, in Equation 7.42 stands only in the case of flow around relatively simple shapes, like flat surfaces or inside tubes.

Knowing the heat transfer factor we know the mass transfer factor, as well, and can calculate parameters concerning mass transfer, like the diffusion coefficient. This is important considering that rotary drying includes both heat and mass transfer, as the material receives heat and losses moisture, simultaneously.

Data retrieved from the literature for the drying of some materials in rotary dryers are shown in Table 7.3, which presents the constants of equation $j_H = aRe^n$, and the range in which it is valid. A general expression for the process is also given.

Figure 7.6 presents the heat transfer factor versus Reynolds number for rotary drying processes and various materials, Figure 7.7 shows the ranges of variation of the heat transfer factor versus Reynolds number for the rotary drying process in comparison with other thermal processes, and Figure 7.8 presents the estimated equation of heat transfer factor for the rotary drying versus Reynolds number, in comparison with other thermal processes.

The inlet gas temperature in a direct-heat rotary dryer is generally fixed by the heating medium, i.e., 400–450 K for steam and 800–1100 K for oil- and gas-fired burners. Lower temperatures should be used only if there are limitations by the shell's material. The exit gas temperature, which is a function of the economics involved, may be determined by the relationship

$$N_t = (t_1 - t_2)/(\Delta t)_m \quad (7.44)$$

where N_t is the number of heat transfer units based on the gas, t_1 is the initial gas temperature (K), and t_2 is the exit gas temperature (K), allowing for heat losses. The most economical operation of rotary dryers can be achieved for N_t in the range 1.5–2.5 as it has been found empirically.

The diameter of a rotary dryer may vary from less than 0.3 m to more than 3 m whereas the length-to-diameter ratio, L/D , is most efficient between 4 and 10 for industrial dryers. In a dryer design the value of N_t may change until the ratio mentioned above fall within these limits.

The volume of the dryer that is filled with material during operation is 10–15%. Lower fillage is insufficient to utilize the flights, while a greater one causes a short-circuit in the feed of solids across the top of the bed [35]

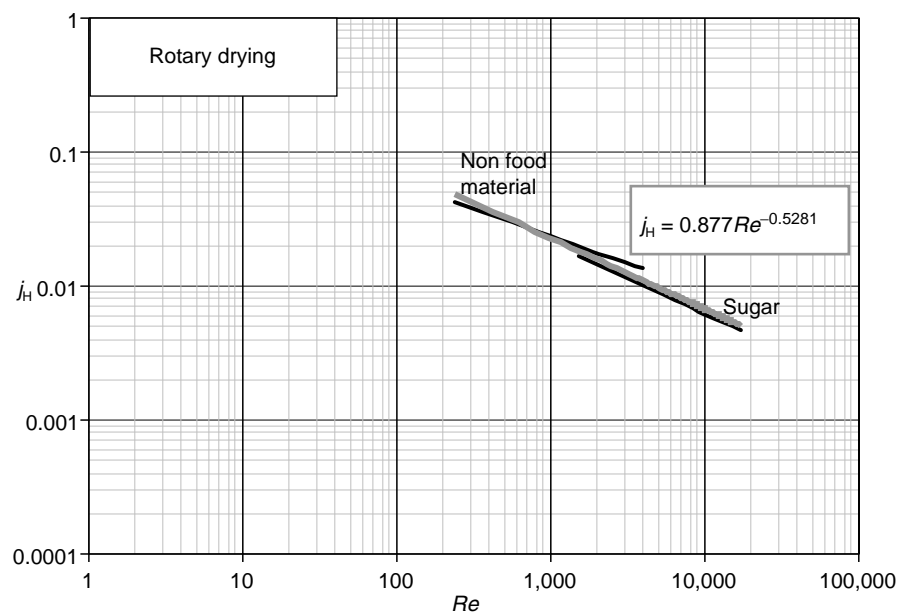


FIGURE 7.6 Heat transfer factor versus Reynolds number for the rotary drying process and various materials.

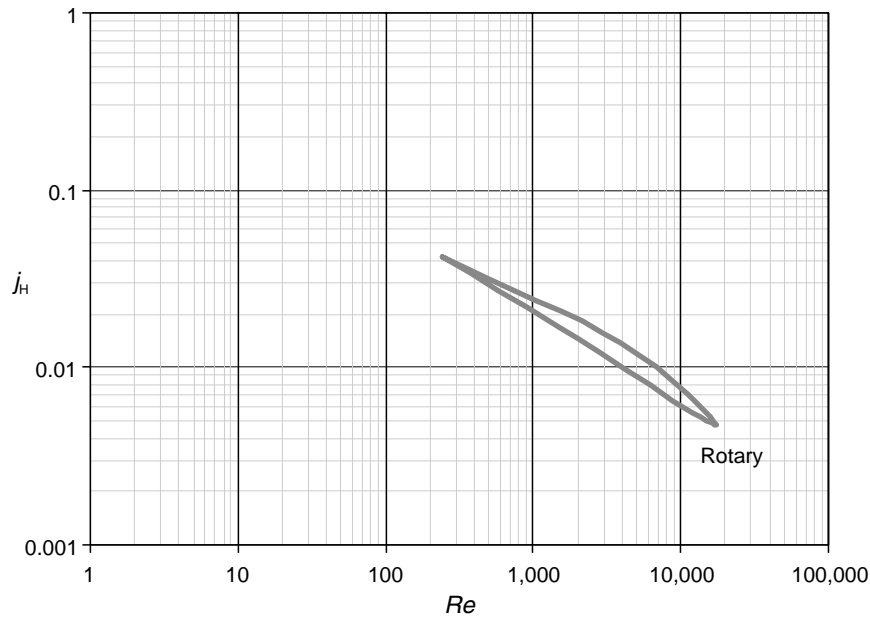


FIGURE 7.7 Ranges of variation of the heat transfer factor versus Reynolds number for the rotary drying process in comparison with other thermal processes.

$$\theta = \frac{0.23L}{SN^{0.9}D} \pm 0.6 \frac{BLG}{F} \quad (7.45)$$

$$B = 5(D_p)^{-0.5} \quad (7.46)$$

7.6 ENERGY AND COST ANALYSIS

The power required to drive a dryer with flights may be calculated by the following equation, proposed by the CE Raymond Division, Combustion Engineering Inc.,

$$\text{bhp} = \frac{N(4.75Dw + 0.1925D'W + 0.33W)}{100,000} \quad (7.47)$$

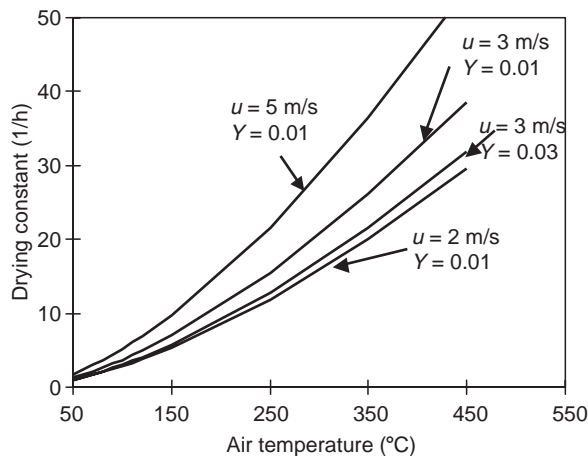


FIGURE 7.8 Typical drying curves.

where bhp is the break horsepower required (1 bhp = 0.75 kW), N is the rotational speed (r/min), D is the shell diameter (ft), w is the load of the material (lb), W is the total rotating load (equipment plus material) (lb), and D' is the riding-ring diameter (ft), which for estimating purposes can be considered as $D' = (D + 2)$.

The estimated cost of a steam-heated air rotary dryer, including auxiliary subsystems such as finned air heaters, transition piece, drive, product collector, fan and duct, ranges from about \$100,000 for a dryer size 1.219 m \times 7.62 m to \$320,000 for a dryer size 3.048 m \times 16.767 m. Their evaporation capacity is 136 and 861 kg/h, respectively, whereas they have a discharge value ranging from 408 kg/h for the smaller dryer to 2586 kg/h for the bigger one. In case that combustion chambers and fuel burner are required for operation at higher temperatures the cost is higher. The total installation cost that includes allocated building space, instrumentation, etc., is 150–300% of the purchase cost. Operating costs include fuel, power, and 5–10% of one worker's time, the yearly maintenance cost is 5–10% of the installation cost, and the power required for fans, dryer drive, and feed and product conveyors ranges from $0.5D^2$ to $1.0D^2$. The above prices are referring to carbon steel construction; when 304 stainless steel has to be used the prices are increased by about 50%.

High-temperature direct-heat rotary dryers present thermal efficiency in the range 55–75%, which is reduced to 30–55% for dryers that employ steam-heated air as heating medium.

7.7 A MODEL FOR THE OVERALL DESIGN OF ROTARY DRYERS

One way to estimate the time for the material to be dried is through the drying constant, k_M , that can be determined experimentally using an apparatus in which air passes through the drying material and air temperature, humidity, and velocity are controlled, while the material moisture content is monitored. A number of experiments have to be carried out for different temperatures, humidities, and velocities. The application of these methods proved that the drying constant depends on those parameters of the drying air and that it can be expressed as a function of them through a general equation of the type

$$k_M = f(T_A, Y_A, u_A) \quad (7.48)$$

A derived analytical correlation that can be produced by fitting the above equation to experimental data is given by the following equation:

$$k_M = k'_0 \left(\frac{T}{T_0} \right)^{k'_1} \left(\frac{Y}{Y_0} \right)^{k'_2} \left(\frac{u}{u_0} \right)^{k'_3} \quad (7.49)$$

where T_0 , Y_0 , u_0 are the parameters which express the mean values of the intervals of air temperature, humidity, and velocity that are used for the experiments, and k'_0 , k'_1 , k'_2 , and k'_3 are parameters. Figure 7.8 presents typical curves, which express the drying constant versus temperature for various air humidities and velocities.

Krokida et al. [28] proposed a model for the design of a rotary dryer, based on the estimation of the drying kinetics of the material that express data from laboratory experiments, and the calculation of residence time of the dryer from empirical equations. The dryer size and characteristics as well as the operating conditions can be calculated for given process

specifications by minimizing the total drying cost. The specifications include the solids feedrate F_s (kg/h db), and the inlet and outlet moisture content of the material, X_0 (kg/kg db) and X_s (kg/kg db). Among the characteristics of the dryer are its diameter D (m), the length-to-diameter ratio L/D , the total holdup-to-volume ratio H/V , the number of flights-to-diameter ratio n_f/D , and the slope of the cylindrical shell s (%). The drying conditions include the inlet temperature T_{AC} ($^{\circ}\text{C}$) and the gas velocity u (m/s) at temperature T_A ($^{\circ}\text{C}$).

A simplified diagram of the dryer is shown in Figure 7.9. The mathematical model of the process consists of two parts, the model of the burner and the model of the dryer.

7.7.1 BURNER

Assuming that the fuel is hydrocarbon with heat of combustion ΔH_f (kJ/kg) and fraction of hydrogen C_H (kg/kg) and that the combustion reactions are $\text{C} + \rightarrow \text{CO}_2$ and $\text{H}_2 + 1/2\text{O}_2 \rightarrow \text{H}_2\text{O}$, then 9 C_H kg of water vapor are produced per kg of fuel. Thus

$$R_w = 9C_H Z \quad (7.50)$$

where R_w is the production rate of water vapor (kg/h) and Z is the feedrate of fuel (kg/h).

The total and moisture balances over the burner are given by the following equations, which describe the combustion process:

$$F_{AC}(1 + Y_{AC}) = F_{AO}(1 + Y_O) + Z \quad (7.51)$$

$$F_{AC}Y_{AC} = F_{AO}Y_O + R_w \quad (7.52)$$

where F_{AO} and F_{AC} are the inlet and outlet flowrate of gases at the burner (kg/h db), and Y_O and Y_{AC} are

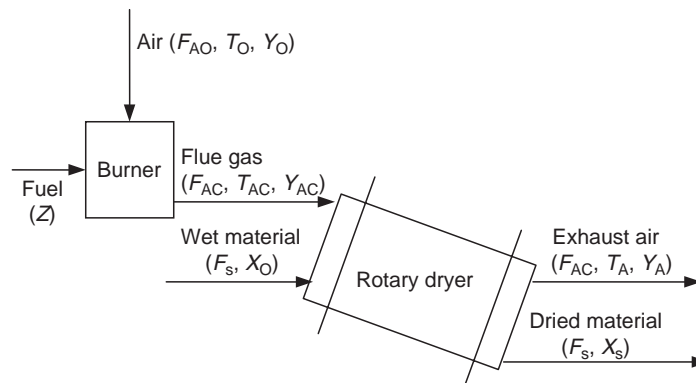


FIGURE 7.9 Simplified diagram of the dryer and burner constituting the drying process unit.

the inlet and outlet humidity of the gases at the burner (kg/kg db), respectively.

Assuming that the gases have the same thermophysical properties as air, the corresponding energy balance over the dryer is given by the following relationship:

$$F_{AC}(1 + Y_{AC})C_{PA}(T_{AC} - T_0) = Z\Delta H_f \quad (7.53)$$

where T_{AC} is the outlet gas temperature at the burner ($^{\circ}\text{C}$), T_0 is the ambient temperature ($^{\circ}\text{C}$), and C_{PA} is the specific heat of the gases (kJ/kg K).

7.7.2 DRYER

The following equations describe the mass and energy balances upon the dryer.

Mass balance on water

$$F_{AC}(Y_A - Y_{AC}) = F_S(X_O - X_S) \quad (7.54)$$

where Y_{AC} is the inlet humidity of the gases at the dryer (equal to the outlet humidity of the gases at the burner) (kg/kg db), and Y_A is the outlet humidity of the gases at the dryer (kg/kg db).

Energy balance (simplified)

$$F_{AC}C_{PA}(1 - Y_{AC})(T_{AC} - T_A) + F_S\Delta H_V(X_O - X_S) = 0 \quad (7.55)$$

where C_{PA} is the specific heat of the air–vapor mixture (kJ/kg K), ΔH_V is the latent heat of vaporization of water at the reference temperature (kJ/kg), and T_A is the mean air–vapor temperature at the dryer output.

7.7.3 DRYING KINETICS

The following well-known first-order kinetic model is selected to express the drying kinetics:

$$\frac{(X - X_{SE})}{(X_0 - X_{SE})} = \exp(-k_M t) \quad (7.56)$$

where X is the material moisture content (kg/kg db) after a time interval t (h), k_M is the drying constant (per hour), and X_{SE} is the equilibrium material moisture content.

The drying constant is a function of gas conditions and the following empirical equation can be used:

$$k_M(T, Y, u) = k'_0 T^{k'_1} Y^{k'_2} u^{k'_3} \quad (7.57)$$

where T , Y , u are the temperature ($^{\circ}\text{C}$), humidity (kg/kg db), and velocity of the drying gas, and k'_0 , k'_1 , k'_2 , and k'_3 are parameters, which express the effect of various factors on the drying constant.

The equilibrium moisture content of the solids, as a function of water activity and temperature of the surrounding air, can be calculated by the following correlation:

$$X_{SE} = b_1 \exp(b_2/T_A)[a_w/(1 - a_w)]^{b_3} \quad (7.58)$$

where a_w is the water activity of the gas stream and b_1 , b_2 , and b_3 are characteristic constants.

The absolute humidity of the drying airstream can be evaluated by the relationship

$$Y = m[a_w P_0(T_A)]/[P - a_w P_0(T_A)] \quad (7.59)$$

where $m = 0.622$ is the water-to-air molecular ratio and $P_0(T_A)$ is the water vapor pressure at temperature T_A .

The water vapor pressure at temperature T_A can be obtained from the Antoine equation

$$\ln P_0(T_A) = A_1 - A_2/(A_3 + T_A) \quad (7.60)$$

where A_1 , A_2 , and A_3 are constants.

7.7.4 RESIDENCE TIME

The residence time ($\bar{\tau}$) is defined by the equation

$$\bar{\tau} = M/F_S \quad (7.61)$$

where M is the total product mass in the dryer, which relates to the product holdup of the dryer (H) by the following expression

$$M = (1 - \varepsilon)\rho_p H \quad (7.62)$$

where ρ_p is the density of the material (kg/m³) and ε is its porosity.

Generally, the residence time in a rotating dryer is a function of its length, diameter, slope, and rotating velocity. An empirical equation can be used [24] for this correlation as follows:

$$\bar{\tau} = \frac{kL}{NDs} \quad (7.63)$$

where k is an empirical constant.

An empirical equation is also used by Kelly to correlate the total holdup to the flights load per unit length. This relationship underestimates the true

holdup value as it ignores the particles cascading through the gas. The equation can be written as

$$H = 0.5(n_f + 1)h_0L \quad (7.64)$$

where h_0 is the holdup per meter (m^2).

7.7.5 GEOMETRICAL CONSTRAINTS

The following geometrical constraints should be added to the mathematical model:

$$\begin{aligned} 5\% < H/V < 15\% \\ 2 < L/D < 20 \\ 5 < n_f/D < 10 \end{aligned}$$

Cost estimation

The process unit cost of wet product (\$/kg wb) has to be minimized

$$C_p = \frac{C_T}{t_{op}} F_S(1 + X_S) \quad (7.65)$$

where C_p is the cost of the product due to the drying process, t_{op} is the operating time per year (h/y), and C_T is the total annual cost of the drying process that can be expressed by the following equation:

$$C_T = eC_{eq} + C_{op} \quad (7.66)$$

where eC_{eq} is the yearly capital cost (\$/yr), C_{op} is the operating cost (\$/yr), and e is the capital recovery factor that is given by the equation

$$e = \frac{i(1+i)^N}{(1+i)^N - 1} \quad (7.67)$$

where i is the annual interest rate and N is the time of the loan (yr).

The equipment cost is affected by the size of the dryer and the consumption rate of fuel, assuming that a furnace is used for heat supply. Thus,

$$C_{eq} = \alpha_D A^{n_D} + \alpha_Z Z^{n_Z} \quad (7.68)$$

where α_D , α_Z are unit costs and n_D , n_Z are scaling factors for the dryer and burner, respectively.

The operating cost involves electrical energy and fuel cost:

$$C_{op} = h_p C_e t_{op} + Z C_Z t_{op} \quad (7.69)$$

where C_e and C_Z are the electricity and fuel cost, respectively.

The electrical power h_p for the cylinder rotation is given as follows (Kelly [24]):

$$h_p = qND(M + W') \quad (7.70)$$

where q is an empirical constant and W' is the dryer weight (kg).

The calculation of the dryer weight is based on its geometrical characteristics and is given by

$$W = \rho_M \left(\frac{2\pi D^2}{4} + \pi DL \right) dx \quad (7.71)$$

where dx is the dryer wall thickness (m), and ρ_M is the metal density (kg/m^3).

A degree of freedom analysis suggests that five design variables are available for the design problem described above. It can be proved that an effective solution algorithm can be based on the following selection of design variables: T_{AC} , u , H/V , L/D , and n_f/D , where the first and second express the operating conditions and the rest the dryer shape.

7.8 CASE STUDY 1

The solution of a typical dryer problem for an industrial olive cake rotary dryer is presented. The data required for process design calculations are given in Table 7.4. The results of calculations using the model proposed by McAdams [28] are presented in Table 7.5, and are obtained by minimizing the process unit cost, and evaluating the design variables.

A sensitivity analysis of the process unit cost is achieved by changing the two significant decision variables: the drying air temperature and the velocity. As the air-drying temperature is allowed to vary, air velocity is maintained constant and each time all other variables are calculated. It must be noted that as the air temperature increases and thus the operating cost increases, whereas the size of the equipment and, consequently, the cost of equipment decreases. For a given air velocity, the total cost reaches a maximum at a specific air temperature (see Figure 7.10). In Figure 7.11 the total unit cost is presented as function of air temperature for different air velocities.

The model was adapted to an industrial rotary dryer with the following characteristics: length 22 m, diameter 2.5 m, and number of flights 24.

The drying conditions are 650°C inlet drying air temperature, 2.4 m/s mean gas-vapor velocity, and the fuel consumption rate is 1500 kg/h. The operating conditions obtained from process design calculations are close to the real ones.

TABLE 7.4
Data for Process Design Calculations

<i>Process specifications</i>			
Solids flow rate	F_s	5000	kg/h
Input material moisture content	X_0	1.00	kg/kg db
Output material moisture content	X	0.10	kg/kg db
<i>Fresh air characteristics</i>			
Temperature	T_0	25	°C
Humidity	Y_0	0.01	kg/kg db
<i>Thermophysical properties</i>			
Water to air molar fraction	m	0.622	—
Air specific heat	C_{PA}	1.18	kJ/kg °C
Water specific heat	C_{PV}	1.98	kJ/kg °C
Heat of combustion	ΔH_f	15	MJ/kg
Latent heat of vaporization of water	ΔH_0	2500	kJ/kg
Porosity	ε	0.48	—
<i>Empirical constants</i>			
Empirical constant in Equation 7.14	k	0.003	—
Empirical constant in Equation 7.24	q	1	—
<i>Economic data</i>			
Dryer unit cost	a_D	8	k\$/m ²
Dryer scaling factor	n_D	0.62	—
Burner unit cost	a_Z	200	\$/kg
Burner scaling factor	n_Z	0.4	—
Lifetime	N	10	yr
Interest rate	i	8	%
Operating time	t_{op}	2000	h/yr
Electricity cost	C_e	0.07	\$/kW h
Fuel cost	C_z	0.05	\$/kg

7.9 CASE STUDY 2

For the drying of catalyst pellets, the engineers of a certain industry decided that a direct rotary dryer will be appropriate, and studied the performance of a pilot plant rotary dryer in order to obtain data for the scale-up. The production F will be 350 kg/h on a dry basis. The pellets have cylindrical shape, about 1 cm long and 1 cm in diameter, their bulk density ρ_b is 570 kg/m³, the specific heat C_{ps} is 1 kJ/kg K, and the initial moisture content X_0 , as a result of the previous unit operation, is 0.65 kg/kg db. The final product, in order to be stable, must have moisture content X no more than 0.05 kg/kg db. It is nonsticking, but it is sensitive at high temperatures. Therefore, cocurrent operation has to be used and the initial air temperature T_1 will not exceed the range of 150–170°C. The heating medium will be hot air. A steam-air heat exchanger is going to be used for the heating. The air velocity has to be limited to avoid entraining of the material by the air. Table 7.6 presents the values of the operating parameters of the pilot plant rotary dryer.

The following calculations aim at a preliminary design of the dryer.

TABLE 7.5
Results of Process Design Calculations

<i>Design variables</i>			
Input air temperature	T_{AC}	700	°C
Mean air–vapor velocity	u	2.4	m/s
Total holdup to volume fraction	H/V	15	%
Length-to-diameter fraction	L/D	20	—
Number of blades to diameter fraction	n_f/D	10	1/m
<i>Drying air characteristics</i>			
Mean air temperature	T_A	298	°C
Humidity outlet	Y	0.37	kg/kg db
<i>Operating characteristics</i>			
Residence time	t	0.3	h
Total holdup	H	8.4	m ³
Rotating velocity	N	8.6	rpm
<i>Dryer characteristics</i>			
Diameter	D	1.5	m
Length	L	30.6	m
Blade number	n_f	15	—
<i>Utilities</i>			
Fresh air flow rate	F_{A0}	15,048	kg/h
Fuel rate	Z	1066	kg/h
<i>Economics</i>			
Electricity cost	C_e	6286	\$/yr
Fuel cost	C_z	106,606	\$/yr
<i>Operating cost</i>			
Cost of equipment	C_{eq}	55,619	\$/yr
Total cost	C_T	168,510	\$/yr
Unit cost	C_p	0.00843	\$/kg wb

The overall material mass (kg/h) that is fed is

$$F_1 = F(1 + X_0) \quad (7.72)$$

whereas the mass (kg/h), which exits the dryer is

$$F_2 = F(1 + X) \quad (7.73)$$

Therefore, the evaporating water m_w (kg/h) is

$$m_w = F_1 - F_2 \quad (7.74)$$

The heat supplied by the hot air is used for five different operations:

1. To evaporate the water, that leaves the material

$$Q_1 = m_w \Delta H_w \quad (7.75)$$

2. To heat the vapor from the initial wet-bulb temperature of the air to the exit air temperature

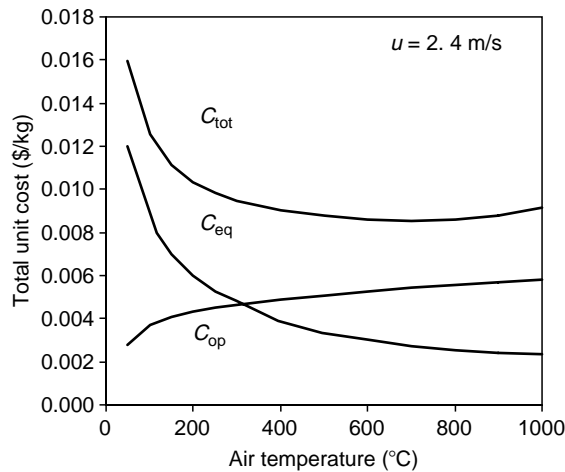


FIGURE 7.10 Total unit cost versus air temperature for air velocity 2.4 m/s.

$$Q_2 = m_w C_{pv}(T_2 - T_w) \quad (7.76)$$

3. To heat the water that evaporates, from its initial temperature, as it enters the dryer, to the inlet wet-bulb temperature of the air, in order to evaporate

$$Q_3 = m_w C_{pw}(T_w - T_{m1}) \quad (7.77)$$

4. To heat the dry solid from its inlet temperature to its exit temperature

$$Q_4 = FC_{ps}(T_{m2} - T_{m1}) \quad (7.78)$$

5. To heat the water that remains in the final product from the inlet to the exit temperature of the material

$$Q_5 = FXC_{pw}(T_{m2} - T_{m1}) \quad (7.79)$$

where ΔH_w is the latent heat of vaporization (kJ/kg), C_{pv} , C_{pw} , C_{ps} , are the specific heat of vapor, water,

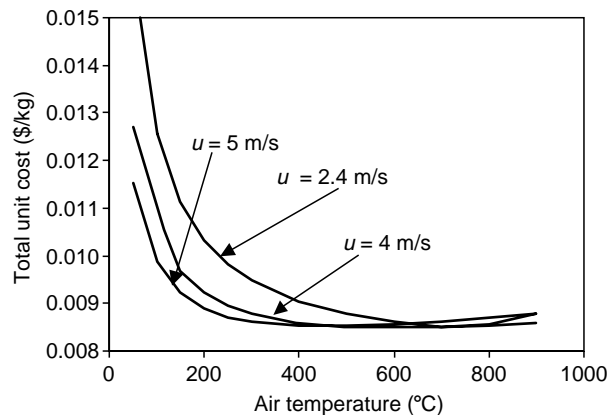


FIGURE 7.11 Total unit cost versus air temperature for three different air velocities.

TABLE 7.6
Data Obtained by the Pilot Plant Dryer

Inlet temperature of drying air	T_1	160	°C
Exit temperature of drying air	T_2	65	°C
Wet-bulb temperature of inlet air	T_w	40	°C
Exit temperature of product	T_2	45	°C
Permittable air mass velocity	u_{perm}	3	kg/m ² s
Retention time of product	τ	0.35	h

and solid (kJ/kg °C), respectively, T_w is the inlet wet-bulb temperature of the drying air, T_2 is the outlet temperature of the air (°C), and T_{m1} , T_{m2} are the inlet and exit temperature (°C) of the material (dry solid and moisture content), respectively.

The overall heat transferred to the product is given by the correlation

$$Q = (1 + \alpha)(Q_1 + Q_2 + Q_3 + Q_4 + Q_5) \quad (7.80)$$

where α is a factor that represents the heat losses due to the conduction between the outer surface of the dryer and the atmospheric air and especially, because of radiation. These losses are estimated to be about 7.5–10% of the heat consumption for the reasons mentioned above. The largest amount of heat is used for the evaporation of moisture content and is expressed by the ratio

$$\beta = Q_1/Q \quad (7.81)$$

The air mass rate G required in order to transfer sufficient amount of heat for the drying is

$$G = \frac{Q}{C_{p,air}(T_1 - T_2)} \quad (7.82)$$

where T_1 is the inlet air temperature (°C) and $C_{p,air}$ is the specific heat of air (kJ/kg °C).

For the estimation of the diameter D of the dryer (m) two points have to be examined. First it must be large enough so that the air mass velocity u (kg/m²s) will not exceed the value that causes entrainment of the product, and second we must assume that only a percentage of the dryer cross section represents a free area for the air to pass. This percentage is about 85% ($j = 0.85$), as can be estimated by operating rotary dryers. Therefore the diameter of the cylindrical shell is calculated by the following equation:

$$D = \sqrt{\frac{4G}{3600\pi ju}} \quad (7.83)$$

where 3600 is a factor for the arrangement of the units.

The humidity of the exit air should be checked for not exceeding the maximum mass of vapor the air can hold under the specific condition on the exit (for %RH = 100). The initial air humidity Y_1 is about 0.01 kg/kg dry air (for $T_1 = 160^\circ\text{C}$ and $T_w = 40^\circ\text{C}$). The humidity of the exit air Y_2 is

$$Y_2 = Y_1 + \frac{m_w}{G} \quad (7.84)$$

The volume V of the dryer (m^3) is calculated by the expression

$$V = \frac{\tau F_2}{H \rho_s} \quad (7.85)$$

where τ is the retention time of the product (h), and H is the dryer holdup, that is assumed to be about 0.07–0.08 of the dryer volume, as values in this range give good performance in industrial dryers. The retention time could be calculated by the geometric characteristics of the dryer, but it is desirable to obtain it by experiments rather than through theoretical calculations. In this case study it is estimated on the basis of pilot plant data, and the volume is calculated by the above expression. The length L of the dryer (m) is given by the correlation

$$L = \frac{4V}{\pi D^2} \quad (7.86)$$

In practice the ratio L/D should be within the range 4 to 10, for optimum performance. The number of heat transfer units N_T is defined by the equation

$$N_T = \ln \frac{T_1 - T_w}{T_2 - T_w} \quad (7.87)$$

and should be in the range 1.5 to 2.5 [35]. These ranges have been estimated through practical experience by the study of industrial direct rotary dryers in order for efficient operation to be achieved.

For the heating of the air, a steam-air heat exchanger is to be used. Its energy load should be sufficient for the heating of the airstream from the initial atmospheric temperature T_0 ($^\circ\text{C}$) to the inlet air temperature in the dryer T_1 , and given by the equation

$$Q_{\text{he}} = G C_{p,\text{air}} (T_1 - T_0) \quad (7.88)$$

Steam at temperature T_{st} ($^\circ\text{C}$) will be used as heating medium in the exchanger. The consumption of steam is

$$F_{\text{st}} = \frac{Q_{\text{he}}}{\Delta H_{\text{st}}} \quad (7.89)$$

TABLE 7.7
Data for Process Design Calculations

<i>Product specifications</i>			
Production rate (dry basis)	F	350	Kg/h db
Initial moisture content	X_0	0.65	kg/kg db
Final moisture content	X	0.05	kg/kg db
Inlet product temperature	T_{m1}	25	$^\circ\text{C}$
<i>Thermophysical properties</i>			
Evaporation heat of water	ΔH_w	2350	kJ/kg
Specific heat of product	C_{ps}	1.0	kJ/kg $^\circ\text{C}$
Specific heat of water	C_{pw}	4.18	kJ/kg $^\circ\text{C}$
Specific heat of vapor	C_{pv}	1.88	kJ/kg $^\circ\text{C}$
Specific heat of air	$C_{p,\text{air}}$	1.01	kJ/kg $^\circ\text{C}$
Bulk density	ρ_b	570	kg/m ³
<i>Properties of air</i>			
Atmospheric air temperature	T_0	15	$^\circ\text{C}$
Humidity of inlet air	Y_1	0.01	kg/kg db
<i>Constants</i>			
Dryer holdup	H	0.075	—
Factor	α	0.1	—
Factor	j	0.85	—

The thermal efficiency of the dryer is

$$n_{\text{th}} = \frac{Q_1 + Q_2 + Q_3 + Q_4 + Q_5}{Q_{\text{he}}} \quad (7.90)$$

Table 7.7 presents the specifications, thermophysical properties, and factor j for the design of the dryer, and Table 7.8 shows the values of the parameters calculated by the above equations.

TABLE 7.8
Results of Process Design Calculations

Overall inlet material	F_1	578	kg/h
Overall exit material	F_2	368	kg/h
Evaporating water	m_w	210	kg/h
Overall heat consumption	Q	577,500	kJ/h
Heat for evaporation	Q_1	493,500	kJ/h
Heat for vapor	Q_2	9,870	kJ/h
Heat for liquid	Q_3	13,167	kJ/h
Heat for product solid	Q_4	7,000	kJ/h
Heat for product water	Q_5	1,463	kJ/h
Air mass rate	G	6,019	kg/h
Diameter	D	0.9	m
Volume	V	3.0	m ³
Length	L	4.6	m
Number of heat transfer units	N_T	1.6	—
Heat load of exchanger	Q_{he}	881,447	kJ/h
Heat consumption	F_{st}	316	kg/h
Thermal efficiency	n_{th}	0.60	—

7.10 CONCLUSION

Until recently, the design of the industrial dryers was based on the experience of manufacturers and suppliers of these units, who used both data obtained in pilot plant rotary dryers and operating characteristics from units already installed. Because of the variety in drying equipment and solid materials that are processed, little consideration was given to mathematical models and theoretical approaches. It was common practice for the dryer to be built a bit oversized and inefficient, but mechanically sound and well proven in operation, instead of an optimization process to be followed, even if the capital and operating costs were larger. In recent years, many models and simulation techniques have been published, which can be useful for the design of dryers, especially when the drying material is the same or similar to the one the model refers to. Nevertheless, the development of a universal model of the rotary dryer, that combines the cascading motion of the particles with the heat and mass transfer, is questionable. The mathematical expressions and models that have been described can promote the understanding of the individual processes that take place during drying and the particular effect of each design parameter to the drying process.

NOMENCLATURE

A_1	constant in Equation 7.60
A_2	constant in Equation 7.60
A_3	constant in Equation 7.60
α_D	unit cost for the dryer, \$/m ²
α_Z	unit cost for the burner, \$/kg
b	constant in Equation 7.29
b'	constant in Equation 7.29
bhp	break horsepower
b_1	constant in Equation 7.58
b_2	constant in Equation 7.58
b_3	constant in Equation 7.58
C_e	electricity cost, \$/kW h
C_{eq}	equipment cost, \$
C_H	hydrogen mass fraction, kg/kg
C_{op}	operating cost, \$/yr
C_P	process unit cost, \$/kg wb
C_T	total annual cost, \$/yr
C_Z	fuel cost (\$/kg)
D	drum inside diameter, m
D'	riding-ring diameter, ft
D_e	effective drum diameter, m
d_p	particle diameter, μ m
dx	dryer wall thickness, m
e	capital recovery factor
F	solids mass feedrate, kg/s

f	drag coefficient, or friction factor
$f()$	function of parameter indicated in parentheses
G	gas flowrate per unit area of dryer cross section, kg/hm ²
g	acceleration due to gravity, m/s ²
H	actual volumetric holdup of drum, m ³
H^*	design volumetric holdup of drum, m ³
h^*	design volumetric holdup of solids per unit length of flight, m ³ /m
h_0^*	value of h^* at $\theta = 0^\circ$
h_0	actual volumetric holdup in the flight of solids per unit length of flight at $\theta = 0^\circ$, m ³ /m
h_p	electrical power, kW
i	annual interest rate
j	percentage of the dryer cross section represents a free area for the air to pass
j_D	mass transfer factor
j_H	heat transfer factor
K	constant in Equation 7.22
K	proportional constant in Equation 7.37
K'	constant in Equation 7.24
K_s	constant in Equation 7.36
k	dimensionless constant in Equation 7.20
k	empirical constant in Equation 7.63
k'	constant
k_c	constant
k_M	drying constant, per hour
k_0'	constant in Equation 7.49
k_1'	constant in Equation 7.49
k_2'	constant in Equation 7.49
k_3'	constant in Equation 7.49
L	drum length, m
l	radial flight depth, m
l'	height of flight lip, m
L_{eff}	length of drum over which particle travels by cascade motion, m
M	ratio of actual drum holdup to the design holdup
M	total material mass in the dryer, kg
m	constant in Equation 7.20
m	constant in Equation 7.59
m'	empirical constant in Equation 7.23
m_0	ratio of actual flight holdup to the design holdup at $\theta = 0^\circ$
m_w	rate of water vaporization, kg/h
N	period of the loan, y
N	rotational speed, rpm
N_t	number of heat transfer units based upon the gas dynamic angle of repose of the solids (degrees)
n	constant in Equation 7.37 and in Equation 7.41
n'	constant
n_D	scaling factor for the dryer
n_f	number of flights
n_{th}	thermal efficiency
n_Z	scaling factor for the burner

P_0	water vapor pressure, kPa
Q	rate of heat transfer between gas and solids, J/s
Q_i	partial rates of heat transfer in Case Study 2, kJ/h
q	empirical constant in Equation 7.70
r	drum inside radius, m
r_e	effective drum radius
R_w	production rate of water vapor in burner, kg/h
s	slope of the cylindrical shell, %
T_0	ambient temperature, °C
T	temperature, °C
t_f	average falling time of the particles
t_{op}	annual operating time, h/yr
U_L	overall heat transfer coefficient based on dryer length
U_v	overall heat transfer coefficient based on dryer volume
u	gas velocity, m/s
u_{perm}	permissible air mass velocity, kg/m ² s
u_r	relative velocity between particles and gas
V	drum volume, m ³
w	load of the material, lb
W	total rotating load, lb
W'	dryer weight, kg
X	material moisture content, kg/kg
X_0	solids holdup expressed as percentage of drum volume without airflow
X_a	solids holdup expressed as percentage of drum volume with airflow
Y	absolute humidity of air, kg/kg db
Y	particle length of fall, m
\bar{Y}	average particle length of fall, m
Z	parameter in Equation 7.25
Z	flowrate of fuel, kg/h

GREEK SYMBOLS

a	ratio of the effective area to the volume of the dryer
α	constant in Equation 7.80 and Equation 7.41
α	drum slope (degrees)
α'	constant
a_w	water activity of airstreams
β	angle subtended by flight at center of the drum
β	factor in Equation 7.81
γ	dynamic coefficient of friction of particle
ΔH_f	heat of combustion of fuel, kJ/kg
ΔH_v	latent heat of vaporization of water, kJ/kg
$(\Delta T)_m$	true mean temperature difference between the hot gases and the material, °C
ΔT_{lm}	logarithmic mean temperature difference between the wet-bulb depressions of the drying air at the inlet and outlet of the dryer, K
ε	porosity

θ	angle subtended by flight lip with horizontal at center of the drum
$\bar{\theta}$	angle to which average particle is carried in flights before cascading
θ_i	angular spacing between flights
ν	ratio of centrifugal to gravitational forces acting on particle
ρ	air (or gas) density, kg/m ³
ρ_b	bulk density of solids, kg/m ³
ρ_M	metal density, kg/m ³
ρ_p	particles density, kg/m ³
σ	parameter in Equation 7.24
$\bar{\tau}$	residence time of average particle (or, time of passage)
φ	angle between horizontal and free surface of solids
φ_0	angle between horizontal and free surface of solids at $\theta = 0^\circ$
φ_c	centrifugal force on particle, N
φ_f	frictional force on particle, N
φ_g	gravitational force on particle, N
φ_n	normal reaction of surface on particle
ω	angular speed of drum

SUBSCRIPTS

a	gas
air	air
he	heat exchanger
s	solid
st	steam
AC	drying airstream
A0	fresh airstream
SE	at equilibrium
v	vapor
w	wet-bulb
1	inlet
2	outlet

DIMENSIONLESS NUMBERS

Nu	Nusselt number
Pr	Prandtl number
Re	Reynolds number
St	Stanton number

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8 Fluidized Bed Dryers

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8.1 INTRODUCTION

Fluidized bed dryers (FBD) are used extensively for the drying of wet particulate and granular materials that can be fluidized, and even slurries, pastes, and suspensions that can be fluidized in beds of inert solids. They are commonly used in processing many products such as chemicals, carbohydrates, foodstuff, biomaterials, beverage products, ceramics, pharmaceuticals in powder or agglomerated form, health-care products, pesticides and agrochemicals, dyestuffs and pigments, detergents and surface-active agents, fertilizers, polymer and resins, tannins, products for calcination, combustion, incineration, waste management processes, and environmental protection processes. Fluidized bed operation gives important advantages such as good solids mixing, high rates of heat and mass transfer, and easy material transport.

For drying of powders in the particle size range of 50 to 2000 μm , fluidized beds compete successfully with other more traditional dryer types, e.g., rotary, tunnel, conveyor, continuous tray (see Table 8.1).

Conventional fluidized bed is formed by passing a gas stream from the bottom of a bed of particulate solids. At low gas velocities the bed is static (packed).

The bed of particles rests on a gas distributor plate. The fluidizing gas passes through the distributor and it is uniformly distributed across the bed. Pressure drop across the bed increases as the fluidizing gas velocity is increased. At a certain gas velocity, the bed is fluidized when the gas stream totally supports the weight of the whole bed. This state is known as minimum fluidization and the corresponding gas velocity is called minimum fluidization velocity, u_{mf} . Pressure drop across the bed remains nearly the same as pressure drop at minimum fluidization even if the gas velocity is increased further. Figure 8.1 shows various regimes of the particulate bed from packed to bubbling bed when the gas velocity is increased. The graphs show the bed pressure drops and bed voidage under various regimes.

A fluidized bed is operated at superficial gas velocities higher than the minimum fluidization velocity, u_{mf} , normally at 2–4 u_{mf} . The minimum fluidization velocity is typically obtained from experiments. There are several ways to determine the minimum fluidization velocity experimentally. It can also be estimated using various correlations. A list of minimum fluidization velocity can be obtained from Gupta and Sathiyamoorthy [1]. It should be noted that these correlations have limitations such as

TABLE 8.1
Comparison of Fluidized Bed Dryers (Conventional Types and Modified Types) with Other Competing Dryers for Particulate Solids

Criterion	Rotary	Flash ^a	Conveyor	Conventional FBDs	Modified FBDs
Particle size	Large range	Fine particles	500 μm –10 mm	100–2000 μm	10 μm –10 mm
Particle size distribution	Flexible	Limited size distribution	Flexible	Limited size distribution	Wide distribution
Drying time (approx.)	Up to 60 min	10–30 s	Up to 120 min	Up to 60 min	Up to 60 min
Floor area	Large	Large length	Large	Small	Small
Turndown ratio	Large	Small	Small	Small	Small
Attrition	High	High	Low	High	High
Power consumption	High	Low	Low	Medium	Medium
Maintenance	High	Medium	Medium	Medium	Medium
Energy efficiency	Medium	Medium	High	High	High
Ease of control	Low	Medium	High	High	High
Capacity	High	Medium	Medium	Medium	High

^aFlash dryer is used only for removing surface moisture from smaller particles at relatively short drying times typically in the range of 10–30 s.

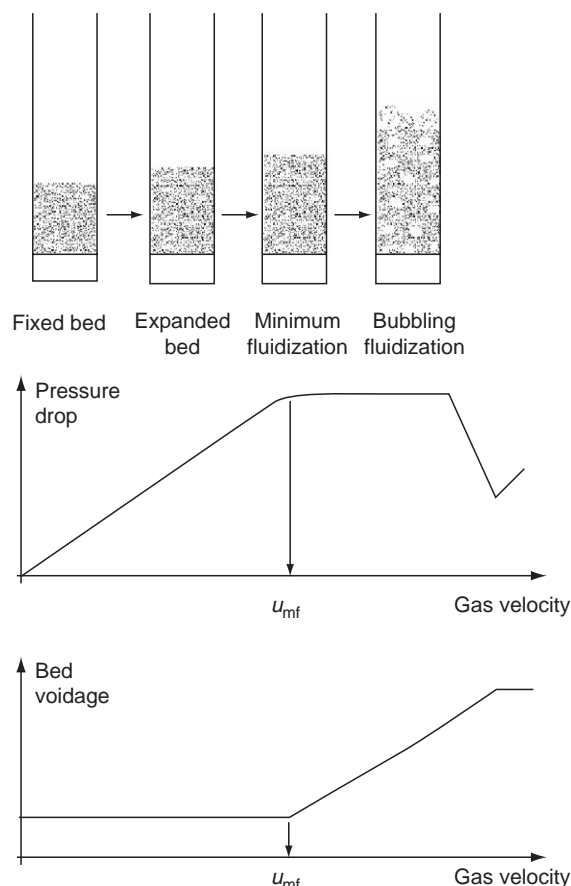


FIGURE 8.1 Various regimes of a bed of particles at different gas velocities.

particle size, column dimensions, operating parameters, etc. Thus, they are valid in a certain range of criteria and operating conditions. The effect of wetness of the particles is, however, not included.

Particles with high initial moisture content require a higher minimum fluidization velocity than similar bed of dry particles. Due to dominant cohesive forces exerted by wetted surfaces, only the top layer of the bed of solids is fluidized. The bottom layers may remain stationary during the initial stage of drying when the solids are quite wet.

For the case of dry (or partially dry, no surface moisture) particles, if the fluidizing gas is further increased, the bed of particles goes through different types of fluidization regimes depending on the types of particles with reference to the Geldart classification of powders [2,3]. Based on fluidization quality, powders can be classified into four groups: group A (aeratable particles, easy-to-fluidize when dry), group B (sandlike particles, easy-to-fluidize when dry), group C (fine and ultrafine particles, difficult-to-fluidize due to dominated cohesive forces between particles), and group D (large and dense particles, poor fluidization

quality due to formation of large bubbles in the bed). Figure 8.2 shows the various fluidization regimes exhibited by a bed of dry particles of different classes with increasing gas velocity. Fluidized bed dryers are normally operated in the regimes of smooth and bubbling fluidization.

After passing through the fluidized bed, the gas stream is introduced into gas-cleaning systems to separate fine particles (dusts) from the exit gas stream before discharging it to the atmosphere. Figure 8.3 shows a typical setup of fluidized bed drying system. A typical fluidized bed drying system consists of a gas blower, heater, fluidized bed column, gas-cleaning systems such as cyclone, bag filters, precipitator, and scrubber. To save energy, sometimes the exit gas is partially recycled.

The bubbling fluidized bed (Figure 8.3) is divided vertically into two zones, namely a dense phase and a freeboard region (also known as lean phase or dispersed phase). The dense phase is located at the bottom; above the dense phase is the freeboard in which the solids hold-up and density decreases with height (Figure 8.3).

Fluidizing gas after passing through the bed of particles enters the freeboard region, and carries with it fine particles which are terminal velocities smaller than the operating gas velocity. This phenomenon is known as elutriation. Solids hold-up in the freeboard region decreases as the freeboard height is increased until a height beyond which the solids hold-up remains unchanged. This point is known as the transport disengagement height (TDH). TDH can be estimated from several empirical correlations; these correlations are expressed in terms of one or two operating parameters thus, the predictions are generally poor. However, there is no universally accepted equation for calculating TDH. As a result, it is best to determine the transport disengaging height experimentally.

In designing a fluidized bed dryer for solids drying, it is important to take note about the occurrence of entrainment of fine particles, especially if the solids are polydispersed (i.e., have wide particle size distribution). The gas exit should be placed at a height above the TDH to minimize elutriation of fines.

On the other hand, by means of fines elutriation, solids in fluidized bed can be classified into fine and coarse products. Particles that are elutriated by the fluidized gas stream are known as fine products whereas particles retained in the bed are known as coarse products. This process is called fluidized bed separation or classification or dedusting. For processes that require a certain degree of dedusting (removal of undesirable fine particles) or classification, operating gas velocity and location of gas exit should

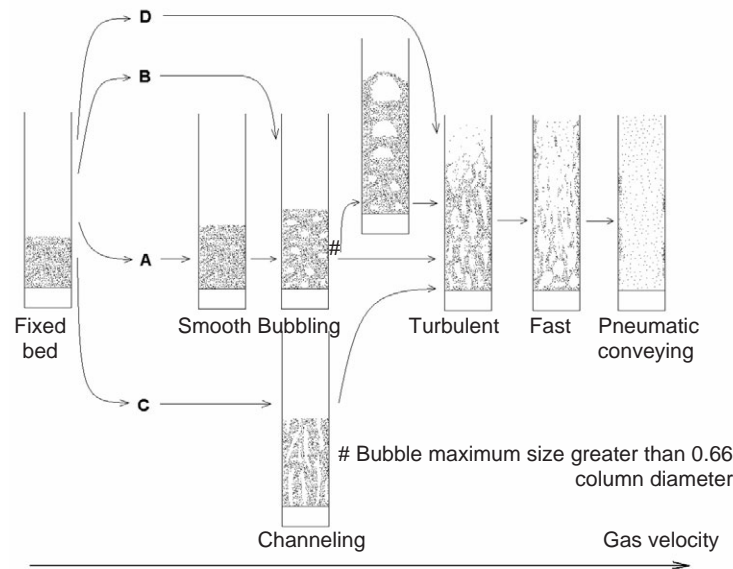


FIGURE 8.2 Various fluidization regimes exhibited by different classes of particles with increasing gas velocity.

be chosen carefully in order to achieve the appropriate product cut size. Cut size refers to the critical size that separates the fine (elutriated) and coarse (remain in bed) particles.

To ensure uniform and stable fluidization, the type of distributor has to be chosen carefully. This is to prevent poor fluidization quality of solids in certain regions in the fluidized bed, to prevent plugging of distributor-perforated holes, and to avoid solids from dropping into windbox or gas plenum located beneath the fluidized bed. There are many types of distributors available. Figure 8.3 (lower right image) shows four common types of distributors, namely, ordinary (i), sandwiched (ii), bubble cap tuyere (iii), and sparger (iv). It should be noted that pressure drop

across the distributor must be high enough to ensure good and uniform fluidization.

As a rule of thumb, for upwardly and laterally directed flow, pressure drop across the distributor must exceed 30% of the pressure drop across the bed [4]. Whereas for downwardly directed flow, the pressure across the distributor must be greater than 10% of the pressure drop across the bed. Upwardly directed flow is normally found in ordinary perforated plates (Figure 8.3, lower right image-i). Sandwich-type distributor is used if reinforcement of the distributor is needed due to heavy load of bed of particles (Figure 8.3 lower right image-ii). Laterally directed flow is normally obtained with bubble caps and nozzle types of distributors (Figure 8.3, lower

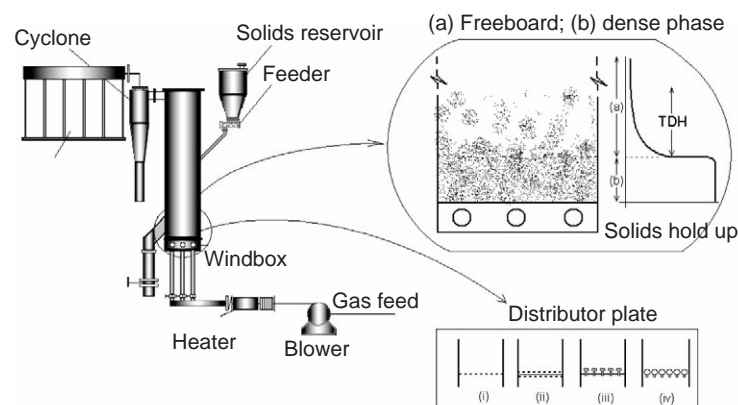


FIGURE 8.3 Typical fluidized bed drying setup. Zones in a fluidized bed with its corresponding solids hold-up are shown in upper right side image. Types of perforated distributor plates that can be used are shown in lower right side image.

right image-iii), whereas the sparger type gives laterally or downwardly directed flow (Figure 8.3, lower right image-iv).

8.2 ADVANTAGES AND LIMITATIONS OF FLUIDIZED BED DRYERS

Commonly recognized advantages of fluidized bed drying include: high rate of moisture removal, high thermal efficiency, easy material transport inside dryer, ease of control, and low maintenance cost. Limitations of fluidized bed dryer include: high pressure drop, high electrical power consumption, poor fluidization quality of some particulate products, nonuniform product quality for certain types of fluidized bed dryers, erosion of pipes and vessels, entrainment of fine particles, attrition or pulverization of particles, agglomeration of fine particles, etc. See Mujumdar and Devahastin [5] for detailed discussion.

Besides drying, fluidized bed has found wide ranges of industrial applications in various industries for mixing, dedusting, granulation, coating, agglomeration, cooling, chemical reactions, incineration, combustion, gasification, etc. Many of these processes can be incorporated with fluidized bed drying in one unit processor to accomplish two or more processes in the same unit. Processes that can be advantageously incorporated with fluidized bed drying are described briefly in the following paragraphs.

The mixing effect in a fluidized bed is generally good for particle sizes between 50 and 2000 μm . For fine particles (particle size less than 50 μm), or for particles that are difficult-to-fluidize when wet, vibration is normally applied to improve the fluidization quality and the mixing effect. For large particles, insertion of internals or use of the spouting mode can help to improve the operation. For fluidized bed drying, good particle mixing is essential. Thus, knowledge on particle fluidization characteristics and their properties is required to ensure good performance of a fluidized bed dryer. In addition, the bed of particles can be fluidized by a pulsating flow or by fluidizing sections of the bed periodically such that the entire bed is fluidized in sequence once over a cycle. Clearly, this operation results in saving of drying air and hence electrical power but it also leads to a longer operating time due to the intermittent mode of heat input. Besides, intermittent fluidization can reduce problem of mechanical damage to the particles due to continuous vigorous particle-particle collision as well as attrition-induced dusting.

Spray drying, granulation, coating, and agglomeration share the same basic operating principle. A fine spray of solution-paste-slurry-suspension is

atomized and sprayed in the fluidized bed of the drying material itself or inert particles, which are already loaded in the drying chamber. Formation and growth of solid particles takes place in the chamber as evaporation and drying carry away moisture. In granulation, growth of solid particles is carried out by successive wetting and coating of liquid feed onto the solid particles, and solidification of the coated layer by hot drying air. In coating, a layer of expensive active agent can be coated on a less expensive substrate, or to add a surface agent on solid particles, which is needed for downstream processing. By spraying a suitable binder onto the bed of solid particles, agglomerated or granulated solid particles of large particle size are produced.

In most cases, spray drying alone is not energy efficient to remove all moisture content inside the solids. This is because considerable amount of heat and time is needed to remove internal moisture that is trapped inside the solids internal. Fluidized bed drying can be incorporated as the second-stage drying to remove the internal moisture. This can be followed by a third-stage fluidized bed cooling to avoid the condensation problem during packaging in some applications.

8.3 HEAT TRANSFER IN FLUIDIZED BEDS

Heat transfer in gas-fluidized bed can occur by conduction, convection, and radiation depending on the operating conditions. The contribution of the respective modes of heat transfer to the coefficient of heat transfer depends on particle classification, flow condition, fluidization regimes, type of distributor, operating temperature, and pressure. Heat transfer between a single particle and gas phase can be defined by the conventional equation of heat transfer:

$$q = h_p A_p (T_p - T_g) \quad (8.1)$$

where q is the rate of heat transfer (W), h_p is the heat transfer coefficient ($\text{W}/(\text{m}^2\text{K})$), A_p is the surface area of a single particle (m^2), T_p is the temperature of the particle (K), and T_g is the temperature of gas (K).

The value of heat transfer coefficient of a single particle in a fluidized bed system is generally not high. It is in the range of 1 to 700 $\text{W}/(\text{m}^2\text{K})$. However, due to the large interfacial surface area, in the order of 3,000 to 45,000 m^2/m^3 , extremely high rates of heat transfer are achieved in this system. The heat capacity is in the order of $10^6 \text{ J}/(\text{m}^3\text{K})$. As a result, thermal equilibrium is reached quickly. In designing fluidized bed dryers, an isothermal condition is often assumed.

The heat transfer coefficient, h_p , is a function of the operating parameters, particulate characteristics,

and dryer geometry. It can be estimated from the following correlations depending on the particle Reynolds number, Re_p :

$$h_p = \frac{k_g}{d_p} Nu_p \quad (8.2)$$

where k_g is the gas thermal conductivity (W/(m K)), d_p is the particle diameter (m), and Nu_p is the particle Nusselt number, and Pr_g is the gas Prandtl number [6]. For $0.1 \leq Re_p \leq 50$, $Nu_p = 0.0282 Re_p^{1.4} Pr_g^{0.33}$ and for $50 \leq Re_p \leq 1 \times 10^4$, $Nu_p = 1.01 Re_p^{0.48} Pr_g^{0.33}$. Tubes, single or multiple, as well as flat channels can be immersed in a fluidized bed to provide additional heat for drying by conduction. These surfaces may be vertically or horizontally oriented. Empirical correlations are available in the literature for various geometries and operating conditions.

The surface-to-bed heat transfer coefficient, $h_w = q/a_w (T_b - T_w)$, is based on the surface area of the submerged object. This coefficient consists of two components, convective and radiative if the temperature is high. Here a_w is wall surface area (m^2) and T_w is wall temperature (K), T_b is bed temperature (K).

The convective heat transfer coefficient, h_c , can be estimated using correlation by Vreedenberg [7] for horizontal immersed objects:

$$\frac{h_c d_t}{k_g} = 420 \left(\frac{\rho_s}{\rho_g} Pr_g \frac{\mu_g^2}{g \rho_s^2 d_p^3} \right)^{0.3} Re_t^{0.3} \quad \text{if } \frac{\rho_s}{\rho_g} Re_p \geq 2550 \quad (8.3)$$

$$\frac{h_c d_t}{k_g} = 0.66 Pr_g^{0.3} \left(\frac{\rho_s (1 - \varepsilon)}{\rho_g \varepsilon} \right)^{0.44} Re_t^{0.44} \quad \text{if } \frac{\rho_s}{\rho_g} Re_p \leq 2050 \quad (8.4)$$

In these equations, d_t is the column diameter (m), ρ_s is the particle density (kg/m^3), ρ_g is the gravitational acceleration (m/s^2), μ_g is the gas viscosity (Ns/m^2), ε is the void fraction, and Re is the Reynolds number defined by

$$Re_t = \frac{d_t \rho_g u_g}{\mu_g} \quad (8.5)$$

and

$$Re_p = \frac{d_p \rho_g u_g}{\mu_g} \quad (8.6)$$

The radiant heat transfer coefficient, h_r ($W/m^2 K$) can be estimated using the following equation among others [8]:

$$h_r = \left(\frac{e_b e_w}{e_b + e_w - e_b e_w} \right) \frac{\sigma (T_b^4 - T_w^4)}{(T_b - T_w)} \quad (8.7)$$

where σ is the Stefan–Boltzmann constant. Radiative heat transfer is insignificant at temperatures, T , lower than $700^\circ C$. Typically bed emissivity, ε_b , is approximately 0.9 and wall emissivity, ε_w , is between 0.9 and 1.125 [8]. Since most drying processes are carried out at temperatures lower than $700^\circ C$, radiant heat transfer can be neglected.

The effect of various operating parameters on the heat transfer coefficient is given in Table 8.2.

8.4 MATHEMATICAL MODELS OF FLUIDIZED BED DRYING

Many mathematical models of fluidized bed drying have been proposed in the literature and verified with experimental data. These models have been developed based on different assumptions.

8.4.1 DIFFUSION MODEL

This model assumes that drying of single particles in a fluidized bed is totally controlled by diffusion of moisture inside the particle. For the analysis of particulate drying, diffusion equation for spheres of an equivalent diameter can be used. Zahed and Epstein [23] developed a diffusion model for spout bed drying and later Martinez-Vera et al. [24] applied the same model for fluidized bed drying.

This model assumes

- Solids are spherical, isotropic, uniform size, and homogeneous. They are perfectly well mixed in fluidized bed.
- Physical properties of the dry solids remain constant with time.
- Solids shrinkage and temperature gradient inside the solid are negligible.
- Drying kinetics is governed by internal moisture diffusion. Thus, moisture at the solid surface is in equilibrium with the bed air humidity.
- Air is perfectly mixed. Exhaust air is in thermal equilibrium with bed.
- The dryer is perfectly insulated.

The diffusivity is assumed constant. The following diffusion equation defines moisture transport:

$$\frac{\partial X}{\partial t} = D \left[\left(\frac{\partial^2 X}{\partial r^2} \right) + \frac{2}{r} \left(\frac{\partial X}{\partial r} \right) \right] \quad (8.8)$$

TABLE 8.2
Effect of Operating Parameters on Particle Heat Transfer Coefficient

Parameter	Effect on Heat Transfer Coefficient, h	Reference
<i>Particle</i>		
Diameter, d_p	For fine particles, h is higher; for coarse particles, h is lower	9
Shape	Higher for rounded and smooth surface particles	10,11
Specific heat, c_p	$h \propto c_p^n$, where $0.25 < n < 0.8$	12,13
Thermal conductivity, k_p	No influence for small Biot number	14,15
<i>Gas</i>		
Velocity, u_g	Increases above u_{mf} to a maximum value at an optimum velocity, u_{opt} and decreases thereafter	9
Density, ρ_g	Increases with increasing, ρ_g	10,11
Viscosity, μ_g	Increases with decreasing, μ_g	
Specific heat, c_g	At moderate pressure and velocity, no information available	16,17
	At high pressure, increases with increasing c_g	
Thermal conductivity, k_g	$h \propto k_g^n$, where $0.5 < n < 0.66$	14,18
	h Increases as bed temperature increases, due to increasing of k_g	
<i>Fluidized bed</i>		
Bed height, H_b	No influence	12,19
Bed diameter, d_b	No information available	
Bed temperature, T_b	Gas-convective: increases for small particles; decreases for coarse particles	20
Bed pressure, P_b	No influence on particle-convective heat transfer	21
	Gas-convective heat transfer increases	
<i>Heat transfer surface</i>		
Length, L	No influence	22
Tube diameter, d_{tube}	Increases with decreasing d_{tube}	17

where X is the free moisture content, i.e., that in excess of the equilibrium value, D is the diffusivity (m^2/s), and r is radial dimension (m).

If diffusivity is variable and dependent on the radial distance of drying boundary from the center of the solids, the following diffusion equation is used instead:

$$\frac{\partial X}{\partial t} = D \left[\left(\frac{\partial^2 X}{\partial r^2} \right) + \frac{2}{r} \left(\frac{\partial X}{\partial r} \right) \right] + \frac{\partial D}{\partial X} \left(\frac{\partial X}{\partial r} \right)^2 \quad (8.9)$$

Once the diffusivity is known, numerical analysis is applied to the diffusion equation in order to find moisture content profile inside the solid. Diffusivity of various food products can be obtained from Sablani et al. [25]. Average moisture content, \bar{X} , can be obtained from the following equation:

$$\bar{X} = \frac{4\pi}{V_p} \int_0^{r_p} r^2 X dr \quad (8.10)$$

where V_p is the particle volume (m^3). Note that moisture content and temperature-dependent diffusivity values can be used to solve the equation numerically.

8.4.2 EMPIRICAL MODEL

In this model, the drying process is divided into different periods where drying mechanisms in each dry-period are different.

The general solution of Fick's diffusion expresses the moisture content in terms of the drying time in exponential function. The solution for spherical solids is given in the following the equation [26–28]:

Sphere:

$$\frac{X - X_{eq}}{X_o - X_{eq}} = \frac{6}{\pi^2} \sum_{n=1}^{\infty} \frac{1}{n^2} e^{-n^2(\pi^2 D_{eff} t / r_{sph}^2)} \quad (8.11)$$

where r_{sph} is the sphere radius (m), D_{eff} is the effective diffusivity (m^2/s) and L is slab half thickness. Subscript “eq” denotes equilibrium and “o” indicates initial state.

Since the general solution of the diffusion equation is expressed as a series of exponential functions, experimental data obtained from fluidized bed drying can be correlated as an exponential function. Many empirical exponential equations have been proposed.

Equation 8.12 is a simple exponential equation. It assumes that the drying rate is proportional to the

difference between the average moisture content and the equilibrium moisture content [29]:

$$\frac{X - X_{eq}}{X_{ind} - X_{eq}} = e^{-kt} \quad (8.12)$$

where subscript “ind” denotes induction period. Equation 8.13 is a modified version of Equation 8.12 by Henderson and Pabis [30]. This equation is also analogous to the theoretical diffusion equation solution for an infinite slab [26,31]. Comparing Equation 8.13 and Equation 8.11, $b = D_{eff}\pi^2/r_{sph}^2$ [32]:

$$\frac{X - X_{eq}}{X_{ind} - X_{eq}} = ae^{-bt} \quad (8.13)$$

Equation 8.12 tends to overpredict the early stage and underpredict the later stage of drying. Equation 8.14 is an empirical modification of Equation 8.12 by introducing an exponent y [33]. It has been used most commonly because most experimental data can be fitted very well with the following equation:

$$\frac{X - X_{eq}}{X_{ind} - X_{eq}} = e^{-xty^y} \quad (8.14)$$

Equation 8.15 uses the first two terms from Fick’s second law of diffusion. This equation has been used regardless of solids geometry [34]:

$$\frac{X - X_{eq}}{X_{ind} - X_{eq}} = a_1e^{-b_1t} + a_2e^{-b_2t} \quad (8.15)$$

It should be noted that drying constants in the models mentioned above are empirical and depend on the type of materials, operating conditions as well as dryer dimensions. If one of these models is used for fluidized bed dryer design, experimental investigation on drying kinetics has to be conducted to obtain the drying constant for the particular material prior to the dryer design.

8.4.3 KINETIC MODEL

Chandran et al. [35] developed a kinetic model for fluidized bed drying of solids. For a batch fluidized bed kinetic model, it is assumed that the drying process has both constant and falling rate periods. Drying rate in the falling rate period falls linearly with decreasing moisture content. Feed conditions and total contact area between solids and hot airstream remain the same throughout the whole drying process. In the batch drying operation, there is little interaction between the particles (wet and dry particles) in the system. Thus, data on drying kinetics is

sufficient to estimate the residence time of solids in order to achieve the desirable final moisture content. Moisture content of solids in different drying periods can be estimated from the following equations.

In the constant rate period,

$$X = X_o - at \quad (8.16)$$

In the falling rate period,

$$X = X_{eq} + (X_{cr1} - X_{eq})e^{-a(t-t_{cr1})/(X_{cr1}-X_{eq})} \quad (8.17)$$

where subscript “cr1” denotes the first critical point that distinguishes constant and falling rate periods. For a single-stage continuous fluidized bed kinetics model, solids exit the fluidized bed system with a distribution of moisture content due to the wide residence time distribution. An average value of the moisture content and residence time is used.

The average moisture content of solids in a continuous fluidized bed drying is given by

$$\bar{X} = \int \left(\frac{X}{X_o} \right)_b E(\theta) d\theta \quad (8.18)$$

where $(X/X_o)_b$ is the moisture ratio in batch fluidized bed dryer, $E(\theta)$ is the residence time density for the solids, and $E(\theta) = e^{-\theta}$. $\theta = t/t_{cr1}$ is dimensionless time. Subscript “b” denotes batch process [36].

In the constant rate period,

$$\frac{\bar{X}}{X_o} = 1 - \frac{a\bar{t}}{X_o} \quad (8.19)$$

In the falling rate period,

$$\frac{\bar{X}}{X_o} = 1 - \frac{a\bar{t}}{(b\bar{t} + 1)X_o} \quad (8.20)$$

For a continuous fluidized bed that exhibits both constant and falling rate periods, the moisture content is then given by the following equation:

$$\frac{\bar{X}}{X_o} = 1 + \frac{a\bar{t}}{X_o} \left[\frac{b\bar{t}e - \theta_c}{b\bar{t} + 1} - 1 \right] \quad (8.21)$$

where $b = a/(X_{cr1} - X_{eq})$, $\theta_c = t/t_{cr1}$, \bar{X} is the average moisture content, \bar{t} is the average residence time, and a is the drying coefficient.

Once the average moisture content is known, equations obtained from mass and energy balances in the following models can be used to calculate the humidity and temperature of the exhaust air as well as the solids temperature. The simplest model is the

single-phase model that treats the fluidized bed as a continuum. As the number of phases considered in the model goes higher, the fluidized bed drying model becomes more complex and involves more transport properties. Complicated fluidized bed drying models that account for many transport processes that occur within and across the phases are beyond the scope of this chapter.

8.4.4 SINGLE-PHASE MODEL

In a single-phase model, the fluidized bed is regarded essentially as a continuum (Figure 8.4). Heat and mass balances are applied over the fluidized bed. It is assumed that particles in the bed are perfectly mixed. Equation 8.22 and Equation 8.23 are the equations of moisture balance and energy balance, respectively [24].

Moisture Balance:

$$-M_s \frac{d\bar{X}}{dt} = G_g(Y_{out} - Y_{in}) \quad (8.22)$$

where M_s is the mass hold-up of dry solid in bed (kg), \bar{X} is the average moisture content (kg/kg), G_g is the mass flow rate of dry air (kg/s), and Y is the air humidity (kg(water vapor)/kg(dry air)).

Energy Balance:

$$M_s c_{ps} \frac{dT}{dt} = G_g(c_g + Y_{in} c_v)(T_{in} - T_{out}) - G_g(Y_{out} - Y_{in})\lambda \quad (8.23)$$

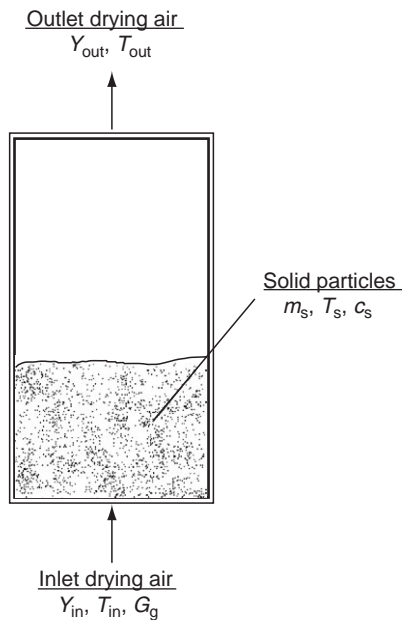


FIGURE 8.4 Schematic diagram of the single-phase model of fluidized bed dryer.

where c_p is the heat capacity at constant pressure (kJ/(kg K)) and λ is the latent heat of vaporization (kJ/kg). Subscript “s” denotes wet solid, “g” denotes dry air, and “v” denotes water vapor. Equation 8.23 neglects sensible heat of the water in solids.

8.4.5 TWO-PHASE MODEL

A simple two-phase model of fluidized bed drying treats the fluidized bed to be composed of a bubble phase (dilute phase) and an emulsion phase (dense phase). The bubble phase contains no particles or the particles are widely dispersed. This model assumes that all gas in excess of minimum fluidization velocity, u_{mf} , flows through the bed as bubbles whereas the emulsion phase stays stagnant at the minimum fluidization conditions [37]. Figure 8.5 shows a schematic diagram of the simple two-phase model.

Zahed et al. [38] have presented mass and energy balance equations for the dense phase and the bubble phase for fluidized bed drying. Mass balance of liquid in the bubble phase gives the following equation:

$$\rho_g \epsilon_{bb} \frac{dY_{bb}}{dt} + \rho_g \frac{V_{g,bb}}{V_b t} (Y_{bb} - Y_{in}) = \frac{6K_c \rho_g \epsilon_{bb}}{d_{bb}} (Y_d - Y_{bb}) \quad (8.24)$$

where subscript “bb” denotes bubble phase and “d” denotes dense phase. The rate of change of mass in the bubble phase can be assumed to be negligible [38] and Equation 8.24 can be rearranged to express humidity in the bubble phase, Y_{bb} in terms of humidity in the dense phase, Y_d . In the equation, $V_{g,bb}/V_b$ is the gas flow rate in bubble phase per unit volume of bed.

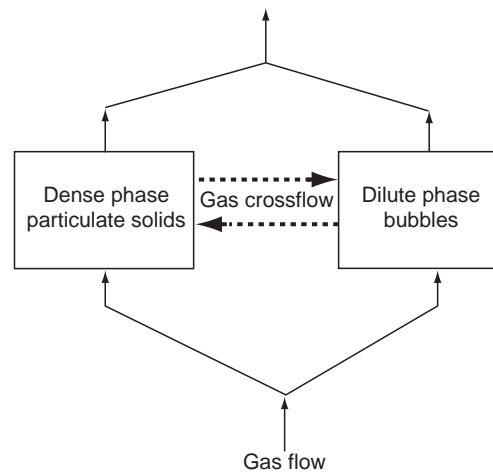


FIGURE 8.5 Schematic diagram of a two-phase model for fluidized bed drying.

K_c is the mass transfer coefficient across the bubble boundary.

Mass balance of liquid in the interstitial gas in the dense phase gives the following equation:

$$\begin{aligned} & \frac{6K_c\rho_g\varepsilon_{bb}}{d_{bb}}(Y_{bb} - Y_d) - \rho_g \frac{V_{g,d}}{V_b t}(Y_d - Y_{in}) + \dot{m} \\ & = \rho_g \varepsilon_{mf}(1 - \varepsilon_{bb}) \frac{dY_d}{dt} \end{aligned} \quad (8.25)$$

Likewise, the rate of change of mass in the interstitial gas can be assumed to be negligible. In this equation, \dot{m} is the mass rate of evaporation of water per unit volume of bed, which in turn can be obtained from mass balance on dense phase. $V_{g,d}/V_b$ is the gas flow rate in dense phase per unit volume of bed.

Mass balance of liquid in the dense-phase particles yields the following equation:

$$\dot{m} = -\rho_p(1 - \varepsilon_{mf})(1 - \varepsilon_{bb}) \frac{d\bar{X}}{dt} \quad (8.26)$$

The coupled mass and energy balance in dense phase that consists of particles and interstitial gas phases is given in the following equation:

$$\begin{aligned} & \rho_p(1 - \varepsilon_{mf})(1 - \varepsilon_{bb})(c_{ps} - c_{pl}\bar{X}) \frac{dT_p}{dt} \\ & = \rho_g \frac{V_{g,d}}{V_b t}(c_{pg} + Y_{in}c_{pv})(T_{g,in} - T_p) - \Delta H_{evap} \\ & \times \left[\rho_g \frac{V_{g,d}}{V_b t}(Y_d - Y_{in}) - \frac{6K_c\rho_g\varepsilon_{bb}}{d_{bb}}(Y_{bb} - Y_d) \right] \end{aligned} \quad (8.27)$$

The above equation expresses the change of particle temperature in the dense phase in terms of average moisture content, \bar{X} , which can be determined from any one equation from Equation 8.8 through Equation 8.21 depending on the operating conditions and the drying model, humidity of dense and bubble phases, Y_d , Y_{bb} , enthalpy of evaporation, ΔH_{evap} , bubble diameter, d_{bb} , and mass transfer coefficient of bubble boundary. Solving Equation 8.27 yields the solids temperature at different drying times.

8.5 EFFECT OF OPERATING PARAMETERS ON FLUIDIZED BED DRYING

8.5.1 EFFECT OF BED HEIGHT

For materials with high mobility of internal moisture such as iron ore, ion-exchange resins, silica gel, most drying takes place close to the distributor plate. Bed height has no effect on its drying rate that increasing

bed height beyond a particular value leads to no differences in drying rates. For materials with main resistance to drying within the material, e.g., grains, drying rate decreases with increasing bed height.

8.5.2 EFFECT OF PARTICLE SIZE

For group B particles (sandlike particles, according to Geldart Classification of Powders), drying time that is required to remove a given amount of moisture increases as the square of the particle diameter provided that all other conditions remain unchanged. However, this effect is much smaller for group A (aeratable particles, according to Geldart Classification of Powders) particles because these particles are finer than group B and it exhibits smooth fluidization before entering bubbling fluidization regime.

8.5.3 EFFECT OF GAS VELOCITY

Gas velocity has a dominant effect on removing surface moisture. Increasing the gas velocity increases the drying rate. However, gas velocity has no effect at all for particles with high internal resistance to moisture transfer. High internal moisture resistance dominates at the end of the falling rate period.

8.5.4 EFFECT OF BED TEMPERATURE

Bed temperature is increased by high external heat fluxes. This in turn leads to higher moisture diffusivities and hence higher drying rate. This effect is complex and depends on the relative significance of external and internal resistances to moisture transfer.

8.6 TYPES OF FLUIDIZED BED DRYERS: CLASSIFICATION AND SELECTION

Various types of fluidized bed dryers have been studied, developed, and operated in many industrial processes according to the respective process, product, operational safety, and environmental requirements. It is important to become familiar with the specific characteristics of different fluidized bed types in order to make a logical and cost-effective selection. It should be noted that in many instances several different types may provide similar performance at the same cost.

Some novel fluidized bed dryers, which have not found application in industrial drying, are used to overcome disadvantages and difficulties that may occur in conventional fluidized bed dryers. It should be noted that not all modified fluidized bed dryers are necessarily better than the conventional dryers in terms of product quality, or energy efficiency, or drying performance.

TABLE 8.3
Classification of Fluidized Bed Dryers

Criterion	Type of Dryer	Subclassification
Processing mode/feed and discharge	<ul style="list-style-type: none"> • Batch FBDs (well-mixed) • Semicontinuous FBDs • Continuous 	<ul style="list-style-type: none"> • Well-mixed FBDs • Plug flow FBDs or • Single stage • Multistage FBDs • Hybrid/combined FBDs
Particulate flow regime	<ul style="list-style-type: none"> • Well-mixed FBDs • Plug flow FBDs • Circulating FBDs • Hybrid 	<ul style="list-style-type: none"> • Multistage FBDs (well-mixed—plug flow) • Hybrid/combined FBDs
Operating pressure	<ul style="list-style-type: none"> • Low (for heat-sensitive products, low pressure strategy) • Near atmospheric (most common) • High (5 bars, superheated steam FBDs) 	
Fluidization gas flow	<ul style="list-style-type: none"> • Continuous • Pulsed FBDs 	
Fluidizing gas temperature	<ul style="list-style-type: none"> • Constant • Time-dependent 	<ul style="list-style-type: none"> • Step down • Step up • Periodic (zigzag) • Combined
Heat supply	<ul style="list-style-type: none"> • Convective • Convective/conduction (immersed FBDs) <p>or</p> <ul style="list-style-type: none"> • Continuous • Intermittent (multiple variable strategy) 	
Fluidization action	<ul style="list-style-type: none"> • By gas flow (pneumatic) • By jet flow • With mechanical assistance • With external field 	<ul style="list-style-type: none"> • Ordinary FBDs • Circulating FBDs • Spouted FBDs • Recirculating FBDs • Jetting FBDs • Vibration (vibrated FBDs) • Agitation (agitated FBDs) • Rotation (centrifugal FBDs) • Microwave–radio frequency field (MW–RF FBDs) • Acoustic field • Magnetic field
Fluidized material	<ul style="list-style-type: none"> • Particulate solid (most common) • Paste/slurry 	<ul style="list-style-type: none"> • Group A and B (most common, conventional FBDs) • Group C (vibrated FBDs, agitated FBDs) • Group D (vibrated FBDs, baffled FBDs, spouted FBDs) • Spray onto a bed of inert particles (inert solids FBDs) • Spray onto absorbent particles (silica gel, biomass) • Spouted FBDs
Fluidizing medium	<ul style="list-style-type: none"> • Heated air/flue gases/direct combustion gas • Superheated steam/vapor • Dehumidified cool air (heat pump FBDs) • Air below freezing point of liquid being removed (fluidized bed freeze dryers) 	

Table 8.3 classifies the diverse variants of fluidized bed dryers according to various criteria.

8.7 CONVENTIONAL FLUIDIZED BED DRYERS

8.7.1 BATCH FLUIDIZED BED DRYERS

A batch fluidized bed dryer is used when production capacity required is small (normally 50 to 1000 kg/h) or several products are to be produced in the same production line. It is preferable to operate batchwise if upstream and downstream processes are operated in batch mode, or several processes are to be carried out in sequence (e.g., mixing, drying, granulation/coating, cooling) in the same processing unit.

Drying air temperature and flow rate are normally fixed at a constant value. However, by adjusting the airflow rate and its temperature, it is possible to save energy and reduce attrition. Mechanical assistance such as agitation or vibration is normally applied for processing materials that are difficult-to-fluidize. Figure 8.6 shows a typical batch fluidized bed dryer with expanded freeboard and built-in internal bag filters. Expanded freeboard is used to reduce elutriation of fine particles.

8.7.2 SEMICONTINUOUS FLUIDIZED BED DRYERS

In semicontinuous fluidized bed drying system, the drying chamber consists of a series of subprocessors. The wet product is accurately dosed and charged into the batches. The product is either transported batchwise from one processor to another processor or the batches (the processors with the batches of product) rotates along the process line [39]. This gives uninterrupted continuous operation over a long period. Figure 8.7 shows a schematic diagram of a semicon-

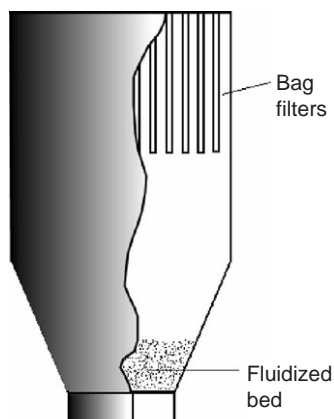


FIGURE 8.6 Batch fluidized bed dryer.

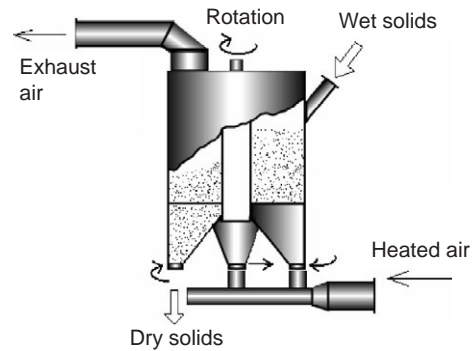


FIGURE 8.7 Semicontinuous fluidized bed dryer.

tinuous fluidized bed dryer where the batches are rotated. In addition, gas temperature and velocity at different batches can be varied.

8.7.3 WELL-MIXED, CONTINUOUS FLUIDIZED BED DRYERS

The well-mixed continuous fluidized bed dryer (Figure 8.8) is one of the most common fluidized bed dryers used in the industry. As the bed of particles is perfectly mixed, the bed temperature is uniform and is equal to the product and exhaust gas temperatures. However, particle residence time distribution is necessarily wide, thus resulting in wide range of product moisture content. On the other hand, as the feed material is continuously charged into the fluidized bed of relatively dry particles, this gives the added advantage of enhanced fluidizability and better fluidization quality. In some cases, a series of well-mixed continuous dryers may be used with variable operating parameters. In addition, a well-mixed continuous

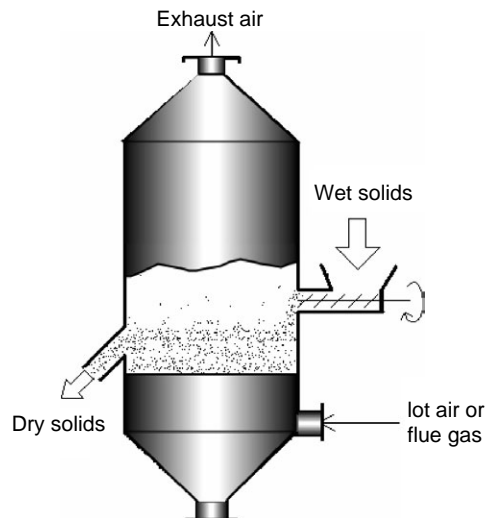


FIGURE 8.8 Well-mixed fluidized bed dryers.

fluidized bed dryer can be incorporated with other types of dryers such as plug flow fluidized bed dryers to give better drying performance.

8.7.4 PLUG FLOW FLUIDIZED BED DRYERS

In plug flow fluidized bed dryers, vertical baffles are inserted to create a narrow particle flow path, thus giving relatively narrow particle residence time distribution. Particles flow continuously as a plug from the inlet toward the outlet through the path. This ensures nearly equal residence time for all particles irrespective of their size and ensures uniform product moisture content. Various paths can be designed such as straight or spiral paths. Length-to-width ratio is normally in the range of 5:1 to 30:1. Figure 8.9 shows a plug flow fluidized bed dryer of straight and reverse paths.

Operational problems might occur at the feed inlet because wet feedstock must be fluidized directly rather than when mixed with drier material as in the case of a well-mixed unit. To overcome the problem of fluidizability at the feed inlet, the inlet region may be agitated with an agitator, or by applying backmixing of solids, or by using a flash dryer to remove the surface moisture prior to plug flow fluidized bed drying.

8.8 MODIFIED FLUIDIZED BED DRYERS

Various types of modified fluidized bed dryers have been developed and applied in many industrial processes. Modified fluidized bed dryers are applied to overcome some of the problems and disadvantages encountered in conventional fluidized beds.

8.8.1 MULTISTAGE AND MULTIPROCESS FLUIDIZED BED DRYERS

As fluidized beds offer many distinct features and advantages for processing of particulate materials,

two or more processes can be carried out and accomplished in a fluidized bed column. This can be achieved by simply changing the operating conditions of fluidized bed to mix, dry, granulate, or coat [40], or cool in a single unit without discharging the material from the unit.

In a fluidized bed spray dryer, spray drying is carried out in the upper part of the chamber followed by fluidized bed drying or agglomeration (Figure 8.10a). The large-scale fluidized bed coal dryer is also a particle classifier (Figure 8.10b). Drying and classification (separation of fines) are carried out in the same fluidized bed. By changing the fluidizing gas velocity, cut size (particle size that separates fine and coarse particles) can be adjusted in the classification process. Another example is upper stage of fluidized bed drying that can be followed by a lower stage of fluidized bed cooling (Figure 8.11a). A fluidized bed dryer or cooler consists of first-stage fluidized bed dryer followed by second-stage fluidized bed cooling (Figure 8.11b).

In addition, different types of fluidized bed systems can be incorporated in a processing unit as well. For instance, first-stage well-mixed fluidized bed pre-drying can be incorporated with second-stage plug flow fluidized bed drying (Figure 8.11a). By incorporating different processes and combining different types of fluidized beds, space requirement, installation costs, and energy consumption can be reduced appreciably.

8.8.2 HYBRID FLUIDIZED BED DRYERS

Hybrid fluidized bed dryers are useful for through drying of solids that contain surface and internal moistures. Surface moisture can be removed in the first-stage drying using a flash or cyclone dryers. Second-stage drying is then carried out in fluidized bed dryers in which residence time can be easily controlled. Figure 8.12 shows an example of hybrid

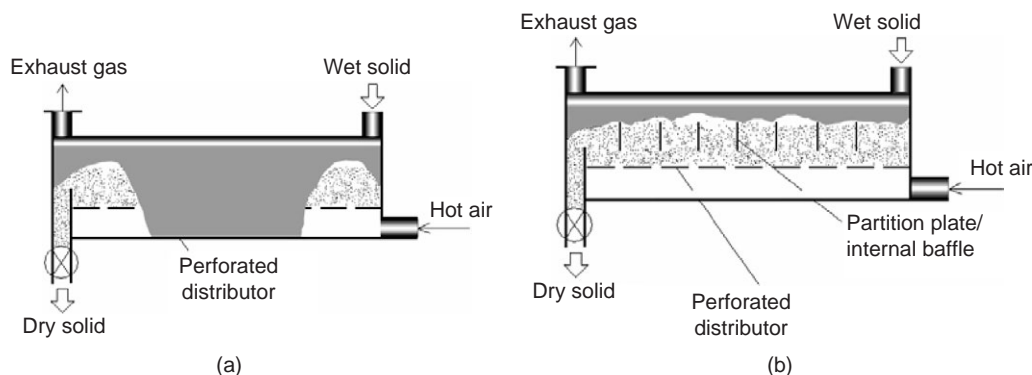


FIGURE 8.9 Plug flow fluidized bed dryers. (a) Straight path; (b) reversing path.

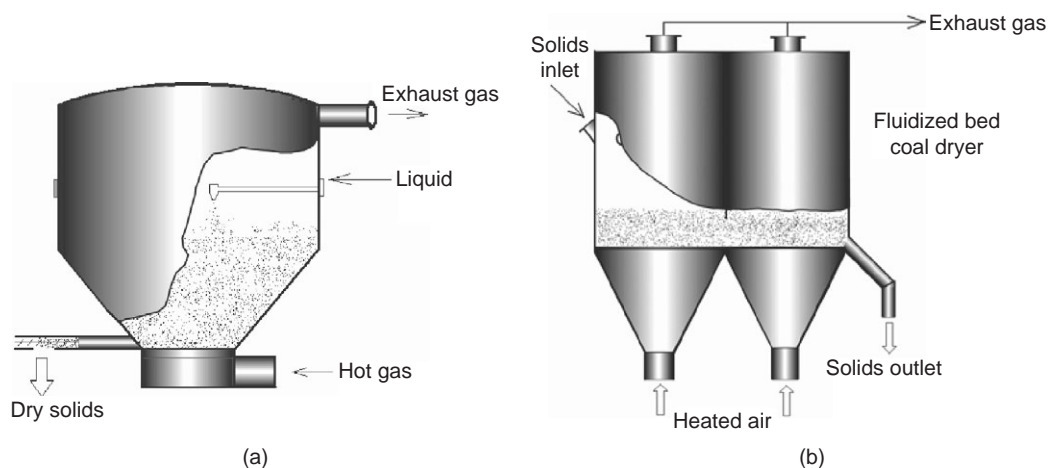


FIGURE 8.10 (a) Spray fluidized bed dryer; (b) fluidized bed coal dryer and classifier.

cyclone fluidized bed dryer [39]. Wet solids are first charged into the cyclone dryer by exiting fluidizing gas from fluidized bed dryer. Surface moisture content of solids is quickly removed with the gas in the cyclone dryer. Solids and gas are separated in the cyclone. Partially dried solids are then pneumatically conveyed into the fluidized bed for second-stage drying. Other types of hybrid fluidized bed dryers include flash-fluidized bed dryer, filter-fluidized bed dryer [41].

A multistage spray fluidized bed dryer consists of a spray chamber followed by first-stage fluidized bed drying and second-stage fluidized bed cooling (Figure 8.13). When solid powders are formed in the spray dryer, these powders still contain some internal moisture. It is costly to use a spray dryer to remove all of the internal moisture. Instead, using a second-stage fluidized bed dryer is more cost-effective. Lisboa et al. [42] applied fluidization technique in a conventional rotary. The dryer is known as roto-fluidized dryer.

It was found that the roto-fluidized dryer performs better than the conventional rotary dryer.

8.8.3 PULSATING FLUIDIZED BED DRYERS

Pulsating fluidized bed dryers are used to overcome the problems of restricted particle size and size distribution, as well as aggregative fluidization and channeling that occur in a conventional fluidized bed dryer when processing certain types of powders. By pulsating the fluidizing gas stream, the fluidized bed either the whole bed or part of the bed is subjected to variable fluidizing gas velocity (Figure 8.14) [43–46]. This contributes to effective energy costs saving and enhanced drying performance without affecting the fluidization quality and process performance or added extra capital costs. For larger particles (group D particles), intermittent spouting of the bed with a rotating spouting jet has been shown to reduce energy consumption with only a marginal increase in drying time for batch drying.

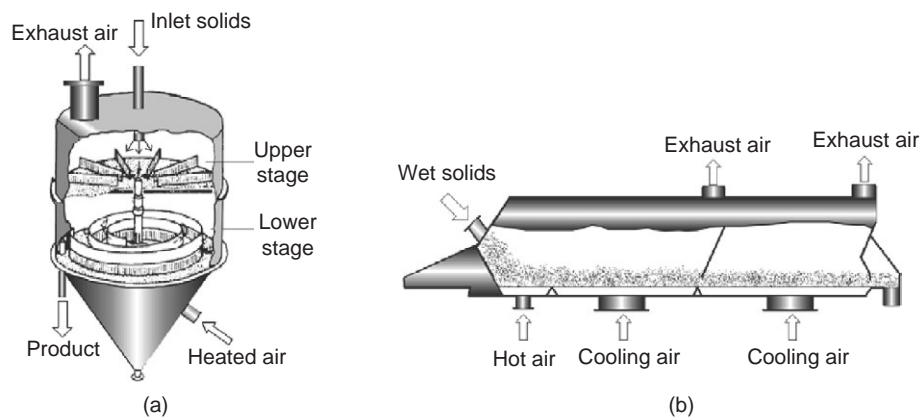


FIGURE 8.11 Two-stage fluidized bed dryers. (a) Upper stage well-mixed fluidized bed followed by lower stage plug flow fluidized bed; (b) first-stage dryer followed second-stage cooler.

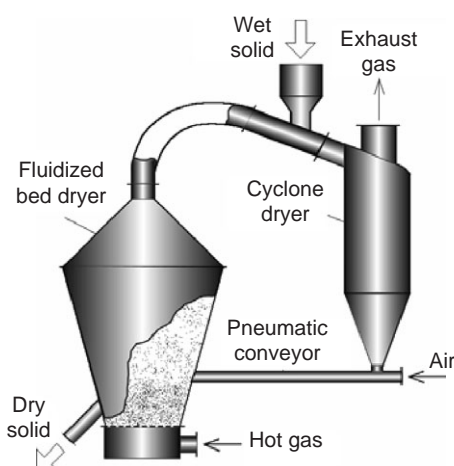


FIGURE 8.12 Hybrid cyclone fluidized bed dryer.

8.8.4 FLUIDIZED BED DRYERS WITH IMMERSSED HEAT EXCHANGERS

Fluidized beds equipped with internal heaters or immersed tubes transfer heat indirectly to the drying material. Horizontal tube bundles (Figure 8.15) are used extensively compared to vertical type. Tube pitch is an important design parameter. Fluidizing gas stream fluidizes the material and carries over the evaporated moisture. As a result, total sensible heat of gas and thus quantity of gas required are reduced. Immersed tubes or internally heated fluidized bed dryers are used to dry smaller size or fine powders. This is because heat transfer coefficient decreases with increasing particle size. Instead of tubes, vertical plates are also used as immersed heaters.

Heat transfer is highly dependent on the particle heat capacity and mixing. Vigorous bubble action gives better particle circulation and mixing, and thus

increases the contacting efficiency between the bed and the heat transfer surface. However, heat transfer coefficient reaches a maximum value. Beyond this point, increasing superficial gas velocity will hinder heat transfer between the bed and the heating surface. This is because of increasing preponderance of bubble (not particles) at the heating surface, which decreases particle-to-wall heat transfer.

8.8.5 MECHANICALLY ASSISTED FLUIDIZED BED DRYERS

Fluidization quality of fine and large particles can be enhanced by the assistance of external means such as vibration or agitation. Moreover, these particles can be immersed in a bed of fluidizable inert particles to improve their fluidization quality [47].

8.8.6 VIBRATED FLUIDIZED BED DRYERS

Vibration combined with upward flow of air in an aerated bed enables particles to pseudofluidize smoothly. The gas velocity required for minimum fluidization is considerably lower than the minimum fluidization velocity in conventional fluidized bed dryer. Attrition due to vigorous actions between particle-particle and particle-wall is thus minimized appreciably. Hence, application of fluidized bed can be extended to fragile, abrasive, and heat-sensitive materials. The problem of fine particle entrainment is also avoided. For polydisperse powders, low gas velocity fluidizes the fine particles gently whereas vibration keeps the coarse particles in a mobile state.

Vibrating fluidized beds are generally plug flow type (Figure 8.16). Vibrating fluidized beds are relatively shallow as the effect of vibration imparted by the vibrating grid decays with distance from the grid.

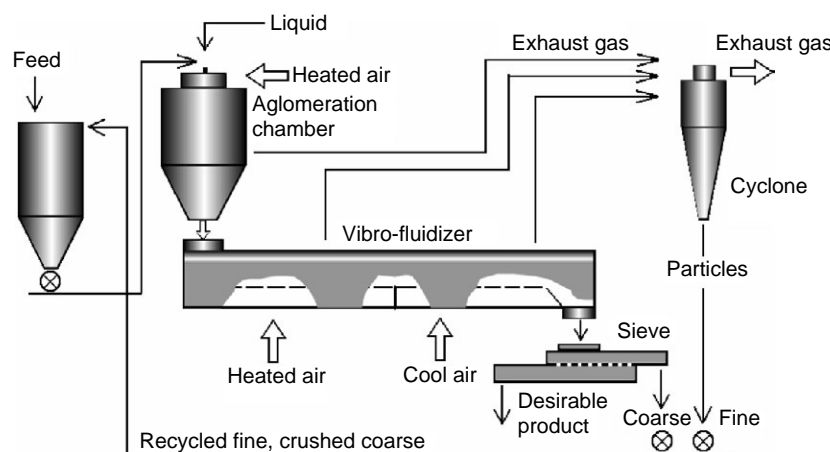


FIGURE 8.13 Multistage fluidized bed spray dryer.

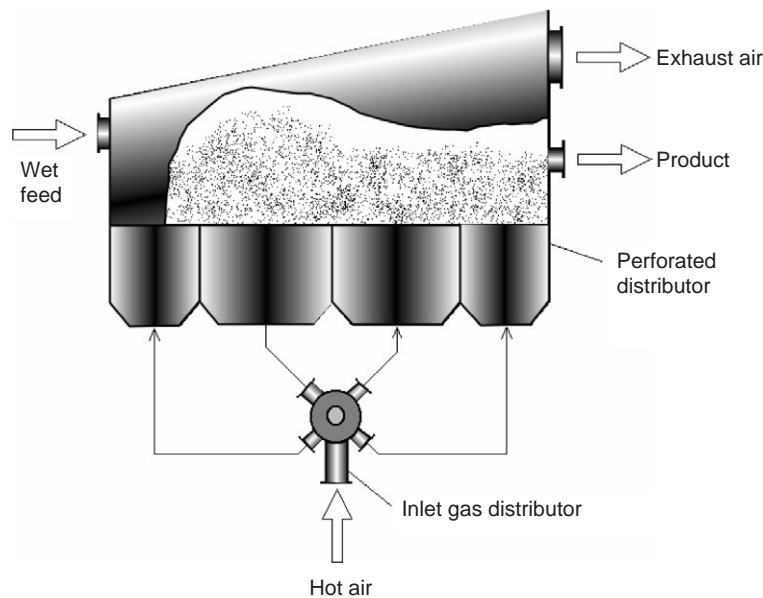


FIGURE 8.14 Pulsating fluidized bed. Parts of the bed are fluidized periodically.

There are some acoustic noise issues associated with such devices. These units can operate in batch as well as continuous modes.

8.8.7 AGITATED FLUIDIZED BED DRYERS/SWIRL FLUIDIZERS

Another way to improve fluidization quality of fine particles is to impart mechanical agitation to the bed (Figure 8.17). By agitation, a homogeneous fluidized

bed is formed without channeling or formation of large bubbles. Moreover, agitated fluidized bed dryers are useful for drying pastes or cakes consisting of fine particles [48]. In this case, agitation helps to disintegrate and disperse the pasty feed. The agitator serves as a mixer in the dryer [49]. Moreover, deeper bed depth is possible if the bed is agitated whereas its fluidization quality is maintained.

8.8.8 FLUIDIZED BED DRYERS OF INERT PARTICLES

In recent years, the application of fluidized bed drying has been extended to drying of fine powders, pastes, slurries, suspensions, pulp, and enzymes-containing aqueous medium [50–55]. This is accomplished by using inert particles of high heat capacity (Figure 8.18) [56]. Inert particles must be able to fluidize well in a fluidized bed. By mixing the inert particles whose fluidization quality is generally good with the materials mentioned above, the fluidization quality of the materials is improved appreciably [57,58]. In addition, the inert particles with high heat capacity serve as energy carriers that enhance heat transfer [59,60]. Drying on inert particles can be performed in a variety of fluidized beds namely ordinary fluidized bed, spouted bed, spouted fluidized bed, jetting-spouted bed, as well as vibrated fluidized bed [61].

The liquid to be dried is sprayed into the fluidized bed; it coats the inert particle surfaces. The coated layer dries as a result of combined convective heat transfer from hot air and contact heat transfer due to sensible heat of the particles. When the thin layer is

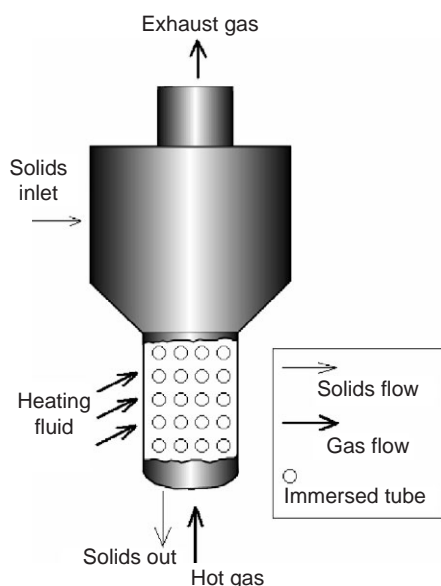


FIGURE 8.15 Immersed tubes fluidized bed dryer.

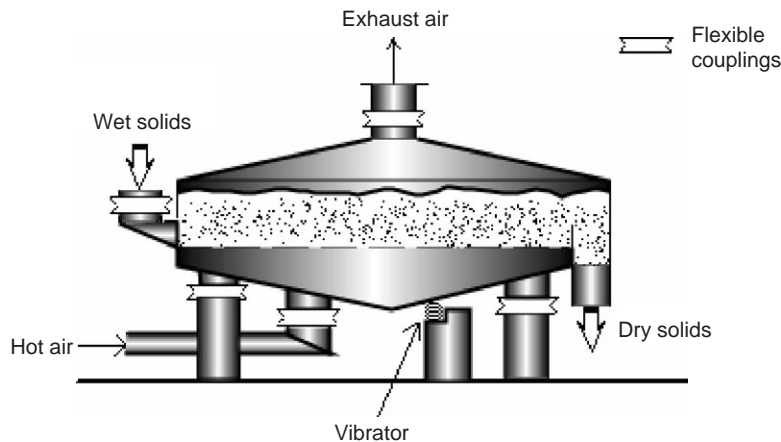


FIGURE 8.16 Vibrating fluidized bed.

dry, it becomes brittle, cracks, and is peeled off due to attrition by particle-particle and particle-wall collisions. As a result, a fine powder is formed and is carried over by the exhaust gas to be collected and separated in suitable gas-cleaning devices such as cyclones or bag filters.

8.8.9 SPOUTED BED DRYERS

Spouted bed dryers are useful for drying of large (Geldart's group D) particles (>5 mm), which exhibit slugging under normal fluidization. In a spouted bed, a high gas velocity jet of gas penetrates through an opening at the bottom of the bed of particles and

transports the particles to the bed surface. Energetic spouting at the bed surface thrusts the particles into the freeboard region at the center of the bed (Figure 8.19). After losing their momentum, these particles fall back onto the bed surface. Through this fountain-like action, good solid mixing is induced. A cyclical flow of particles is thus created. Details of spouted bed drying are discussed elsewhere in this handbook.

The spout bed has been applied to drying, granulation, coating as well as to drying of pastes, solutions, slurries, and suspensions. Mujumdar [62] has classified spouted beds into at least 30 different variants, each with a specific set of advantages and limitations. Periodically spouted beds, multiple spouted

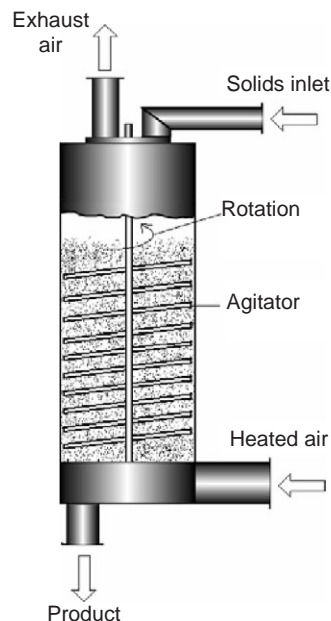


FIGURE 8.17 An agitated fluidized bed dryer.

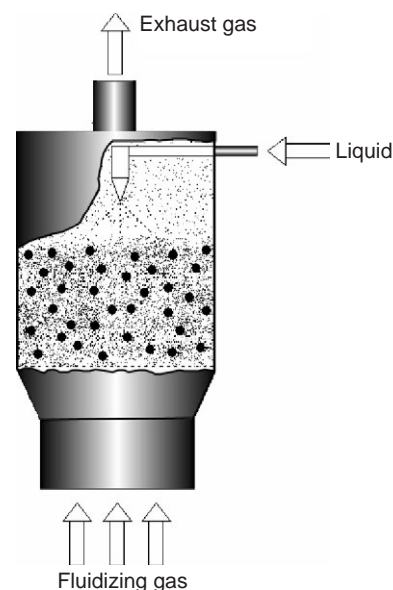


FIGURE 8.18 Inert solids fluidized bed.

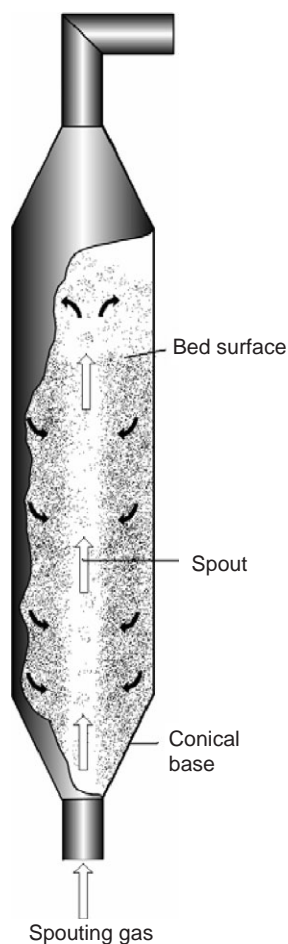


FIGURE 8.19 Spouted bed dryers.

beds, two-dimensional spouted beds, and oscillating spouted beds are some of the ideas introduced by Mujumdar in 1985, which have been examined in the literature in recent years.

8.8.10 RECIRCULATING FLUIDIZED BED DRYERS

Insertion of a tubular draft tube into an ordinary spouted fluidized bed changes its operational and design characteristics. This type of fluidized bed is known as recirculating fluidized bed (or internally circulating fluidized bed, see Figure 8.20). Unlike spouted beds, recirculating fluidized beds do not have limitation of maximum spoutable bed height and minimum spouting velocity. As spouting gas stream passes through the draft tube, it is confined within the tube and does not leak out horizontally toward the downcomer. After passing through the draft tube, particles follow a certain flow pattern in the bed and flow downward in downcomer region.

Since there is more flexibility in operating recirculating fluidized bed, it is applicable to handle all

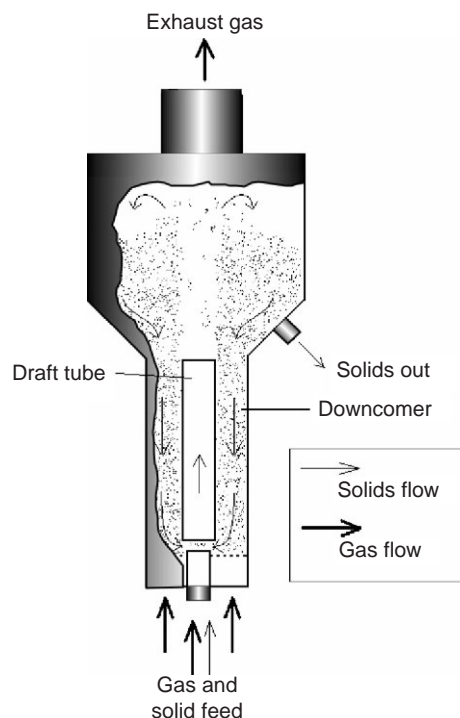


FIGURE 8.20 Recirculating fluidized bed dryer.

groups of powders and particles. Its application in drying has been reported in coating of tablets in pharmaceutical industries, and in drying of dilute solutions containing solids. However, it is not a common dryer type now.

8.8.11 JETTING FLUIDIZED BED DRYERS

In an ordinary fluidized bed, inlet gas is passed through nozzles, which are perforated evenly across the distributor plate. Jetting regions appear above every nozzle. In a spouted bed, inlet gas stream is supplied through a centrally located jet, spout in dilute phase is thus created, and penetrates the center region of the spouted bed (Figure 8.19). However, if a fairly large jet replaces the conical centrally located jet in a spout bed, a jetting fluidized bed is formed. One distinctive feature of jetting fluidized bed is that bubbles are formed instead of dilute phase spout (Figure 8.21). Small-scale jetting fluidized beds have been applied in coating and granulation processes.

8.8.12 FLUIDIZED BED DRYERS WITH INTERNAL Baffles

Internal baffles can be inserted into a fluidized bed to divide the bed into several compartments. Various types of baffles can be used, e.g., wire mesh, perforated plate, turn plate, louver plate, and ring [63]. In

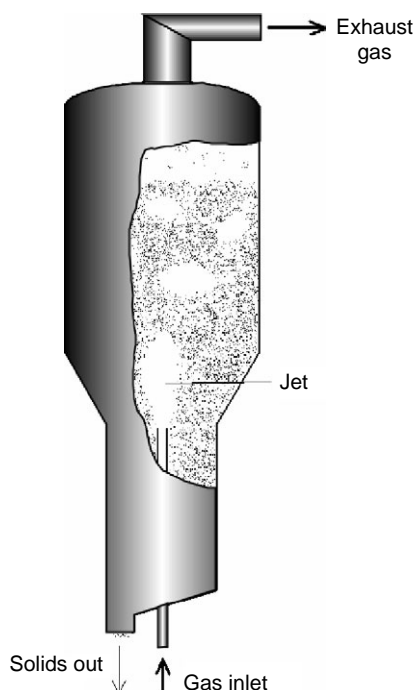


FIGURE 8.21 Jetting fluidized bed dryer.

addition, the internals can be placed horizontally or vertically (Figure 8.22). Horizontal baffles are frequently used. The objective of inserting baffles (horizontal and vertical) is to limit bubble growth and coalescence [64,65]. Hence, the baffled fluidized bed is useful to process group B and D particles because large bubbles are formed with such particles. The effect of baffles on the gas and solids flow is very complex and is dependent on bed diameter, distance between baffles, baffle opening and operating conditions. The optimum conditions for operating an

efficient baffled fluidized bed dryer can only be determined by carrying out pilot testing.

8.8.13 SUPERHEATED STEAM FLUIDIZED BED DRYERS

Superheated steam as the fluidizing medium offers a number of advantages, e.g., no fire or explosion hazards, no oxidative damage, better operation performance (higher drying rate) and product quality, environmental friendliness, high energy consumption efficiency, suitability for drying of products containing toxic or expensive organic liquids, ability to permit pasteurization, sterilization, and deodorization of food products [66,67]. Details are available elsewhere in this handbook.

The application of superheated steam fluidized bed dryer has been reported for drying of paper and pulp, wood-based biofuels (Figure 8.23), sugar beet pulp, and paddy [65,66]. Superheated steam fluidized bed drying of foodstuff, coal, bagasse, sludges, spent grains from breweries, lumber, tortilla, vegetables, herbs, and spice is also possible [67,70].

8.8.14 FLUIDIZED BED FREEZE DRYER

Freeze-drying is one of the low-temperature drying techniques suitable for drying of highly heat-sensitive materials such as drugs, pharmaceutical, biological, and food products. Freeze-drying removes moisture captured inside the solids by sublimation of moisture from solid state (ice) to vapor state.

Ordinary freeze-drying is carried out in vacuum. Over the years, new developments showed that freeze-drying can be carried out at atmospheric pressure and as well as in a fluidized bed (e.g., Refs. [71–73]). Here the drying rate is very slow. Wolff and Gibert [74] showed that fluidized bed freeze-drying at

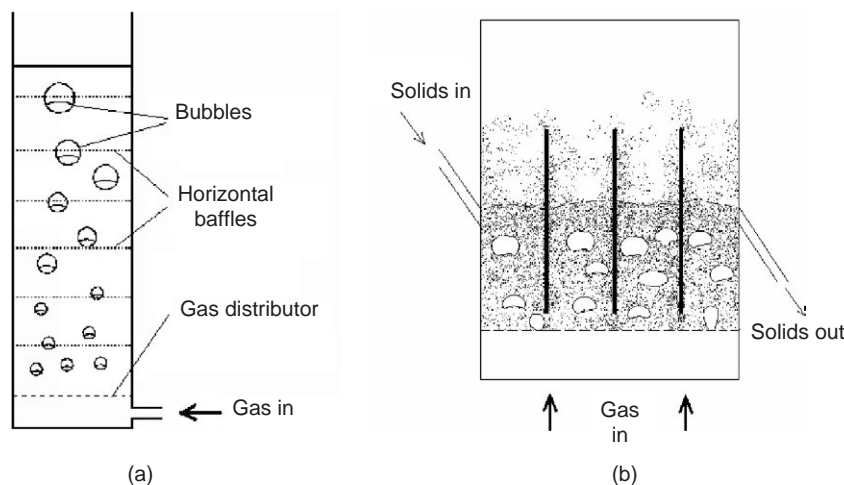


FIGURE 8.22 (a) Horizontal baffled fluidized bed dryer; (b) vertical baffled fluidized bed dryer.

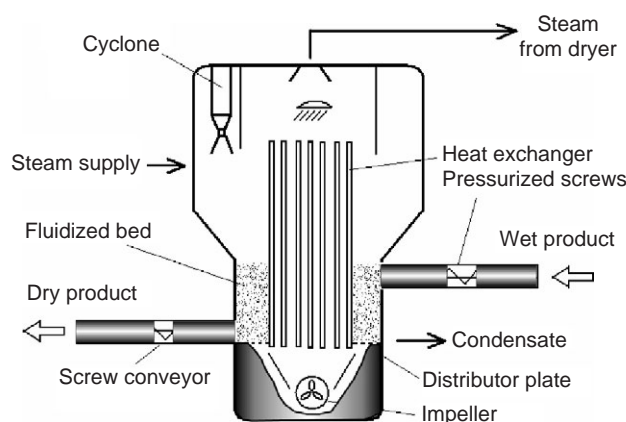


FIGURE 8.23 Pressurized superheated steam fluidized bed dryer.

atmospheric pressure with the use of adsorbents can increase the drying rate appreciably (about sevenfold compared to that without adsorbent). In this case, adsorbent particles play a dual role as transfer agent for both heat and mass transfers. But there is difficulty in separating adsorbent particles and frozen dried products at the end of the process. It is thus suggested to use particles that are edible or compatible with human consumption such as starch. Fluidized bed freeze-drying assisted by adsorbent involves three stages, namely freezing of product, sublimation of free-frozen water, and secondary dehydration by desorption.

Wolff and Gibert [75] suggested that fluidized bed freeze-drying should be carried out at higher temperature, but lower than the freezing point. They showed that fluidized bed freeze-drying with absorbent contributes to about 35% saving in heat requirement, respectively, although much longer drying time is needed as compared to vacuum freeze-drying.

8.8.15 HEAT PUMP FLUIDIZED BED DRYER

An ordinary fluidized bed drying system consists of a blower, heater, dehumidifier (optional), fluidized bed-chamber, and cyclone, whereas an ordinary heat pump drying system consists of evaporator, compressor, condenser, and an expansion valve. By combining fluidized bed and heat pump drying systems, where the evaporator acts as a dehumidifier and the condenser as a heater, a heat pump fluidized bed dryer is formed.

The working fluid (refrigerant) at low pressure is vaporized in the evaporator by heat drawn from the exhaust humid air. At the same time, condensation of moisture occurs as the exhaust air temperature goes below dew point temperature. Thus, the process air is dehumidified. The working fluid then goes to

compressor. The compressor raises the enthalpy of the working fluid and discharges it as superheated vapor at high pressure. Heat is removed from the working fluid and returned to the process air, which has been dehumidified previously at the condenser. As a result, the process air temperature increases. The working fluid is then throttled using an expansion valve to the low-pressure line and enters the evaporator to complete the cycle, whereas the dehumidified and heated process air is charged into the fluidized bed drying chamber to remove moisture of solids. Details on heat pump drying are available elsewhere in this handbook.

Figure 8.24 shows a typical heat pump fluidized bed dryer. The fluidized bed drying chamber receives wet solids and discharges dried product whereas dehumidified and heated air is charged into the chamber from the bottom of the chamber. The drying temperature can be adjusted by monitoring the capacity of condenser, whereas the desired humidity of inlet air can be obtained by controlling the motor frequency of compressor.

The advantages offered by heat pump fluidized bed dryer are: low energy consumption due to high specific moisture extraction rate (SMER), high coefficient of performance (COP), wide range of drying temperature (-20 to 110°C), environmental friendliness, and high product quality. Thus this type of dryer is suitable for heat-sensitive products such as food and products of bio-origin.

As chlorofluorocarbons (CFC) and hydrochlorofluorocarbons (HCFC) are to be phased out very soon, working fluids such as carbon dioxide, ammonia, R717, and R744 can be used as substitutes [76]. Many products have been tested at the Norwegian Institute of Technology, such as food products, fish, fruits, and vegetables [77,78].

8.9 DESIGN PROCEDURE

Design procedures for batch and continuous dryers in constant and falling rate periods vary widely. The discussion here is restricted to particulate solids drying.

8.9.1 DESIGN EQUATIONS

8.9.1.1 Residence Time

If the particles are small, very porous, and sufficiently wet to contain free moisture, the drying rate remains constant throughout the drying process. On the other hand, if the solid particles initially contain surface moisture, falling rate period will occur after a short period of constant rate period. In this case, the design calculation should include two steps: one for the

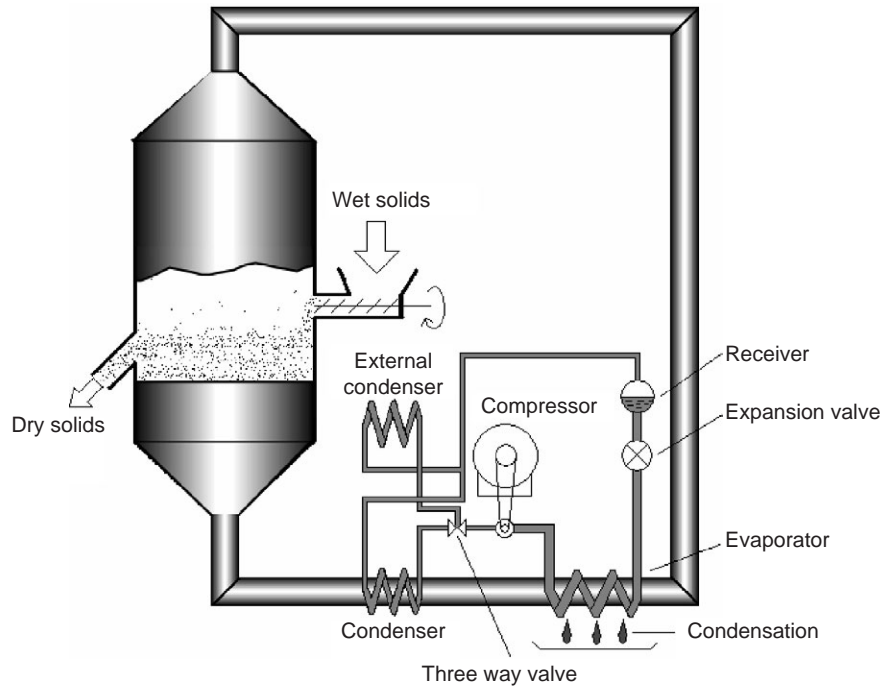


FIGURE 8.24 Heat pump fluidized bed dryer.

constant rate and the other for the falling rate. Table 8.4 shows the equations for calculating residence time at different operating conditions.

8.9.1.2 Sizing of Bed

Sizing of bed is based on simple hold-up mass balance. Cross-sectional area of the fluidized bed can be determined from the following equation after solids flow rate (dry basis), F_s , bed density, ρ_b , and bed height, H_b , are specified, and particle residence time, t_R , is determined:

$$A = \frac{F_s t_R}{\rho_b H_b} \quad (8.28)$$

8.9.1.3 Gas Flow Rate

Gas flow rate (dry basis) is calculated from the following equation. The operating gas velocity, u_g , is specified as a multiple of the minimum fluidization velocity, normally it is $2-3u_{mf}$ for fluidized bed drying. Anyway, the suitable operating gas velocity can be determined from laboratory-scale fluidized bed testing as long as the gas velocity yields good fluidization quality during the operation:

$$G_g = \rho_g u_g A \quad (8.29)$$

where ρ_g is the density of gas.

8.9.1.4 Mass Balance, Continuous Drying, Well-Mixed Bed

$$F_s(X_{in} - X_{out}) = G_g(Y_{out} - Y_{in}) \quad (8.30)$$

In this equation, F_s is the solids flow rate (kg/s), X is the moisture content (kg/kg), G_g is the gas flow rate (kg/s), and Y is the absolute humidity (kg/kg).

8.9.1.5 Heat Balance, Continuous Drying, Well-Mixed

Heat balance for the single-phase model gives the following energy balance:

$$F_s H_{s,in} + G_g H_{g,in} + Q_h = F_s H_{s,out} + G_g H_{g,out} + Q_w \quad (8.31)$$

In this equation, Q_h is the rate of heat input from immersed tubes (kJ/s), Q_w is the rate of heat loss from wall (kJ/s), and H is the enthalpy (kJ/kg). Enthalpy of solids at the inlet and outlet can be obtained from Equation 8.32 and Equation 8.33, respectively:

$$H_{s,in} = (c_{ps} + X_{in} c_l) T_{s,in} \quad (8.32)$$

$$H_{s,out} = (c_{ps} + X_{out} c_l) T_{s,out} \quad (8.33)$$

TABLE 8.4
Equations to Determine Residence Time Required for Drying

	Remarks	Drying Time Required
Batch Drying		
1. Constant rate period	Only surface moisture present	$t_R = \frac{(X_o - X)M_s \lambda}{G_g c_{pg}(T_{in} - T_{out})}$
2. Falling rate period	(i) From diffusion model	$\frac{X - X_{eq}}{X_o - X_{eq}} = \frac{6}{\pi^2} \sum_{n=1}^{\infty} \frac{1}{n^2} e^{[-(n\pi)^2 D t / R^2]}$ t_R is obtained by trial and error
	(ii) Simplified equation	$t_R = \frac{M_s c_{ps}}{G_g c_{pg}} \ln \left(\frac{T_p - T_{in}}{T_{po} - T_{in}} \right)$
	(iii) Empirical formulation	$t_R = \frac{1}{k} \ln \left(\frac{X_{cr1} - X_{eq}}{X - X_{eq}} \right)$
Continuous drying		
(a) Well-mixed	Design curve: $\bar{X} = \int_0^{\infty} X(t)E(t)dt$ [79]	
1. Constant rate period	Only surface moisture	$t_R = \frac{X_o - \bar{X}}{k}$
2. Falling rate period		$t_R = \frac{1}{k} \left(\frac{X_{in} - X_{eq}}{X_{out} - X_{eq}} - 1 \right) \quad [80]$
3. Batch drying curve	(1) Obtain a record of the changing bed temperature T_b during constant inlet air temperature run (2) Divide the constant inlet air temperature batch drying curve $X(t)$ into increments of length, ΔX . For each increment note the time Δt_{T_1} required to accomplish that amount of drying at constant bed temperature of T_1 (3) Calculate the time Δt_{T_2} required to accomplish the same increment of drying at constant bed temperature, T_2 by the use of the following equation: $\frac{\Delta t_{T_1}}{\Delta t_{T_2}} = \frac{[(p_{sat} - p_{in})(\bar{X} - X_e)]_{T_1}}{[(p_{sat} - p_{in})(\bar{X} - X_e)]_{T_2}} \quad [81,82]$ (4) Build up the constant bed temperature batch drying curve by increments (5) Obtain drying equation for each curve (6) Obtain residence time from design curve $t_R = \frac{SA}{f^* G \beta} \left(\frac{X_{in} - X_{out}}{X_{out} - X_{eq}} \right) \quad [83,84]$ where S is the bed loading, A is the bed area, f^* is ratio of bed loading (S) and flux of gas flow rate (G/A) at constant bed temperature, T_1	
(b) Plug flow		
1. Batch drying curve	Residence time distribution function is $E(t) = \frac{1}{2\sqrt{\pi B}} \exp \left[-\frac{(1 - t/\tau_m)^2}{4B} \right]$ where $B = \frac{D\tau_m}{L^2}$ and $D = \frac{3.71 \times 10^{-4}(u - u_{mf})}{u_{mf}^{1/3}} \quad [82];$ $D = \frac{1.49[0.01(H_b - 0.05) + 0.00165\rho_s(u - u_{mf})]u^{0.23}}{u_{mf}^{2/3}} \quad [83]$ Note that validity of Reay's correlation for particle diffusivity has only been confirmed for bed depths up to 0.10 m. There is some evidence in the literature that D may be an order of magnitude larger in much deeper beds [50]. Shallow bed is recommended if the objective is a close approach to plug flow behavior of solid particles in the bed	

For a gas–vapor system, H_{g-in} and H_{g-out} can be obtained from Mollier diagram and for organic vapor–inert gas systems, H_{g-in} and H_{g-out} can be obtained from the following equations:

$$H_{g-in} = (c_{pg} + Y_{in}c_1)T_{g-in} + Y_{in}\lambda \quad (8.34)$$

$$H_{g-out} = (c_{pg} + Y_{out}c_1)T_{g-out} + Y_{out}\lambda \quad (8.35)$$

A summary of steps for fluidized bed dryer design is given in [Figure 8.25](#), whereas a simple guide for selecting suitable fluidized bed dryers (FBD) based on material properties is given in [Figure 8.26](#).

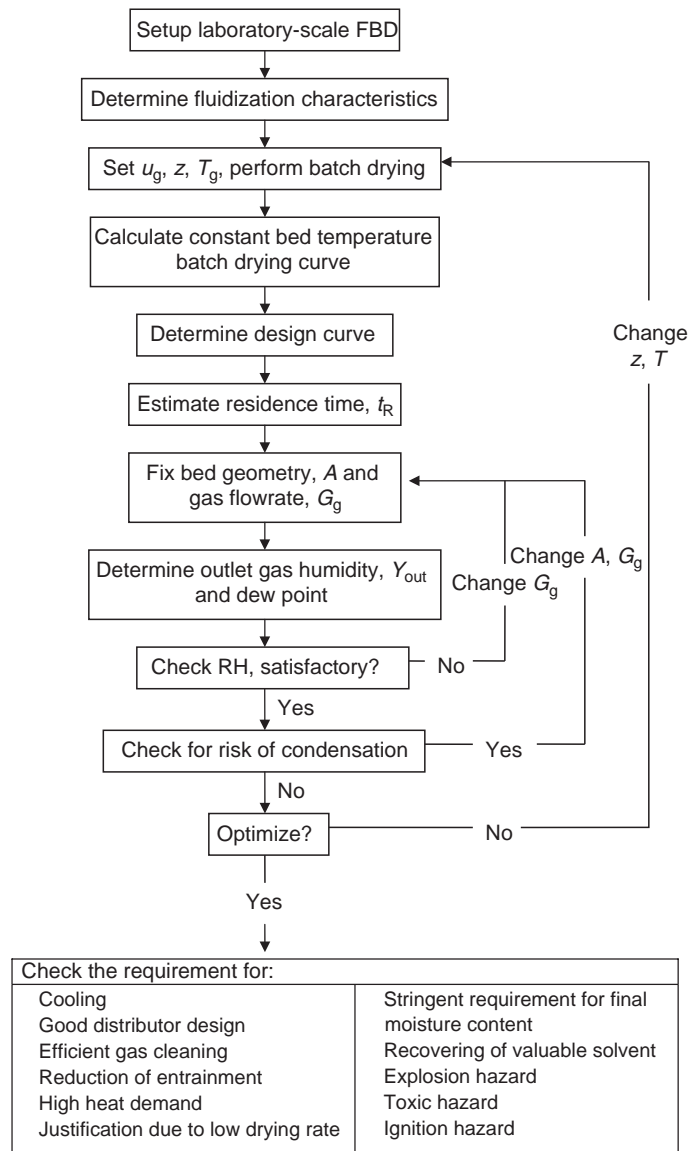


FIGURE 8.25 Design steps starting from laboratory tests.

8.9.2 A SAMPLE DESIGN CALCULATION

Wet particulate solids (6000 kg/h) with an initial moisture content of 20% (db) at 20°C are to be dried to final moisture content of 4% (db). Inlet air at 125°C with humidity of 0.005 kg/kg is used. Bed depth is 20 cm.

Under these conditions, bed density is 500 kg/m³ and equilibrium moisture content is zero. Specific heat of the dry solids and liquid water are 1.0 and 4.2 kJ/kg°C, respectively. Figure 8.27 gives the summary of all available data on this problem. Heat loss at the wall of the dryer is estimated as 5% of the heat content of inlet air. Batch drying curve was obtained at the conditions mentioned above; the relationship

between drying rate and moisture content is given by the following equation:

$$-\frac{dX}{dt} = 0.005X$$

Calculate

- Mean particle residence time
- Bed area
- Mass flow rate of air
- Absolute humidity of exhaust air
- Temperature of exhaust air
- Check whether condensation will occur in cyclone

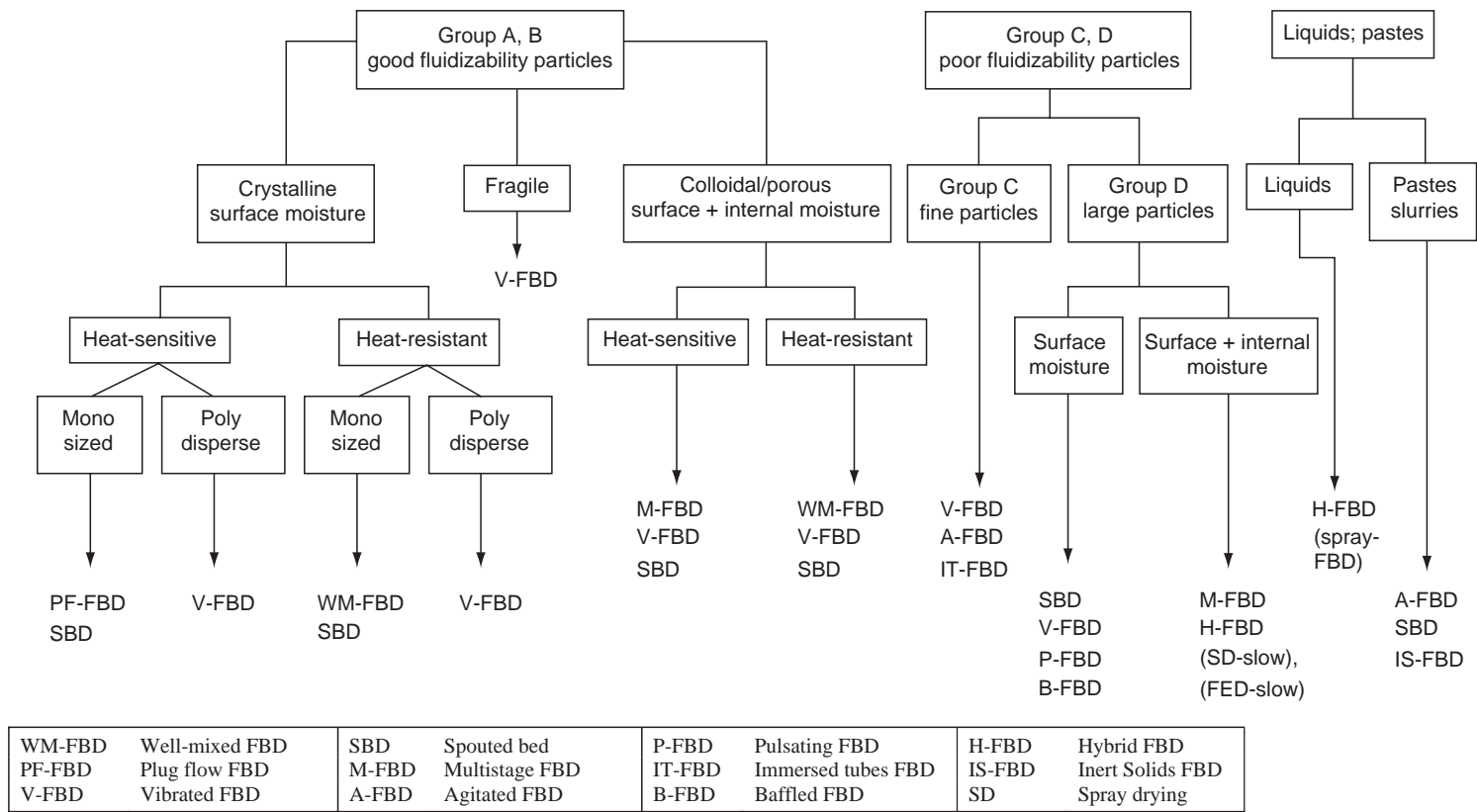


FIGURE 8.26 Dryer selection.

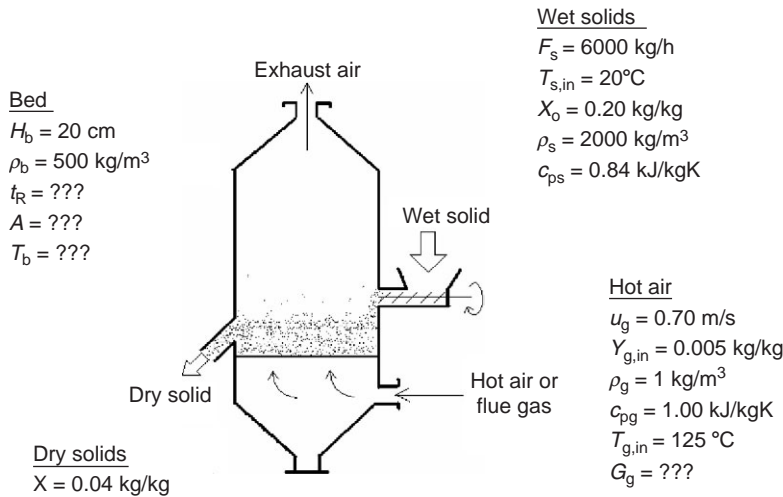


FIGURE 8.27 Sample calculation.

Solution

For continuous, well-mixed dryer operation in the linear falling period, the residence time is given by

$$t_R = \frac{1}{k} \left(\frac{X_o - X_{eq}}{\bar{X} - X_{eq}} - 1 \right)$$

$$t_R = \frac{1}{0.005} \left(\frac{0.20 - 0}{0.04 - 0} - 1 \right)$$

$$t_R = 800 \text{ s}$$

Dry solid mass flow rate is calculated from wet solid mass flow rate, wet solid has initial moisture content of 20%,

$$F = 6000 \frac{\text{kg wet solid}}{\text{h}} \times \frac{1 \text{ kg dry solid}}{1.20 \text{ kg wet solid}} \times \frac{1 \text{ h}}{3600 \text{ s}}$$

$$F = 1.389 \text{ kg dry solid/s}$$

Bed area is given by

$$A = \frac{F_s t_R}{\rho_b H_b}$$

$$A = \frac{1.389 \times 800}{500 \times 0.20}$$

$$A = 11.11 \text{ m}^2$$

Mass flow rate of air is calculated from the equation

$$G_g = \rho_g u_g A$$

$$G_g = 1.0 \times 0.70 \times 11.11$$

$$G_g = 7.777 \text{ kg/s}$$

Outlet air humidity can be obtained from the equation

$$F_s(X_{in} - X_{out}) = G_g(Y_{out} - Y_{in})$$

$$Y_{out} = \frac{F_s}{G_g}(X_{in} - X_{out}) + Y_{in}$$

$$Y_{out} = \frac{1.389}{7.777}(0.20 - 0.04) + 0.005$$

$$Y_{out} = 0.0336 \text{ kg H}_2\text{O/kg dry air}$$

Outlet air temperature can be obtained from the equation

$$F_s H_{s,in} + G_g H_{g,in} + Q_h = F_s H_{s,out} + G_g H_{g,out} + Q_w$$

$Q_h = 0$ since there is no immersed tube.

$Q_w = 0.05 G_g H_{g,in}$ assumes that heat loss to wall is taken as 5% of the enthalpy of inlet air:

$$H_{s,in} = (c_{ps} + X_{in} c_1) T_{s,in}$$

$$H_{s,in} = (0.84 + 0.20 \times 4.2) \times 20$$

$$H_{s,in} = 33.60 \text{ kJ/kg}$$

$$H_{s,out} = (c_{ps} + X_{out} c_1) T_{s,out}$$

$$H_{s,out} = (0.84 + 0.04 \times 4.2) T$$

$$H_{s,out} = 1.008 T \text{ kJ/kg}$$

$$H_{g,in} = (c_{pg} + Y_{in} c_1) T_{g,in} + Y_{in} \lambda$$

$$H_{g\text{-in}} = (1.00 + 0.005 \times 4.2) \times 125 + 0.005 \times 2370$$

$$H_{g\text{-in}} = 139.5 \text{ kJ/kg}$$

$$H_{g\text{-out}} = (c_{pg} + Y_{out}c_1)T_{g\text{-out}} + Y_{out}\lambda$$

$$H_{g\text{-out}} = (1 + 0.0336 \times 4.2)T + 0.0336 \times 2370$$

$$H_{g\text{-out}} = 1.141T + 79.63$$

$$F_s H_{s\text{-in}} + G_g H_{g\text{-in}} + Q_h = F_s H_{s\text{-out}} + G_g H_{g\text{-out}} + Q_w$$

$$\begin{aligned} F_s H_{s\text{-in}} + G_g H_{g\text{-in}} + Q_h \\ = F_s H_{s\text{-out}} + G_g H_{g\text{-out}} + 0.05 G_g H_{g\text{-in}} \end{aligned}$$

$$F_s H_{s\text{-in}} + 0.95 G_g H_{g\text{-in}} + Q_h = F_s H_{s\text{-out}} + G_g H_{g\text{-out}}$$

$$\begin{aligned} 1.389 \times 33.6 + 0.95 \times 7.777 \times 139.5 + 0 \\ = 1.389 \times (1.008T) + 7.777 \times (1.141T + 79.63) \end{aligned}$$

$$T = 44.6^\circ\text{C}$$

From the psychrometric chart, air at absolute humidity of 0.0336 kg/kg has a dew point of 33.5°C and relative humidity is 55%. Since the outlet air leaves the dryer at 44.6°C (10°C higher than the dew point), there is no risk of condensation.

8.10 CONCLUSION

Fluidized bed dryers have replaced some of the conventional dryers, e.g., rotary or conveyor dryers in many instances. Among some of the recent developments in fluidized bed drying is the idea of applying microwave energy field continuously or intermittently in a fluidized or spouted bed. Use of superheated steam will probably become more popular in some applications in the future.

The presence of moisture on particle surface can cause major changes to fluidization quality as compared with that of dry particles to which most of the available literature is applicable. Also, there are numerous variants of the fluidized bed that require different design information and design strategies. For example, drying using a time-dependent heat input or drying under low-pressure conditions or using superheated steam as the drying medium, etc. must be handled with some modifications and new data sets and models.

NOTATION

a, b, x, y, k	drying constant, drying coefficient
A	area, m ²
c	heat capacity, J/(kg K)

c_p	heat capacity at constant pressure J/(kg K)
d	diameter, m or μm
D	diffusivity, m ² /s
e	emissivity
$E(\theta)$	residence time density
F	solids mass flow rate, kg/s
g	gravity acceleration = 9.80665 m/s ² , m/s ²
G	gas mass flow rate, kg/s
H	height, m
h	heat transfer coefficient, W/(m ² K)
K_c	mass transfer coefficient across bubble boundary, m/s
k	thermal conductivity, W/(m K)
L	length, m
M	mass, kg
\dot{m}	mass rate of evaporation of water per unit volume of bed, kg/(m ³ s)
n	integer
Nu	Nusselt number
p	partial pressure, Pa
P	pressure, Pa
Pr	Prandtl number
q	rate of heat transfer, W
Q_h	heat input from immersed tubes, kJ/s
Q_w	heat loss to column wall, kJ/s
r	radius, m
Re	Reynolds number
t	time, s
T	temperature, °C, K
u	velocity, m/s
V	volume, m ³
X	moisture content, kg/kg
Y	absolute humidity, kg/kg

GREEK SYMBOLS

β	root of Bessel function
ε	void fraction
λ	latent heat of vaporization, J/kg
μ	viscosity, Ns/m ²
ρ	density, kg/m ³
σ	Stefan-Boltzmann constant = 5.67×10^{-8} W/ (m ² K ⁴), W/(m ² K ⁴)
ϕ	sphericity

SUBSCRIPT

b	bed
bb	bubble
c	convective component
cr1	first critical
cyl	cylinder
d	dense

eff	effective
eq	equilibrium
evap	evaporation
g	gas
in	inlet
ind	induction
l	liquid
mf	minimum fluidization
o	initial
opt	optimum
out	outlet
p	particle
r	radiant component
R	residence
s	solids
sat	saturated
sph	sphere
t	column
v	vapor
w	wall

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9 Drum Dryers

Wan Ramli Wan Daud

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9.1 INTRODUCTION

The drum dryer is commonly used to dry viscous, concentrated solutions, slurries or pastes on rotating steam-heated drums.^{1,2} It can also be used to dry concentrated solutions or slurries that become more viscous or pasty because of flashing or boiling off of moisture or of irreversible thermochemical transformations of their content that occur on their first contact with the hot drum surface.³⁻⁵

The viscous slurry or paste is mechanically spread by the spreading action of two counter-rotating drums into a thin sheet that adheres on the hotter drum in single drum dryers or split sheets on both hot

cylinders in double drum dryers. The adhering thin sheet of paste is then rapidly dried conductively by the high heat flux of the condensing steam inside the drum. For very wet slurries that produce wet sheets, the drying of the wet thin sheet can be further enhanced by blowing hot dry air on the sheet surface. The thin sheet containing heat-sensitive materials, such as vitamins, can also be dried at a lower temperature in a vacuum.

The irreversible thermochemical transformations during the slurry's first contact with the hot drum can also be used to simultaneously impart certain required quality of the dried product.⁶ Starch slurries can be gelatinized or "cooked" before the sheet is

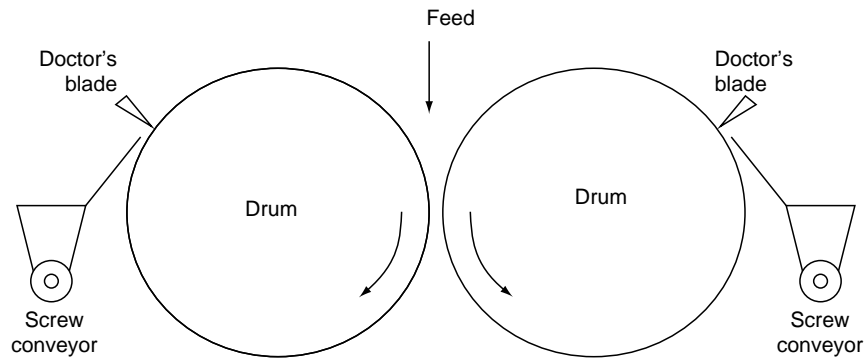


FIGURE 9.1 Double drum dryer with nip feed.

dried to produce pregelatinized or “precooked” starch for instant food formulations. Exposure of the thin sheet to the high heat flux and high temperature for a short period of time can also impart a porous structure to the dried sheet because of the rapid formation of vapor bubbles within the sheet during “boiling-like” drying. Porous products are excellent in instant food formulations because they are more readily wetted and can be easily rehydrated. It is for these reasons that the drum dryer is widely used around the world in the production of pregelatinized starch for instant food formulations.

9.2 TYPES OF DRUM DRYERS

The drum dryer was first patented for use in the manufacture of pregelatinized starch in Germany by Mahler and Supf in 1921. Since then a host of other patents have appeared especially in the United States where extensive variations of feeding methods, number and configuration of drums, heating system and product removal were considered. The diameter of the drum varies from 0.45 to 1.5 m and its length varies from 1 to 3 m. The thickness of the drum wall is between 2 and 4 cm. The drum dryer is classified

according to the number and configuration of the steam-heated drums and the pressure of the atmosphere around the drying sheet.

9.2.1 ATMOSPHERIC DOUBLE DRUM DRYER

This type of dryer has a higher production rate, can handle a wider range of products, and is more efficient.^{1–3,7} The slurry or paste is fed through a pendulum nozzle or through a header with multiple nozzles on to the nip of two steam-heated drums counter-rotating toward each other, forming a boiling pool at the nip (Figure 9.1). The feed can also be fed at the nips of applicator rollers and the drums (Figure 9.2). Starch slurries gelatinize in the boiling pool forming pastes that become more viscous. The counter-rotation of the drums spread the slurry or paste into two thin sheets on both drums that consequently dry conductively.

9.2.2 ATMOSPHERIC SINGLE DRUM DRYERS

The slurry or paste is fed through a pendulum nozzle or through a header with multiple nozzles similar to those of the double drum dryers, on to the nip of a

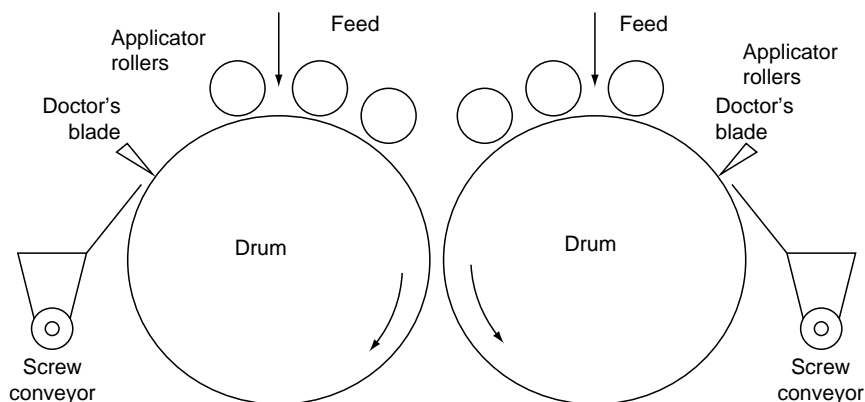


FIGURE 9.2 Twin drum dryer with applicator roller feeds.

steam-heated drum and a much cooler applicator roller counter rotating toward each other, forming a boiling pool at the nip (Figure 9.3).^{1-3,7} Starch slurries gelatinize in the boiling pool, forming pastes that become more viscous. The counter-rotation of the drum and applicator roller spread the slurry or paste into a thin sheet on the hot drum that consequently dries conductively. Alternatively, the slurry can be fed by dip coating a dip or applicator roller in a feed tray at the bottom of the dryer and then roller-coated on the drum (Figure 9.4). The slurry can also be fed by dip coating the drum directly in the feed tray (Figure 9.5) or sprayed or splashed from a feed tray (Figure 9.6).

9.2.3 ATMOSPHERIC TWIN DRUM DRYERS

The slurry is applied by direct dip coating of the twin drums in the feed tray at the bottom of the dryer (Figure 9.7) or by splash or spray feeders from a feed reservoir at the bottom of the dryer (Figure 9.8) on to the surface of the two steam-heated drums that are counter-rotating away from each other.^{1-3,7} The sheet is formed by adhesion on to the drum surface and is held up against gravity by its surface tension. The sheets consequently dry conductively. This type of dryer is suitable for solutions that produce a dusty product.

9.2.4 ENCLOSED DRUM DRYERS

If solvent vapor other than water released during drum drying needs to be recovered or if the dried products generate a lot of dust, atmospheric double- or twin-drum dryers can be enclosed in vapor or dust-tight enclosures.^{1-3,7} The vapor can be recovered by using a suitable condenser and the dust can be removed by using a wet scrubber.

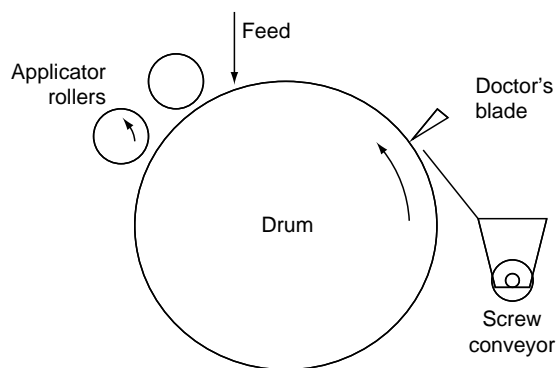


FIGURE 9.3 Single drum dryer with applicator roller feed.

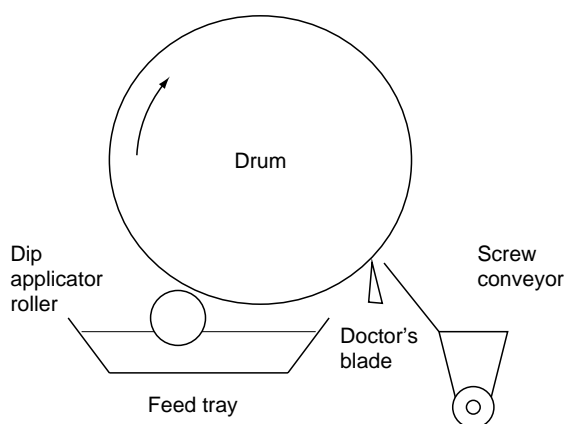


FIGURE 9.4 Single drum dryer with dip roller feed.

9.2.5 VACUUM DOUBLE DRUM DRYER

Heat-sensitive materials can be dried in a vacuum double drum dryer where the dryer is enclosed in an airtight enclosure under vacuum (Figure 9.9).^{1-3,7} This type of dryer is also fitted with a condenser, a scrubber, and a vacuum pump. The operation of the dryer is similar to its atmospheric version except that there are two product troughs, namely one for breaking the vacuum and the other for product discharge.

9.3 PRINCIPLES OF OPERATION OF THE DRUM DRYER

The drum dryer is a highly flexible equipment because its operational variables like the steam pressure, the drum rotational speed, the nip width, and the ratio of drum rotational speeds can be regulated independently. The steam pressure ranges from 2 to 7 bar, the drum rotational speed varies from 2 to 30 rpm, the nip width ranges from 0.05 to 0.5 mm, and the ratio

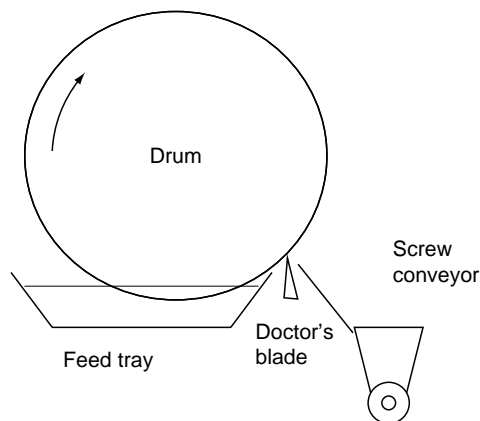


FIGURE 9.5 Single drum dryer with dip feed.

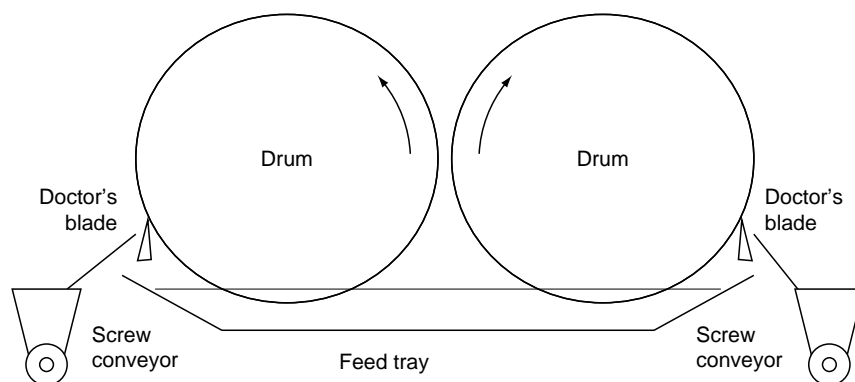


FIGURE 9.6 Twin drum dryer with dip feed.

of drum rotational speeds is from unity to 5. The feed can be preconcentrated and preheated to reduce the drying load but there is a limit to the feed concentration beyond which the sheet may not form well.^{1,2,8}

9.3.1 DRUM DRYER CAPACITY

The capacity of the drum dryer depends on the drying rate of the thin sheet, the amount of product in the sheet, and hence the sheet thickness and the rotation speed of the drums. The drying rate in turn depends on the sheet temperature and hence the steam pressure in the drum, the sheet material, and to a lesser extent the thickness of the sheet. The thickness of the sheet depends on the relative speeds of rotation of the drums, the depth of the boiling pool at the nip, the nip width, and the rheological properties of the liquid.

However, the wide range of material property values of the different feed materials, the different complex thermochemical processes occurring during drying that may change these property values further,

and the wide variety of drying characteristics of sheets of different material make the operation of the drum dryer very complex. In the past, the performance of a drum dryer drying a specific product cannot be adequately predicted by theoretical or semitheoretical models but must be based on drying performance test of the product on a pilot plant drum dryer.

9.3.2 STEAM CONSUMPTION

Typical specific steam consumption of the drum dryer varies from 1.3 to 1.5 kg steam per kg water removed or a steam economy of 0.66 to 0.76 kg water removed per kg steam.¹⁻⁴ It means that the specific heat consumption is typically about 3000 to 3500 kJ/kg water removed. The specific evaporation rate is 10 to 30 kg water evaporated per m²/h for difficult to dry materials and 40 to 50 kg water evaporated per m²/h for easy to dry materials. Recent studies have increased the understanding of the processes in the drum dryer but it is still a long way before the drum dryer can be modeled completely.

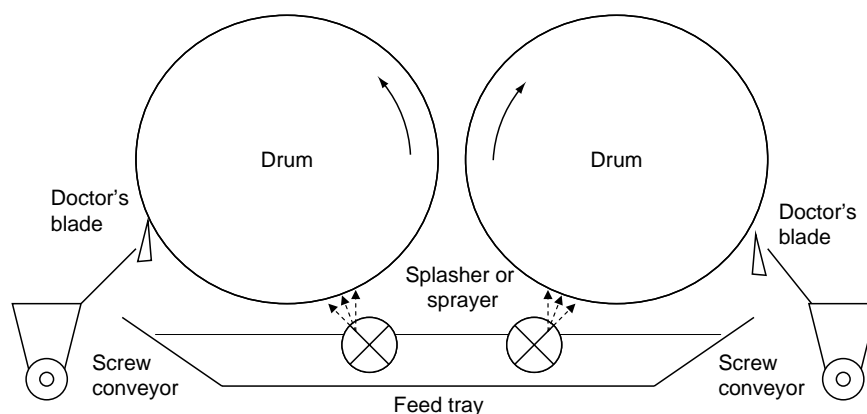


FIGURE 9.7 Twin drum dryer with splasher or sprayer feed.

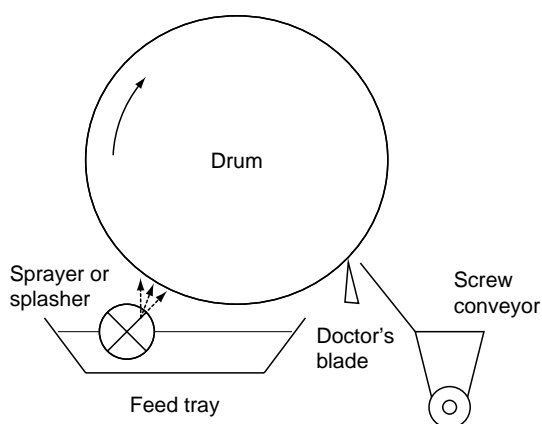


FIGURE 9.8 Single drum dryer with splasher or sprayer feed.

9.3.3 FEEDING AND SPREADING OF LIQUID INTO A THIN SHEET

9.3.3.1 Nip Feeding

Double and single drum dryers are usually fed at the nip between the drums by either a pendulum nozzle or multiple nozzles in a header (Figure 9.1 through Figure 9.3).^{1-3,7} The sudden exposure of the liquid pool to the high temperature and intense heat flux causes the pool to boil and may also cause irreversible thermochemical transformations of the liquid content that may also change the rheological property of the liquid pool. For example, the gelatinization of starch in the boiling pool of starch slurry changes the Newtonian slurry into a non-Newtonian, shear-thinning power law paste. It also swells up several times its original volume. In this case, the boiling pool is transformed into a rotating, asymmetric cylinder of paste at the nip.

The counter-rotation of the drums toward each other draws the liquid pool into the nip and spreads it into a thin sheet that adheres to the hot drum in a

single drum dryer. In the double drum dryer the spreadsheet is split into two sheets that adhere on to both hot drums. Theoretical and empirical studies on the spreading phenomenon show that for a given rotational speed ratio of the drums, the spreadsheet thickness reaches an asymptotic level for a critical height of the boiling pool.^{9,10} Although the asymptotic thickness is proportional to the nip width, it also decreases with increasing rotational speed ratio, and to a lesser extent with increasing diameter ratio of the smaller to the larger drum and the ratio of the nip width and the smaller drum. This means that in practice, the nip can be flooded with the feed beyond the critical height and the sheet thickness is controlled by simply varying the nip width.⁵ For single drum dryers, the asymptotic thickness of the liquid sheet is about 1.2 times the nip width for Newtonian liquids and up to 1.26 times the nip for shear-thinning liquids. For double drum dryers, both the speed and diameter ratios are usually unity and the ratio of the nip width to the drum is very small. The liquid sheet is then split into two sheets of equal thickness, one for each drum.

The gelatinization of starch slurries into a shear-thinning power law paste also changes the spreading and the mixing regime in the pool.¹¹ The sheet is thinner for more shear-thinning liquid ($n < 1.0$). The paste material near the surface of the rotating drums experiences a higher shear rate than those further away in the bulk of the pool. The paste nearer the surface is therefore less viscous and readily rotates with the drum. On the other hand, the bulk of the paste is more viscous and may not move at all. This means that the mean residence time of the paste pool is longer than a corresponding Newtonian liquid pool. Heat-sensitive materials therefore degrade to a larger extent in the paste pool. Empirical studies have shown that the residence time distribution of the shear-thinning

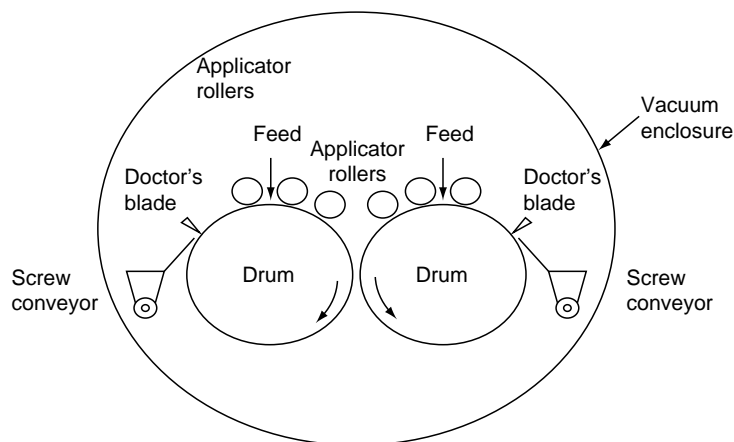


FIGURE 9.9 Vacuum double drum dryer.

paste can be modeled by the Chollete and Coulter model consisting of an equivalent stirred tank and a parallel bypass with a dead region connected to the former.^{12–14} Hence for heat-sensitive materials, the mean residence time can be reduced by having a smaller pool at the expense of losing control of the sheet thickness if the pool height is below the critical height for asymptotic sheet thickness. Alternatively the heat-sensitive material can be fed using roller feeding that does not have a large volume of liquid holdup at high temperature as in nip feeding.

9.3.3.2 Roller Feeding

Single drum dryers are often fed by a cooler applicator roller.^{1–3,7} The liquid can be fed from the top at the nip between an applicator roller and the hot drum and the liquid sheet is formed by spreading (Figure 9.3). The process is similar to nip feeding (see Section 9.3.3.1). The liquid sheet can be further spread by a succession of rollers to ensure that the thickness of the sheet is uniform.

Alternatively the applicator roller can be dip-coated with a liquid sheet from a dipping reservoir or bath below the drum (Figure 9.4).^{1–3,7} The roller then transfers some of the liquid to the counter-rotating drum above it, retaining a thinner sheet that goes back into the bath. This feeding method is called roller feeding and is similar to roller coating.¹⁵ The splitting of the liquid sheet is dependent on the ratio of rotational speeds (N_D/N_R), the ratio of the drum and roller diameters (D_D/D_R), and the index of the power law fluid, n , given by¹⁵

$$\frac{T_D}{T_R} = \left(\frac{N_D}{N_R} \frac{D_D}{D_R} \right)^{2n/(1+2n)} \quad (9.1)$$

The thickness of the liquid sheet on the drum is independent of the pressure applied between the applicator roller and the drum. It depends largely on the viscosity and the surface tension of the liquid as well as on the ratio of rotational speeds of the drum and roller and the nip width. A stable thickness can be achieved for medium viscosity shear-thinning liquids at moderate Reynolds number, Re (the ratio between momentum and viscous forces), low capillary number, Ca (the ratio of the viscous force and surface tension), and low rotational speed ratio.^{15,16} A smaller Ca (<0.01) or a larger surface tension extends coating stability to higher rotation speeds. A stable sheet is established between a minimum and a maximum drum speed and at a critical ratio of rotational speeds of the applicator roller and the drum. The sheet formed beyond the critical point is unstable, leading to ribbing and may entrain air bubbles as

well especially if the liquid contains excessive surfactants.^{15,16} This rarely occurs in drum dryer operation because both the rotation speed and the ratio of rotation speeds are low (linear speed of drum surface rarely exceeds 0.3 m/s).

9.3.3.3 Dip Feeding

Single and double drum dryers can also be dip-coated directly from a reservoir of liquid (Figure 9.5 and Figure 9.6).^{1–3,7} Dip coating depends on the surface tension, viscosity, density, and wall adhesion of the liquid as well as on the angle of immersion and rotation speed of the drum.^{15,17,18} If the drum is immersed deeper in the bath and the angle of immersion is therefore larger, the sheet would be thinner. Apart from Re and the Ca , dip coating also depends to a lesser extent on the Froude number (the ratio of momentum to gravity forces). Higher Re (>1.0) and higher rotation speed yield thinner sheets (>0.5 m/s). Above a certain critical Re (>1), ca (>0.2), and drum rotational speed (linear speed >0.5 m/s), the sheet breaks up into rivulets or ribs.^{15,17,18} A smaller ca or a larger surface tension would extend the critical Re and rotation speed. This rarely happens in drum dryer operation because both the rotation speed and the ratio of rotation speeds are low. The linear speed of the drum surface rarely exceeds 0.3 m/s.

9.3.3.4 Spray and Splash Feeding

The slurry or solution can be fed by spraying or splashing of the liquid from a feed tray at the bottom of the drum (Figure 9.7 and Figure 9.8). Spraying and splashing is a highly inefficient coating technique because most of the spray or splash droplets will bounce back into the feeding tray. In the drum dryer, it would also drools back into the tray due to gravity. The spray efficiency can be as low as 20 to 30%, which means that 70 to 80% of sprayed or splashed liquid fall back to the feed tray. An adhesive liquid or a liquid that becomes adhesive on contact with the hot drum surface may have a higher spraying efficiency. The liquid sheet produced is also thinner than those produced by other comparable spreading techniques.

9.3.4 CONDUCTIVE OR CONTACT DRYING OF THIN SHEETS

9.3.4.1 Initial Application Zone

The sudden exposure of the liquid feed to the high temperature and intense heat flux on the drum surface at the application zone causes it to immediately heat up and boil (Figure 9.10). Most of the free moisture is

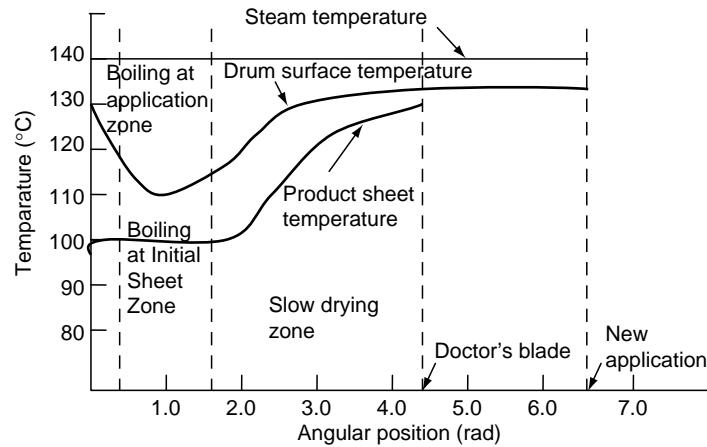


FIGURE 9.10 Temperature profile of the drum dryer.

evaporated during this initial boiling of the feed (Figure 9.11).^{5,19} The boiling removes a large amount of heat from the hot drum surface due to the large latent heat of vaporization of moisture. Since the temperature difference between the feed slurry and the drum is of the order of 30°C, the slurry boils on contact with the drum surface very near the critical point of boiling. The critical boiling flux, E in W/m^2 is given by²⁰

$$E = 2.177\rho_v \left[\frac{\sigma g(\rho_l - \rho_v)}{\rho_v^2} \right]^{0.25} \quad (9.2)$$

where ρ_v is the density of steam (kg/m^3), ρ_l is the density of water (kg/m^3), σ is the surface tension of water (N/m), and g is gravitational constant (m/s^2). The correlation of heat transfer of vapor condensing inside horizontal tubes is given by²¹

$$h_s = 0.555 \left[\frac{g\rho_l(\rho_l - \rho_s)\Delta H_v\lambda_l^3}{\mu_l D_i(T_s - T_i)} \right]^{1/4} \quad (9.3)$$

where μ_l is the viscosity of liquid (Ns/m^2), D_i is the inner diameter of drum (m), T_s is the steam temperature ($^{\circ}\text{C}$), T_i is the inner drum wall temperature ($^{\circ}\text{C}$), ΔH_v is the latent heat of vaporization of steam (J/kg), and λ_l is the thermal conductivity of water ($\text{W/m}^{\circ}\text{C}$). The overall heat transfer coefficient in this zone was reported to be between 2000 and 7000 $\text{W/m}^2\text{ }^{\circ}\text{C}$.^{6,19} The temperature of the drum surface then drops due to the large removal of heat in the boiling to an extent proportional to the drum rotation speed. The faster the rotation, the smaller is the temperature drop. The temperature of the liquid remains at the boiling point of the solution, which is slightly above the normal boiling point of water due to the presence of dissolved solids.

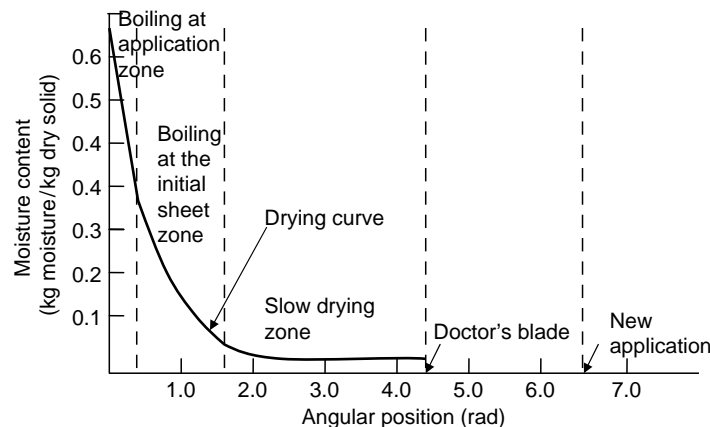


FIGURE 9.11 Moisture content (dry basis) profile of the drum dryer.

9.3.4.2 Initial Sheet Zone

The drying regime of the thin sheet of liquid or paste on the drum surface is dictated by the high temperature and the large heat flux supplied by condensing steam inside the drum (Figure 9.10).^{22–26} The heat flux can be as high as 85 kW/m².²⁷ The overall heat transfer coefficient in this zone was reported lower than at the application zone and is between 600 and 1250 W/m² °C.^{6,19}

Moisture transport in the sheet is predominantly driven by the large temperature gradient and the subsequently large pressure gradient within the sheet (Figure 9.11).^{5,22–24,27} Excess surface water in the wet sheet flashes or boils off the sheet and the temperatures of the drum surface continue to fall whereas the product sheet temperature remains constant.¹⁹ The evaporation flux at this point is given by³

$$E_1 = 8.64554 \times 10^{-8} u_a^{0.8} (P_p - P_a) \quad (9.4)$$

where u_a is the velocity of air over the surface of sheet (m/s), P_p is the vapor pressure at product surface (Pa), and P_a is the vapor pressure at ambient temperature (Pa). As soon as the surface moisture dries up, vapor bubbles generated by boiling within the sheet then flash through the sheet. In gelatinized starch sheets, this drying regime tends to form pores within the sheet and craters at the surface.^{8,28}

9.3.4.3 Slow Drying Zone

The sheet becomes quite dry very quickly and the temperature of the drum surface as well as that of the sheet start to rise since the thermal capacitance of the drum wall is larger than that of the thin sheet and there is little moisture left to lower the temperature further by boiling. Tang et al.⁷ suggests that at this point, the evaporation rate can be estimated by

$$E_1 = 3.6 \times 10^3 h (T_o - T_e) / \Delta H_v \quad (9.5)$$

where h is a coefficient having a value of between 200 and 2000 W/m² °C, depending on the product and its thickness, T_o is the temperature of outer drum surface in °C, and T_e is the temperature of evaporating product surface in °C.

9.3.4.4 Vacuum Drying

The principle of operation of the vacuum drum drying is similar to that of the atmospheric drum drying process as described in the previous section except that both the pressure and temperature of operation are lower.^{1–3,29} A heat-sensitive material dried on a

vacuum drum dryer suffers less damage and retains most of its structure and functionality than that dried in ordinary atmospheric drum dryer.³⁰

9.4 DRY PRODUCT HANDLING

The dry sheet is then scrapped off by a doctor's blade or knife located at about three quadrants away from the feed point. After the sheet is removed by the knife, the temperature of the bare metal rises slightly further until the feed point is reached. The sheet falls into a product trough below the knife and a screw conveyor in the trough transport the dry products away to a milling process where the sheet is crushed to form a powder product.

9.5 CONTROL OF DRUM DRYERS

The control strategy of the drum dryer in the past has always been the basic control for steady operation only without any on-line quality control. The set points for the control variables are developed by trial and error for each product. The temperature of the drum can be controlled independently for the required drying task by setting the steam pressure. In vacuum drum dryers, the pressure of the enclosure and thus the temperature of drying can also be controlled by setting the vacuum pressure. In most cases, the sheet thickness is controlled by setting the nip width and the feed flow rate for asymptotic sheet thickness. The final moisture content is then controlled by varying the drum speed. In cases where the sugar content is high enough for the final sheet to be in a glassy state and sticky, the final temperature and drying rate of the sheet just before the doctor's blade is controlled by blowing cool dry air.

This strategy is adopted because of the sheer complexity of the process and the unavailability in the past of suitable sensors for on-line quality moisture content measurement. The quality of the product such as final moisture content, thickness, porosity, wetting, and rehydration capability (for pregelatinized starch) as well as the right crystal structure (the right therapeutic form for pharmaceuticals) are complex functions of drum speed, temperature, nip width, feed material, feed concentration, and feed-spreading technique. In addition, the final moisture content and thickness of the sheet may not be uniform across the width of the drum dryer that can lead to problems in shelf life and packaging of the product, respectively.

Dynamic drum dryer models that have been developed so far can only predict the final temperature and moisture content.^{31–34} These models cannot

predict other important quality parameters and are only useful for steady process operation. With the advent of improved infrared technology, the moisture content can be measured by inference using infrared temperature sensors.^{35,36}

9.6 DRUM-DRIED PRODUCTS

Products that are suitable for drying on a drum dryer are viscous liquids, slurries, suspensions, and pastes. The final dry products are typically in the form of porous flakes or powders. The drum dryer has been used extensively to dry chemicals and food products. Chemicals that have been successfully dried on a drum dryer are polyacrylamides, and various salts such as silicate, benzoate, propionate, and acetate salts.^{1,2} Drum dryers have been successfully used in drying sludge.³⁷

The drum dryer is also extensively used to dry and gelatinize or “cook” starch slurries, such as potato,³⁸ rice,^{5,39} wheat,^{40–42} maize,^{43,44} corn,⁴⁵ soybean-banana,⁴⁶ and cowpea⁴⁷ slurries to produce pregelatinized starch for instant foods. Nonstarch, low-sugar foods, such as tomato puree, milk, skim milk, whey,⁴ beef broth, yeast,⁴⁸ coffee, and malt extract, have also been successfully dried on a drum dryer.^{1,2} Heat-sensitive products such as pharmaceuticals³⁰ and vitamin-containing products³ can be dried in a vacuum drum dryer.

Sugar-containing slurries, such as apple puree,⁴⁹ apple sauce, citrus pulps,⁵⁰ and other fruit juice, have also been successfully dried on drum dryers. However if the sugar content is high, some of the sugar does not crystallize properly as drying proceeds but becomes molten instead at well above the glass transition temperature.⁷ The uneven scrapping of the doctor's blade at the rubbery and glassy parts of the sheet forms wrinkles in the sheet which eventually become “sticks” in the final product. The “sticks” reduce the quality of the product by making it very hard to disperse and physically unsatisfactory in appearance. The formation of sticks can be controlled by enhanced cooling and drying of the sheet at the doctor blades by a stream of dry, cool air, and by controlling the sheet thickness using takeoff rolls.⁷

NOMENCLATURE

LATIN SYMBOLS

T_D	sheet thickness on the drum m
T_R	sheet thickness on the roller m
N_D	rotational speed of drum, rpm
N_R	rotational speed of roller, rpm
D_D	diameters of drum, m
D_R	diameters of roller, m

n	the index of power law fluid
g	gravitational constant, m/s
D_i	inner diameter of drum, m
T_s	steam temperature, °C
T_i	inner drum wall temperature, °C
ΔH_v	latent heat of vaporization of steam, J/kg
u_a	velocity of air over the surface of sheet, m/s
P_p	vapor pressure at product surface, Pa
P_a	vapor pressure at ambient temperature, Pa
h	heat transfer coefficient, W/m ² °C
T_o	temperature of outer drum surface, °C
T_e	temperature of evaporating product surface, °C

GREEK SYMBOLS

λ_1	thermal conductivity of water, W/m °C
ρ_v	density of steam, kg/m ³
ρ_l	density of water, kg/m ³
σ	surface tension of water, N/m
μ_l	viscosity of liquid, Ns/m ²

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10 Industrial Spray Drying Systems

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10.1 INTRODUCTION

Spray drying is a suspended particle processing (SPP) technique that utilizes liquid atomization to create

droplets that are dried to individual particles when moved in a hot gaseous drying medium, usually air. It is a one-step continuous unit processing operation. It has become one of the most important methods

for drying the fluid foods in the Western world. The development of the process has been intimately associated with the dairy industry and the demand for drying of milk powders. Spray drying used in dairy industry dates back to around 1800, but it was not until 1850 that it became possible in industrial scale to dry the milk. However, this technology has been developed and expanded to cover a large food group that is now successfully spray-dried. Over 25,000 spray dryers are now estimated to be commercially in use to dry products from agrochemical, biotechnology products, fine and heavy chemicals, dairy products, dyestuffs, mineral concentrates to pharmaceuticals in capacities ranging from a few kg/h to over 50 tons/h evaporation capacity.

The spray drying process operates in the following way, i.e., the liquid is pumped from the product feed tank to the atomization device, such as the rotary disc atomizer, pressure nozzle, pneumatic nozzle, and ultrasonic nozzle, which is usually located in the air distributor at the top of the drying chamber. The drying air is drawn from the atmosphere through a filter by a supply fan and is passed through the air heater, e.g., oil furnace, electrical heater, steam heater etc., to the air distributor. The droplets produced by the atomizer meet the hot air and the evaporation takes place cooling the air in the meantime. After the drying of the droplets in the chamber, the majority of the dried product falls to the bottom of the chamber and entrains in the air. Then they pass through the cyclone for separation of the dried particles from the air. The particles leave the cyclone at the bottom via a rotary valve and are collected or packed later. The fine particles will remain entrained in the air and then pass the air through scrubbers for washing out the fine particles from the air. The cycled washing liquid can be transported to the pretreat process of the feed. The air passes from the scrubber to the atmosphere via the exhaust fan. The process control system may comprise the necessary indication of the temperatures at the air inlet and outlet, the voltages and currents of the motors, etc. The automatic controls often are to maintain the inlet temperature by altering the heater operating parameters, such as the steam pressure, the amount of oil to the heater, etc., and the outlet temperature by adjusting the amount of feed pumped to the atomizer.

The advantages of spray dryers are that this technique can

- Handle heat sensitive, non-heat-sensitive, and heat-resistant pumpable fluids as feedstocks from which a powder is produced
- Produce dry material of controllable particle size, shape, form, moisture content, and other

specific properties irrespective of dryer capacity and heat sensitivity

- Provide continuous operation adaptable to both conventional and PLC control
- Handle wide range of production rates, i.e., any individual capacity requirement can be designed by spray dryers
- Provide extensive flexibility in spray dryer design, such as drying of organic solvent-based feedstocks without explosion and fire risk; drying of aqueous feedstocks (where the resulting powders exhibit potentially explosive properties as a powder cloud in air); drying of toxic materials; drying of feedstocks that require handling in aseptic and hygienic drying conditions; drying of liquid feedstocks to granular, agglomerated, and nonagglomerated products

However, they also suffer some limitations, such as

- High installation costs
- A lower thermal efficiency
- Product deposit on the drying chamber may lead to degraded product or even fire hazard

Examples of spray-dried products on industrial scale include the following:

- Chemical industry, e.g., phenol-formaldehyde resin, catalysts, PVC emulsion-type, amino acids, etc.
- Ceramic industry, e.g., aluminium oxide, carbides, iron oxide, kaolin, etc.
- Dyestuffs and pigments, e.g., chrome yellow, food color, titanium dioxide, paint pigments, etc.
- Fertilizers, e.g., nitrates, ammonium salts, phosphates, etc.
- Detergent and surface-active agents, e.g., detergent enzymes, bleach powder, emulsifying agents, etc.
- Food industry, e.g., milk, whey, egg, soya protein, etc.
- Fruits and vegetables, e.g., banana, tomato, coconut milk, etc.
- Carbohydrates, e.g., glucose, total sugar, maltodextrine, etc.
- Beverage, e.g., coffee, tea, etc.
- Pharmaceuticals, e.g., penicillin, blood products, enzymes, vaccines, etc.
- Biochemical industry, e.g., algae, fodder antibiotic, yeast extracts, enzymes, etc.
- Environmental pollution control, e.g., flue gas desulfurization [21], black liquor from paper-making, etc.

10.2 PRINCIPLES OF SPRAY DRYING PROCESSES

10.2.1 GENERAL

The spray drying process transforms a pumpable fluid feed into a dried product in a single operation. The fluid is atomized using a rotating wheel or a nozzle, and the spray of droplets immediately comes into contact with a flow of hot drying medium, usually air. The resulting rapid evaporation maintains a low droplet temperature so that high drying air temperatures can be applied without affecting the product. The time of drying the droplets is very short in comparison with most other drying processes. Low product-temperature and short drying-time allow spray drying of very heat-sensitive products.

Spray drying is used to dry pharmaceutical fine chemicals, foods, dairy products, blood plasma, numerous organic and inorganic chemicals, rubber latex, ceramic powders, detergents, and other products. Some of the spray-dried products are listed in [Table 10.1](#), which also includes typical inlet and outlet moisture content and temperatures together with the atomizer type and spray dryer layout used.

The principal advantages of spray drying are as follows:

1. Product properties and quality are more effectively controlled.
2. Heat-sensitive foods, biologic products, and pharmaceuticals can be dried at atmospheric pressure and low temperatures. Sometimes inert atmosphere is employed.
3. Spray drying permits high-tonnage production in continuous operation and relatively simple equipment.
4. The product comes into contact with the equipment surfaces in an anhydrous condition, thus simplifying corrosion problems and selection of materials of construction.
5. Spray drying produces relatively uniform, spherical particles with nearly the same proportion of nonvolatile compounds as in the liquid feed.
6. As the operating gas temperature may range from 150 to 600°C, the efficiency is comparable to that of other types of direct dryers.

Among the disadvantages of spray drying are the following:

1. Spray drying fails if a high bulk density product is required.
2. In general it is not flexible. A unit designed for fine atomization may not be able to produce a coarse product, and vice versa.

3. For a given capacity larger evaporation rates are generally required than with other types of dryers. The feed must be pumpable.
4. There is a high initial investment compared to other types of continuous dryers.
5. Product recovery and dust collection increase the cost of drying.

Spray drying consists of three process stages:

1. Atomization
2. Spray-air mixing and moisture evaporation
3. Separation of dry product from the exit air

Each stage is carried out according to the dryer design and operation and, together with the physical and chemical properties of the feed, determines the characteristics of the final product. A typical example of a spray drying process with the most important ancillary equipment included is shown in [Figure 10.1](#).

10.2.2 ATOMIZATION

Atomization is the most important operation in the spray drying process. The type of atomizer not only determines the energy required to form the spray but also the size and size distribution of the drops and their trajectory and speed, on which the final particle size depends. The chamber design is also influenced by the choice of the atomizer. The drop size establishes the heat transfer surface available and thus the drying rate. A comparison of spherical droplet surface and droplet size is shown in [Table 10.2](#).

Three general types of atomizers are available. The most commonly used are the rotary wheel atomizers and the pressure nozzle single-fluid atomizers [8]. Pneumatic two-fluid nozzles are used only rarely in very special applications. Existing spray drying systems provide various forms of the dry product—from fine powders to granules. The typical ranges of the disintegrated droplets and particle sizes of various products in a spray dryer are listed in [Table 10.3](#).

10.2.2.1 Drop Size and Size Distribution

The quality of dry powder is the single most important factor that is considerably affected by the operating conditions of the process. Powder character and quality are usually determined by further processing or by consumer requirements. To meet the required bulk density of the dry powder, it is necessary to know how the particle size and size distribution are affected by various parameters. General information about the selection of spray dryer design to meet powder specifications can be found in Refs. [11,23,24], and, for the particular case of food drying, in Ref. [17].

TABLE 10.1
Operating Parameters for Some Spray-Dried Materials

Material	Moisture		Atomizing Device	Air Temperature		
	Inlet (%)	Outlet (%)		Liquid–Air Layout	Inlet (°C)	Outlet (°C)
Skim milk ($D_{3.2} \approx 60 \mu\text{m}$)	48–55	4	Wheel Pressure nozzle (170–200 bar)	95–100		
Whey	50	4	Wheel	Cocurrent	150–180	70–80
Milk	50–60	2.5	Wheel Pressure nozzle (100–140 bar)	Cocurrent	170–200	90–100
Whole eggs	74–76	2–4	Wheel Pressure nozzle	Cocurrent	140–200	50–80
Coffee (instant)	75–85	3–3.5	Pressure nozzle	Cocurrent	270	110
($D_{3.2} \approx 300 \mu\text{m}$)	75–80	3–3.5	Pressure nozzle	Cocurrent	270	110
Tea (instant)	60	≈ 2	Pressure nozzle (27 bar)	Cocurrent	190–250	90–100
PVC emulsions:						
90% particles: $>80 \mu\text{m}$			Pressure nozzle			
5% particles: $>60 \mu\text{m}$	40–70	0.01–0.1	Rotary cup	Cocurrent	165–300	100
Melamine-urethane	30–50	≈ 0	Wheel 140–160 m/s	Cocurrent	200–275	65–75
Detergents:						
Particles: 95–100% $> 60 \text{ m}$ 2–3% $> 150 \text{ m}$	35–50	8–13	Pressure nozzle (30–60 bar)	Countercurrent	350–400	90–110
TiO_2	Up to 60	0.5	Wheel Pressure nozzle	Cocurrent	600	120
				Mixed flow		
Kaolin	35–40	1	Wheel	Cocurrent	600	120
Ammonium phosphate	60	3–5	Pressure nozzle	Cocurrent	400	110–195
Superphosphate	500–600	< 110				
Cream	52–60	4	Wheel	Cocurrent		
Processed cheese	60	3–4	Wheel	Cocurrent		
Whole eggs	74–76	2–4	Wheel			

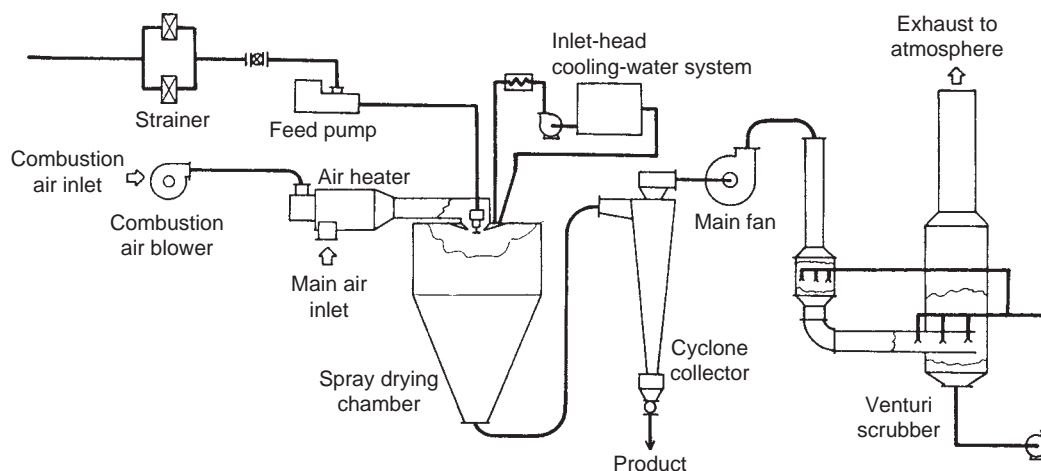


FIGURE 10.1 Spray-drying process and plant.

Particle size and distribution are related to the size of the droplets and their size distribution. Hence, successful prediction of droplet size enables one to control the powder properties as desired.

The mean size of droplet represents a single value that characterizes the whole spray distribution. This value, together with the size distribution, defines the spray characteristics. Many papers have appeared on the subject of drop size prediction. The so-called Sauter mean diameter seems to be the most suitable mean value to characterize the droplet cloud together with the size distribution. This is defined as the ratio of the total droplet volume to the total droplet surface [23]; that is,

$$D_{3,2} = \frac{\sum_i D_i^3 f_i}{\sum_i D_i^2 f_i} \quad (10.1)$$

where f_i is the number frequency of droplet of size D_i . The Sauter mean diameter corresponds to the particle diameter with the same volume-to-surface ratio as the entire spray or powder sample. Sometimes the median diameter D_M is also used in spray drying calculations.

It is that diameter above or below which lies 50% of the number or volume of droplets. It is especially useful when an excessive amount of very large or very small particles are present.

Size distribution can be represented by a frequency or cumulative distributive curve. If occurrence is given by number, a number distribution results. It may be obtained by microscopic analysis. If occurrence is given by area–volume–weights corresponding to a given diameter, then an area–volume–weight distribution results.

The lognormal and Rosin–Rammler distributions are the most common distributions used in the spray drying process calculations. The lognormal distribution has two parameters, the geometric mean size D_{GM} and the geometric standard deviation S_G . The mathematical form is as follows:

$$\frac{dN}{dD} = \frac{1}{DS_G\sqrt{2\pi}} \exp - \left[\frac{(\log D - \log D_{GM})^2}{2S_G^2} \right] \quad (10.2)$$

where N is the number of droplets counted. The graphic form on probability paper is shown in [Figure 10.2](#).

TABLE 10.2
Spherical Droplet Surface versus Droplet Size

Total Volume (m ³)	Diameter of Droplets	No. Droplets	Surface per Droplet	Total Surface of Droplets (m ²)
1	1.234 m	1	3.14 m ²	3.14
1	1 cm	1.986 × 10 ⁶	3.14 cm ²	623.6
1	1 mm	1.986 × 10 ⁹	3.14 mm ²	6,236
1	100 μm	1.986 × 10 ¹²	31,400 μm ²	62,360
1	1 μm	1.986 × 10 ¹³	3.14 μm ²	6,236,000

TABLE 10.3
Range of Droplet and Particle Sizes Obtained
in Spray Dryers (μm)

Rotating wheels	1–600
Pressure nozzles	10–800
Pneumatic nozzles	5–300
Sonic nozzles	5–1000
Milk	30–250
Coffee	80–400
Pigments	10–200
Ceramics	30–200
Pharmaceutics	5–50
Chemicals	10–1000

The lognormal distribution is useful to represent sprays from wheel atomizers [10].

The lognormal distribution also enables us to determine the quantities for evaluation of the dispersion factor D_q , which characterizes the homogeneity of the spray [10,23].

$$D_4 = \frac{D_{95} - D_5}{D_{3,2}} \quad (10.3)$$

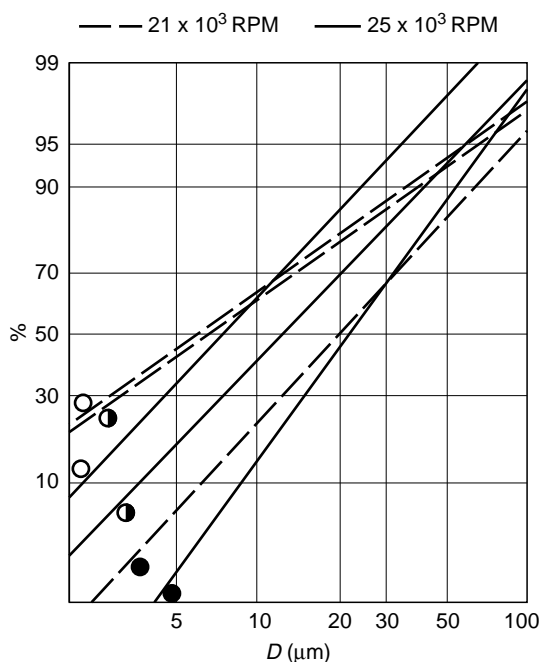


FIGURE 10.2 Lognormal distribution on probability paper, an example. (From Filková, I., Nozzle atomization in spray drying, *Advances in Drying*, Vol. 3, Ed. A.S. Mujumdar, Hemisphere/Springer-Verlag, New York, 1984, pp. 181–216.)

In this equation, D_{95} and D_5 are the droplet diameters at the 95% and 5% probability, respectively.

The Rosin–Rammler distribution is used to represent sprays from nozzles. It is empirical and relates the volume percentage oversize \bar{V}_D to droplet diameter D . The mathematical form is as follows [23]:

$$\bar{V}_D = 100^{-[(D/\bar{D}_R)^{D_4}] \quad (10.4)$$

where D_q is the dispersion factor and \bar{D}_R is the so-called Rosin–Rammler mean diameter. It is the droplet diameter above which lies 36.8% of the entire spray volume. Thus, it follows that the value on the ordinate corresponding to \bar{D}_R equals 0.43. The graphic form of Equation 10.4 on log–log paper is shown in Figure 10.3.

The appropriate correlations for drop size prediction will be presented along with the discussion of each individual type of atomizer in the following sections.

10.2.2.2 Wheel Atomizers

A typical wheel atomizer is shown schematically in Figure 10.4. Liquid is fed into the center of a rotating wheel, moves to the edge of the wheel under the centrifugal force, and is disintegrated at the wheel edge into droplets. The spray angle is about 180° and forms a broad cloud. Because of the horizontal trajectory these atomizers require large-diameter chambers. The most common design of the wheel atomizer has radial vanes.

The linear peripheral speed ranges from 100 to 200 m/s. For the usual wheel diameter, angular speeds between 10,000 and 30,000 rpm are necessary [15] (see Figure 10.5).

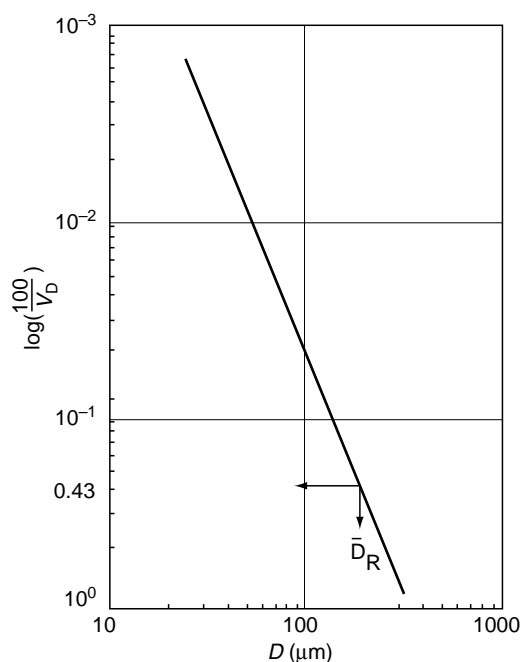


FIGURE 10.3 Rosin–Rammler distribution on log–log paper.

The number and shape of the vanes differ according to the product quality and capacity requirements. The usual shape of vanes is circular, oval, or rectangular, as shown in Figure 10.6. The influence of vane shape on droplet size was studied in Ref. [10]. For high-capacity applications, where the same degree of atomization is required at higher feed rates, the number and height of the vanes are increased to maintain the same liquid-film thickness on each vane. High-capacity wheels often have two tiers of vanes [22] (see Figure 10.7). The largest wheel atomizers allow feed rates up to 200 ton/h. The curved vane wheels are sometimes used instead of the standard straight radial wheels, mainly in the milk industry. The curved vane wheel produces a powder of high bulk density, up to 15% higher than the standard wheels, because of reduced air-pumping effect. The other way to achieve high bulk density of the product is to use the so-called bushing wheels, which are widely used in the chemical industry. The bushings are made from very hard material, such as silicon carbide; thus the wheel can atomize even very abrasive materials [22].

Generally, the wheel atomizer produces a spray of high homogeneity within a wide range of mean droplet size. The size distribution of droplets can be controlled by changing the wheel speed. Feed rate variation produces much less effect. Wheel atomizers are very flexible and can handle a wide assortment of liquids with different physical properties. Factors

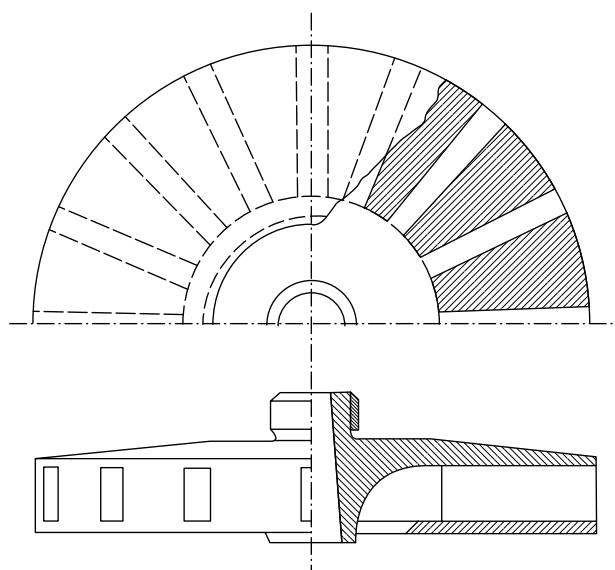


FIGURE 10.4 Wheel atomizer with straight radial vanes.

influencing the wheel atomizer performance are specified in Ref. [25] for instance.

The so-called vaneless disk atomizers should also be mentioned, although their use is limited to specialized applications in which coarse particles are required at high production rates. A typical cup-type disk atomizer is shown in Figure 10.8. Liquid is fed inside the cup and is pressed by centrifugal force against the cup wall, flowing down toward the cup edge, where it is disintegrated.

The wheels are conventionally driven by electric motors, either directly or by belt drives through a worm and worm wheel gearing system (low-capacity wheels) or by helical or epicyclic gearing systems (high-capacity wheels).

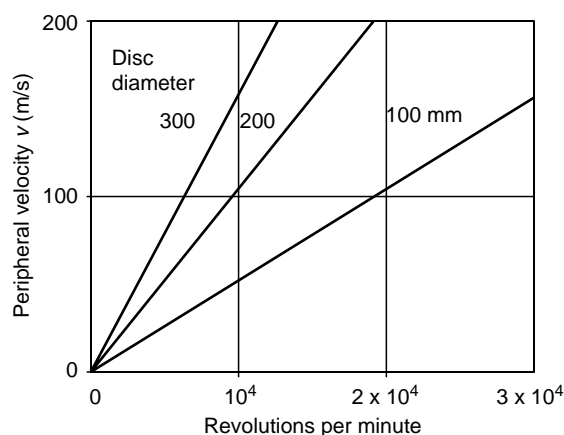


FIGURE 10.5 Peripheral velocity of a wheel atomizer as a function of the wheel diameter and revolutions.

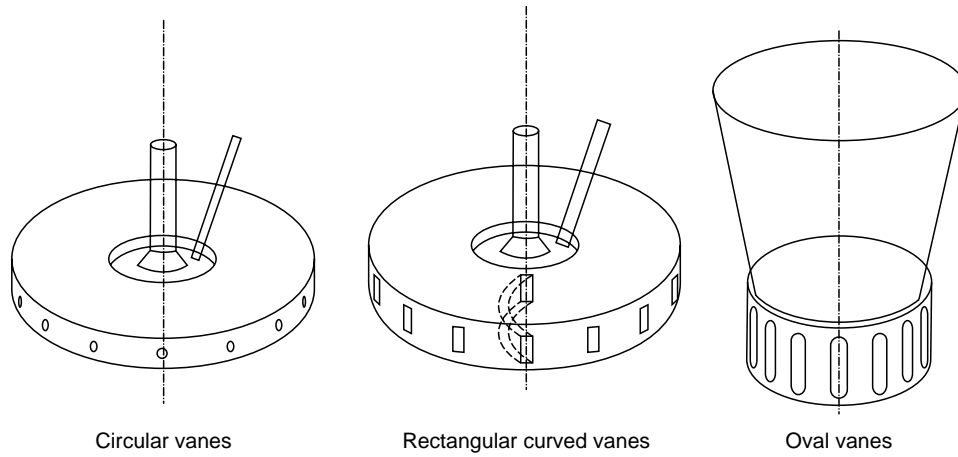


FIGURE 10.6 Various designs of wheel atomizers.

The power necessary to accelerate the liquid to the velocity at the periphery of the wheel is given by

$$E = \frac{\dot{M}}{2} \frac{v^2}{\varepsilon} \quad (10.5)$$

For example, for a mass flow rate $\dot{M} = 1$ kg/s and a peripheral velocity $v = 150$ m/s at an efficiency $\varepsilon = 0.65$, the required electrical power $E = 17.3$ kW.

The drop size (Sauter mean diameter, in m) can be predicted using any one of the following correlations [23]:

$$\frac{D_{3,2}}{r} = 0.4 \left(\frac{\dot{M}}{N_v b p N r^2} \right)^{0.6} \left(\frac{N_v \mu b}{\dot{M}} \right)^{0.2} \times \left(\frac{\sigma b^3 \rho N_v^3}{\dot{M}^2} \right)^{0.1} \quad (10.6)$$

$$D_{3,2} = 0.241 \left(\frac{1}{N} \right)^{0.6} \left(\frac{1}{\rho} \right)^{0.3} \left(\frac{\mu \dot{M}}{2 r p} \right) \left(\frac{\sigma}{N_v b} \right)^{0.1} \quad (10.7)$$

$$D_{3,2} = 1.62 \times 10^{-3} N^{(-0.53)} \dot{M}^{0.21} (2r)^{-0.39} \quad (10.8)$$

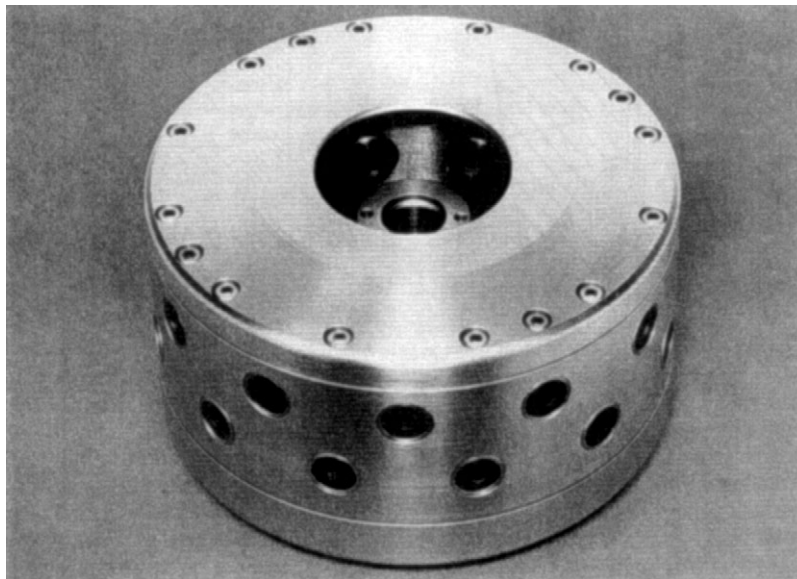


FIGURE 10.7 High-capacity wheel atomizer. (Courtesy of Niro Atomizer.)

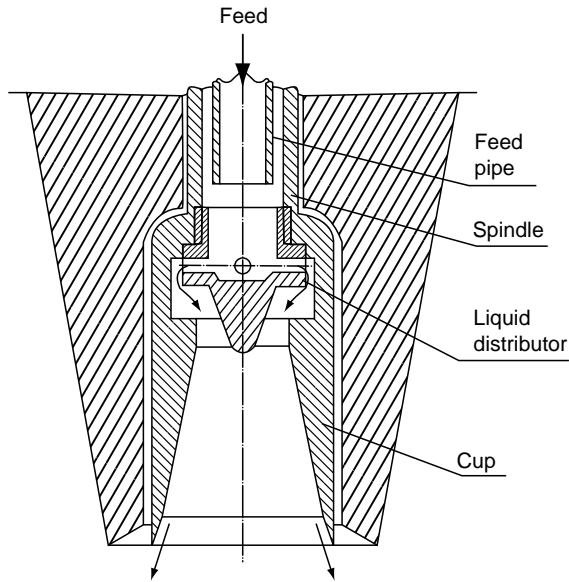


FIGURE 10.8 Cup disk atomizer. (From Masters, K., *Spray Drying*, Leonard Hill Books, London, 1979.)

where \dot{M} (kg/s) is the mass feed rate, N (rpm) is rotational speed, r (m) is wheel diameter, b (m) is vane height, N_v is number of vanes, ρ (kg/m³) is fluid density, σ (N/m) is fluid surface tension, and μ (Pa·s) is fluid viscosity. Equation 10.6 through Equation 10.8 display deviation of $\pm 30\%$ from experimental data reported in the literature.

Most feeds are high-consistency slurries that are generally non-Newtonian, mainly pseudoplastic in nature. The dynamic viscosity of these liquids should be substituted by the apparent viscosity, which is defined as

$$\mu_A = K \left[\left(\frac{r\omega^2\rho}{K} \right)^2 \frac{V_v}{b} \frac{2n+1}{n} \right]^{\frac{n-1}{2n+1}} \quad (10.9)$$

where ω is angular velocity (rad/s), V_v (m³/s) is volumetric feed rate per vane, K (Pa·s) is fluid consistency, and n is the flow index (dimensionless) for power law fluids [9]. The influence of non-Newtonian character on drop formation is reviewed in Ref. [9].

The advantage of Equation 10.8 is that it does not contain any physical parameters on the liquid. It can be used for a rough estimation of the drop size. For the application of more accurate correlations [4,5], a knowledge of the feed properties is needed.

10.2.2.3 Pressure Nozzles

A pressure nozzle, sometimes called a single-fluid nozzle, creates spray as a consequence of pressure to velocity energy conversion as the liquid passes

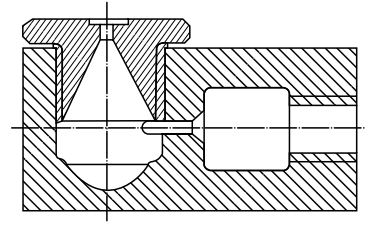


FIGURE 10.9 Pressure nozzle.

through the nozzle under pressure within the usual range of 5–7 MPa (see Figure 10.9). The liquid enters the nozzle core tangentially and leaves the orifice in the form of a hollow cone with an angle that varies from 40 to 140°. The orifice diameter is usually small, from 0.4 to 4 mm, and the usual capacity of one nozzle does not exceed 100 l/h. When larger feed rate is to be processed, several nozzles are used in the drying chamber (see Figure 10.10). Owing to their smaller spray angles the drying chamber can be narrower and taller. With this type of nozzle it is generally possible to produce the droplets within a narrow range of diameters, and the dried particles are usually hollow spheres. Pressure nozzles are not suitable for highly concentrated suspensions and abrasive materials because of their tendency to clog and erode the nozzle orifice.

Energy consumption of a pressure nozzle is very low in comparison with that of the wheel atomizer as well as the pneumatic nozzle. The drop size calculation for a pressure nozzle may be done using the correlation [23]:

$$D_{3,2} = 286[(2.54 \times 10^{-2})D + 0.17] \exp \left[\frac{39}{v_{AX}} - (3.13 \times 10^{-3})v_1 \right] \quad (10.10)$$

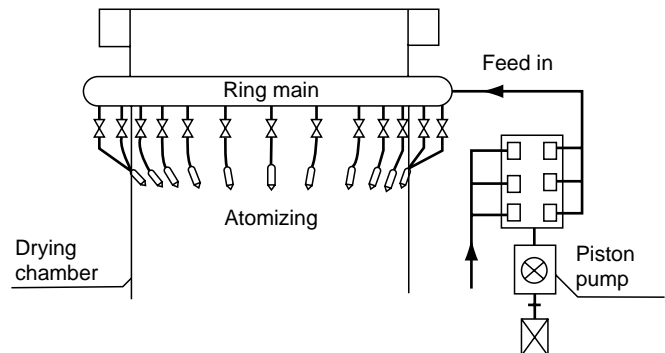


FIGURE 10.10 Multinozzle system in the drying chamber. (From Kessler, H.G., *Food Engineering and Dairy Technology*, Verlag A. Kessler, Freising, Germany, 1981.)

where the axial velocity v_{AX} (m/s) and the inlet velocity v_1 (m/s) are determined as follows:

$$v_{AX} = \frac{D_1^2}{2Db} v_1 \quad (10.11)$$

$$v_1 = \frac{\dot{V}_1}{A_1} \quad (10.12)$$

In this equation, D is orifice diameter (m), D_1 is inlet channel diameter (m), A_1 is inlet channel area (m²), \dot{V}_1 is volumetric flow rate (m³/s), and b is thickness of fluid film in the orifice. The resulting Sauter mean diameter is in μm .

For a rough prediction, the following empirical equation may be used [6]:

$$D_A = \frac{9575}{\Delta P^{1/3}} \quad (10.13)$$

where D_A is the average drop diameter (μm) and ΔP is the pressure drop across the nozzle (Pa).

10.2.2.4 Pneumatic Nozzles

Pneumatic nozzles are also known as two-fluid nozzles as they use compressed air or steam to atomize the fluid. Figure 10.11 shows the most usual type. In this case the feed is mixed with the air outside the body of the nozzle. Less frequently, the mixing occurs inside the nozzle. The spray angle ranges from 20 to 60°C and depends on the nozzle design. Approximately, 0.5 m³ of compressed air is needed to atomize 1 kg of fluid. The capacity of a single nozzle usually does not exceed 1000 kg/h of feed. Sprays of less-viscous feeds are characterized by low mean droplet sizes and a high degree of homogeneity. With highly viscous feeds, larger mean droplet sizes are produced

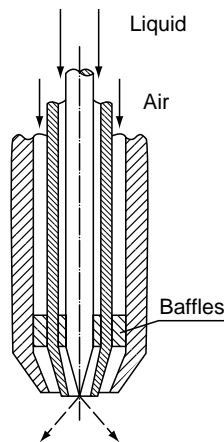


FIGURE 10.11 Pneumatic nozzle with external mixing.

but homogeneity is not as high. Pneumatic nozzles are very flexible and produce small or large droplets according to the air–liquid ratio. The high cost of compressed air (pressure range, 0.15–0.8 MPa) becomes important to the economics of these nozzles, which have the highest energy consumption of all three types of atomizers.

The drop size can be predicted by means of the following correlation, which yields the Sauter mean diameter in μm [23]:

$$D_{3,2} = \frac{535 \times 10^3 \sqrt{\sigma}}{v_{REL} \sqrt{\rho}} + 597 \left(\frac{\mu}{\sqrt{\sigma \rho}} \right)^{0.45} \times \left(\frac{1000 \dot{V}_{FL}}{\dot{V}_{AIR}} \right) \quad (10.14)$$

where σ , ρ , and μ are the fluid surface tension (N/m), density (kg/m³), and viscosity (Pa s), respectively, and \dot{V}_{FL} and \dot{V}_{AIR} are volumetric flow rates of fluid and air (m³/s), respectively. Instead of relative velocity v_{REL} (m/s), the outlet velocity of air may also be substituted.

10.2.2.5 Novel Types of Atomizers

A number of liquids that cannot be atomized successfully by wheels or nozzles method have generated interest in using other methods of atomization that may be more suitable for such liquids. These are, for example, highly viscous and long molecular chain structured materials and some non-Newtonian liquids, which form only filaments instead of individual droplets from the ordinary atomizers. Attention has been paid to the use of sonic energy. The most recent development in the field of pneumatic nozzles is the sonic atomizer [31]. The breakup mechanism is entirely different from that for conventional nozzles. The disintegration of a liquid occurs in the field of high-frequency sound created by a sonic resonance cup placed in front of the nozzle. However, this development has not reached the stage at which sonic nozzles can be industrially competitive with other kinds of atomizers. They have some promising aspects, such as 15% savings in energy [35] and applicability for abrasive and corrosive materials. The four main types of sonic atomizers are the Hartman mono-whistle nozzle, steam jet nozzle, vortex whistle nozzle, and mechanical vibratory nozzle [23].

10.2.2.6 Selection of Atomizers

The selection of the atomizer usually means the selection between wheel atomizer and pressure nozzle, as the use of the pneumatic nozzle is very limited. The selection may be based on various considerations, such as

availability, flexibility, energy consumption, or particle size distribution of the final dry product. The last is the most common case. The sizes of droplets produced by various atomizers are shown in Table 10.3. The advantages and disadvantages of both wheel and pressure nozzle atomizers are summarized below.

Wheel nozzles

Advantages

- Can handle high feed rates in a single wheel
- Suitable even for abrasive materials
- Negligible blockage or clogging tendencies
- Simple droplet size control by changing wheel revolutions

Disadvantages

- Higher energy consumption than pressure nozzles
- Higher capital cost than pressure nozzles
- Broad spray requires large chamber

Pressure nozzles

Advantages

- Simple and compact construction, no moving parts
- Low cost
- Low energy consumption
- Required spray characteristics can be produced by alternation of the whirl chamber design

Disadvantages

- Control and regulation of spray pattern and nozzle capacity during operation not possible
- Swirl nozzles not suitable for suspensions because of phase separation
- Tendency to clog
- Strong corrosion and erosion effects cause enlargement of the orifice, which changes the spray characteristics

The energy consumption of three main types of atomizers is summarized in Table 10.4.

In many cases rotary and nozzle atomizers can be used with equal success and the choice depends entirely on the manufacturer's tradition. However, there are differences in the dry product characteristics, bulk density, and shape between the wheel and nozzle atomizers (see Figure 10.12) [15]. In cases in which both wheel and nozzle atomizers produce similar spray patterns, the wheel atomizer is usually preferred because of its greater flexibility.

10.2.3 CHAMBER DESIGN

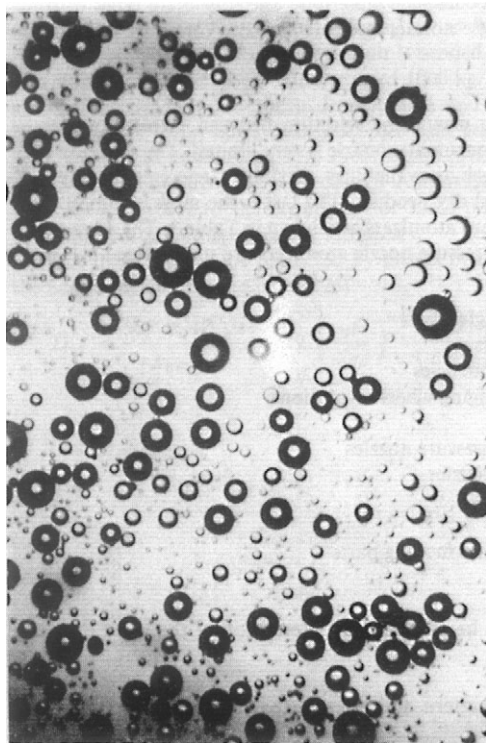
Generally the chamber design depends on the atomizer used and on the air–fluid contact system selected. The selection of atomizer and air–fluid layout is determined by the required characteristics of the dry product and production rate. The product specifications are almost always determined from small-scale tests in an experimental spray dryer. Sometimes they are available from the literature. If heat-sensitive material is involved, attention must be paid to the temperature profile of the drying air along the drying chamber. Another area that requires attention is the droplet trajectory, mainly the trajectory of the largest drops as the size of chamber must be such that the largest drop in the spray is dry before it reaches the chamber wall. This requirement prevents the formation of partially dried material buildup on the chamber walls. The height H of the drying chamber as a function of droplet diameter d and temperature differences ΔT between the drying air and the particle is shown in Figure 10.13.

10.2.3.1 Chamber Shape

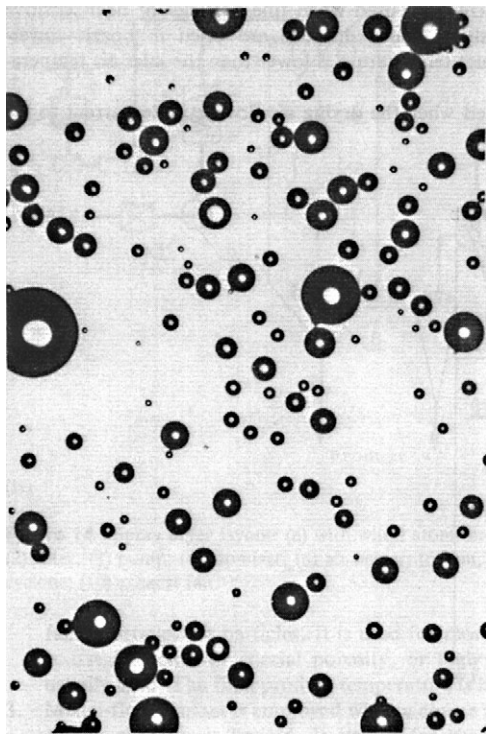
As noted earlier, the chamber shape depends on the type of atomizer employed because the spray angle determines the trajectory of the droplets and therefore the diameter and height of the drying chamber. Typical spray dryer layouts with wheel and nozzle atomization are shown in Figure 10.14. Correlations are available for calculating the drying chamber sizes without pilot tests. However, these equations require simplifying assumptions that make their application

TABLE 10.4
Energy Consumption of Three Main Types of Atomizers

Atomizer Type	Energy Consumption for Atomization of			
	250 kg/h	500 kg/h	1000 kg/h	2000 kg/h
Pressure nozzle, pressure 3–5 MPa	0.4	1.6	2.5	4.0
Pneumatic nozzle, air pressure 0.3 MPa, air mass rate 0.5–0.6 m ³ /kg	10.0	20.0	40.0	80.0
Rotary wheel	8.0	15.0	25.0	30.0



(a)



(b)

FIGURE 10.12 Spray-dried powder: (a) produced by wheel atomizer; (b) produced by pressure nozzle under comparable operating conditions.

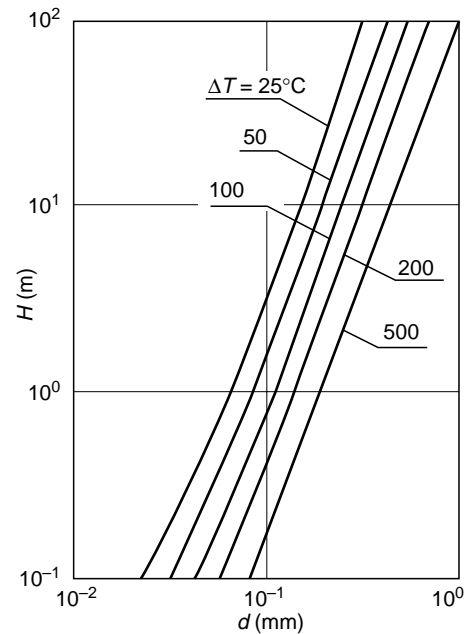


FIGURE 10.13 Height H of the drying chamber versus diameter d ; ΔT = temperature difference between the drying air and the particle.

unreliable. The best method so far has been the scale-up from pilot to commercial size.

10.2.3.2 Air-Droplet Contact Systems

There are three basic types of air-droplet contact systems employed in spray drying processes:

1. Cocurrent contact occurs when the droplets fall down the chamber with the air flowing in the same direction. It is the most common system with both wheel and nozzle atomization. Wheel atomizers are used when fine particles of heat-sensitive material are required; heat-sensitive coarse droplets are dried in nozzle tower-chamber designs. The final product-temperature is lower than the inlet air temperature.
2. Countercurrent contact is achieved when the drying air flows countercurrent to the falling droplets or particles. It is used for more heat-sensitive materials that require coarse particles, or special porosity, or high bulk density. Nozzle atomization is usually used. The final product-temperature is higher than that of the exit air.
3. Mixed-flow contact is employed when a coarse product is required and the size of the drying chamber is limited. It has so far been the most

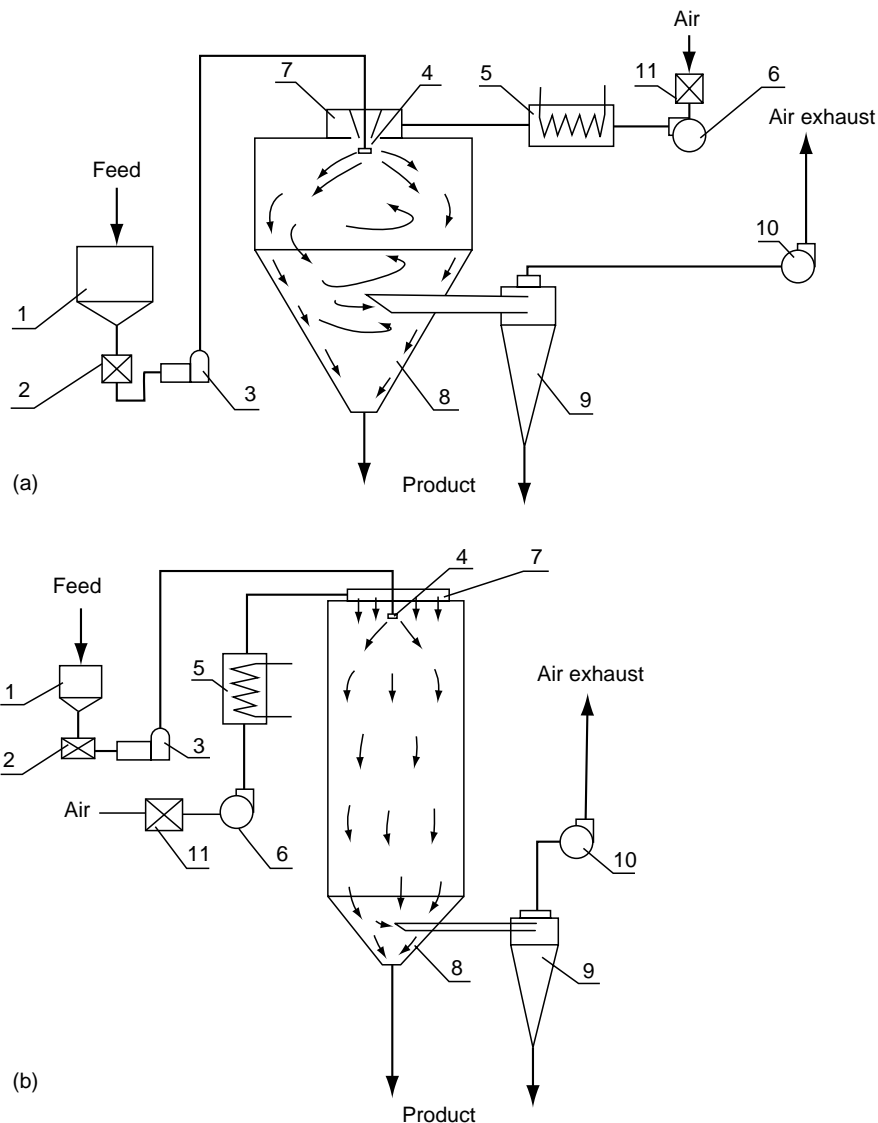


FIGURE 10.14 Spray dryer layout: (a) with wheel atomizer; (b) with nozzle atomizer; (1) feed tank; (2) filter; (3) pump; (4) atomizer; (5) air heater; (6) fan; (7) air disperser; (8) drying chamber; (9) cyclone; (10) exhaust fan.

economical system for a material that can withstand exposure to high temperature in dry form.

The drying chamber layouts for all three systems are shown schematically in Figure 10.15; a typical set of temperature data can be found in Figure 10.16 [20]. Temperature and vapor pressure profiles along the drying chamber are shown in Figure 10.17 for cocurrent and countercurrent contact layouts [15].

The direction of airflow and the uniformity of the air velocity over the whole cross section of the chamber are very important in determining the final product quality. Design of the hot air distribution must prevent

the local overheating due to reverse flow of wet particles into the hot air area, ensure that the particles are dry before they reach the wall of the drying chamber, and prevent insufficient drying in some areas of the chamber. Some arrangements of hot air distribution are shown in Figure 10.18.

The areas sensitive to wall impingement of wet product are shown in Figure 10.19. There are many ways to control wall impingement, depending on the atomizer applied, the product properties, and the air distribution design [23]. The temperature of the chamber wall that comes in contact with the particles must be lower than the melting point of the product to prevent baking on.

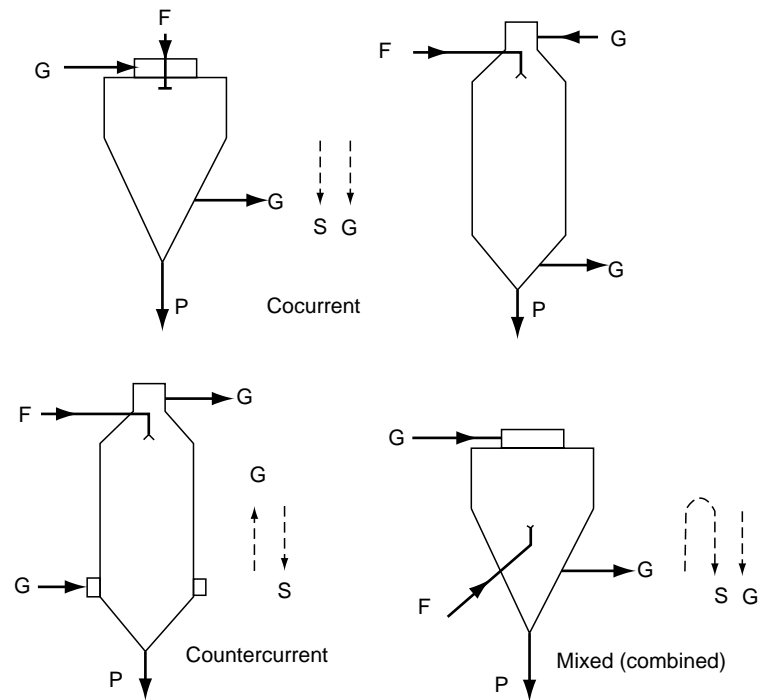


FIGURE 10.15 Drying chamber layouts: F, feed; G, gas; P, product; S, spray. (From Masters, K., *Chemical Age of India*, 30, 1979.)

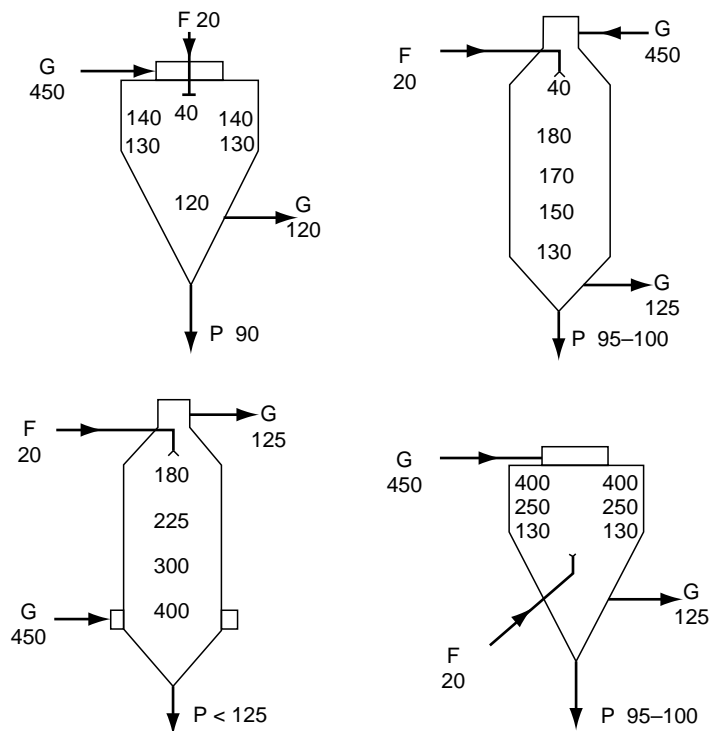


FIGURE 10.16 Temperature data in the drying chamber: F, feed; G, gas; P, product. (From Masters, K., *Chemical Age of India*, 30, 1979.)

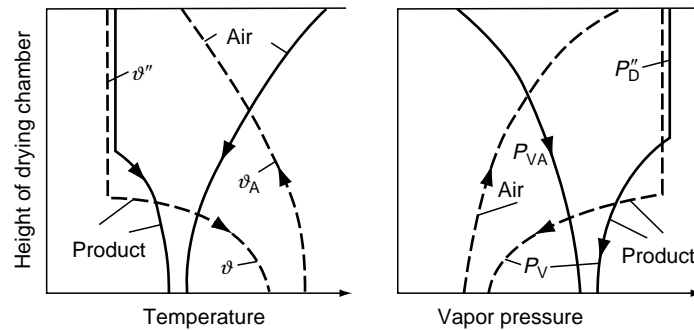


FIGURE 10.17 Temperature and vapor pressure profiles along the drying chamber; dashed lines refer to cocurrent operation, solid lines to countercurrent operation. (From Kessler, H.G., *Food Engineering and Dairy Technology*, Verlag A. Kessler, Freising, Germany, 1981.)

10.2.3.3 Powder and Air Discharge Systems

The dry powder is collected at the bottom of the drying chamber and discharged. The powder–air separation can be done outside the chamber or the main part of the product can be separated inside the chamber, whereas outside, in a separation device, only the fine particles are collected. Examples of different discharge systems are sketched in Figure 10.20.

In case (1) of Figure 10.20, the separation occurs outside. In cases (2), (3), and (4) the product is accumulated at the bottom of the chamber and discharged by means of a valve. Vibrators and rotating scrapers help convey the powder when the cone angle is too large or the chamber bottom is flat. When necessary, the air sweeper can be used to

discharge the dry product along the chamber wall, as in case (4).

10.2.4 ANCILLARY EQUIPMENT

The ancillary equipment used depends generally on the spray drying process layout. Nevertheless, some basic pieces of equipment must be used in any system (see Figure 10.1). They are discussed below:

- Air Heaters
- Steam, indirect
- Fuel oil, direct or indirect
- Gas, direct or indirect
- Electric
- Thermal fluids, indirect

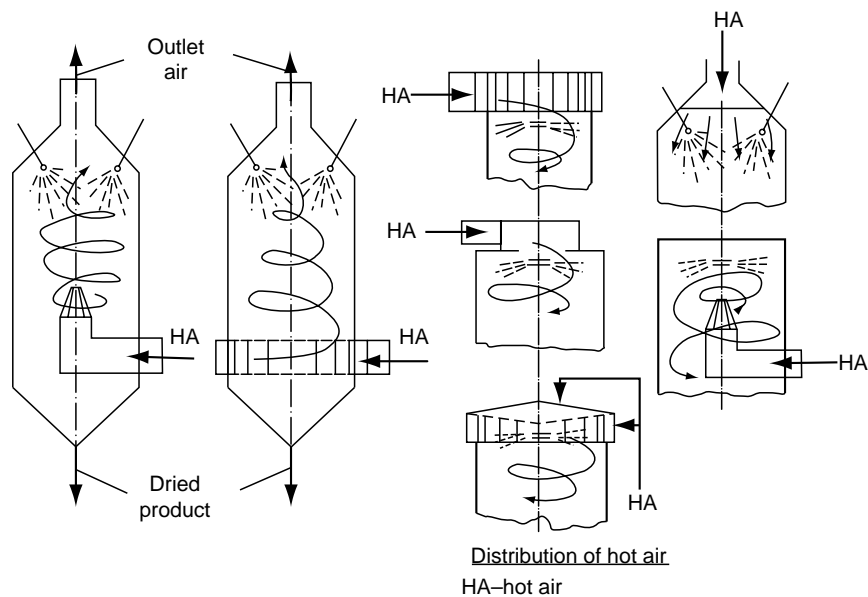


FIGURE 10.18 Hot air distribution layouts. (From Kessler, H.G., *Food Engineering and Dairy Technology*, Verlag A. Kessler, Freising, Germany, 1981.)

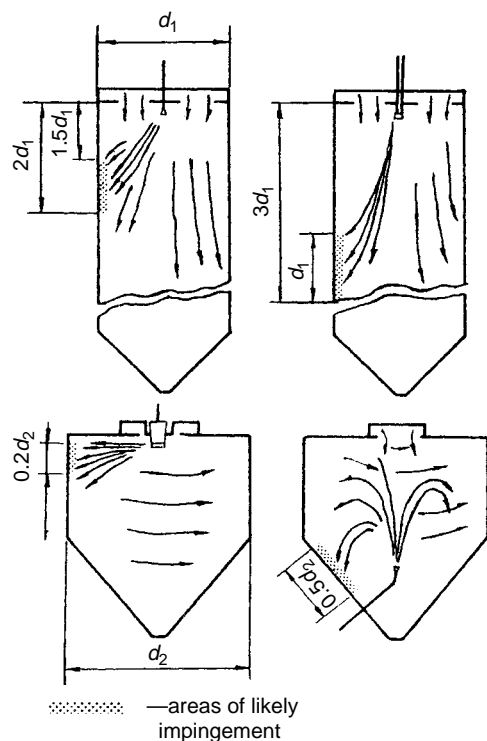


FIGURE 10.19 Possible areas of impingement on the drying chamber wall. (From Masters, K., *Spray Drying*, Leonard Hill Books, London, 1979.)

Direct heaters may be used if the material can come into contact with products of combustion. Otherwise, indirect air heaters must be used. The type of heater used depends on the required temperature of the drying air and on the availability of the heat source; the most common air heater in the food industry is the steam-heated type. Dry saturated steam is usually used within the temperature range 150–250°C. The

air temperature is about 10°C lower than the steam temperature. Steam air heaters use extended fin tubes and are relatively inexpensive. The steam rate M_s (kg/s) necessary to heat the drying air is given by [23]

$$\dot{M}_s = \frac{L_A C_{PA}(T_{Ao} - T_{Ai})}{H_s - H_c} \eta \quad (10.15)$$

where L_A is airflow rate (kg/s); C_{PA} is air specific heat (J/kg·K); T_{Ao} and T_{Ai} are the air outlet and inlet temperatures (K), respectively; H_s and H_c are the enthalpies of steam and condensate (J/kg), respectively; and η is heater efficiency (about 0.95).

Indirect fuel oil heaters and gas heaters have similar features. They have separate flow passages for hot gas and drying air. The maximum air temperature at the outlet of a heater is about 400°C.

Direct air heaters are less complicated, and drying air temperatures up to 800°C can be obtained. Gas-fired direct heaters have been used more frequently in recent years because of the increasing availability of natural gas in many parts of the world. The fuel (or gas) combustion rate M_G (kg/s) is given by

$$\dot{M}_G = \frac{L_A C_{PA}(T_{Ao} - T_{Ai})}{Q_{cv}} \eta \quad (10.16)$$

where Q_{cv} is the calorific value of the fuel or gas (J/kg).

Electric air heaters are used mainly with laboratory and pilot-scale dryers because of high electricity costs. Air temperatures up to 400°C can be achieved using electric heaters.

A relatively new type of air heater is one that uses a thermal fluid, which is a special oil that transfers heat from a boiler to the air heater. The boiler may be gas or fuel oil fired. This type of heater may be used when steam is not available and the temperature of

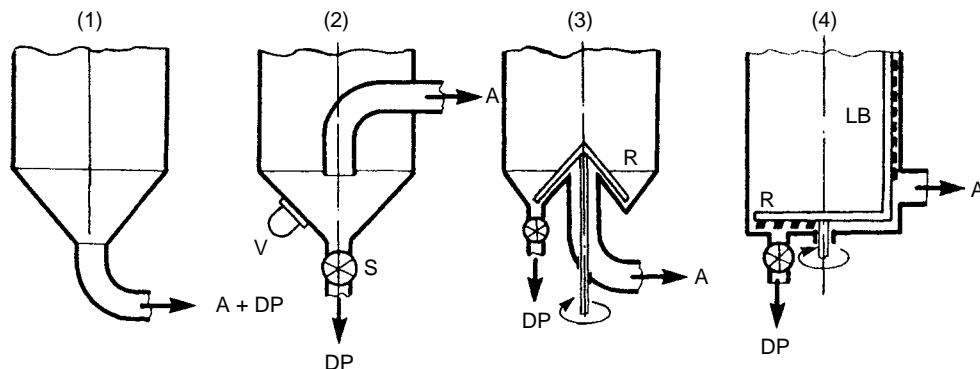


FIGURE 10.20 Examples of discharge systems: A, outlet air; DP, dry product; RS, rotating scraper; V, vibrator; AS, air sweeper.

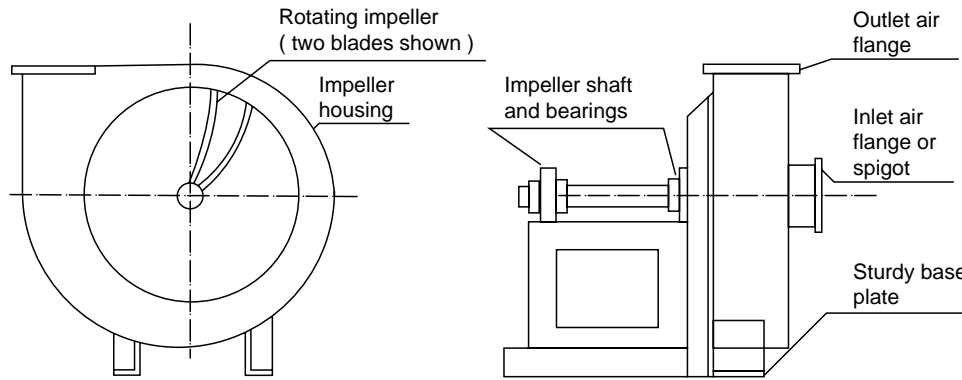


FIGURE 10.21 Typical centrifugal fan.

inlet air up to 400°C is required and when combustion gases cannot be used directly.

10.2.4.1 Fans

In a spray drying process, high flow rates of drying air are generally obtained by the use of centrifugal fans. Usually a two-fan system is used, the main fan situated after the powder recovery equipment and the supply fan located in the inlet duct to the drying chamber. Two fans enable better control of the pressure in the chamber. With a single fan after the cyclone, the whole drying system operates under a high negative pressure. The operating pressure in a drying chamber determines the amount of powder in the exhaust air and hence the capacity of the cyclones and their collection of efficiency. In special cases, more fans may be used in a drying process, for example, a centrifugal fan for the powder pneumatic transport or small fans for blowing cool air to potential hot spots in the drying chamber and atomizer. A typical centrifugal fan is illustrated in Figure 10.21.

The pressure developed by a fan depends upon the blade design. The most common type of fan in spray drying has backward-curved blades, as shown in Figure 10.21. Such blades are also used for the supply fan. If the powder–air ratio is high, the backward-curved blades may cause problems with deposit formation on the backside of the blades. In such cases, the use of a blade profile intermediate between the backward curving and the radial is recommended. It gives better self-cleaning properties. In Figure 10.22 are shown the typical characteristic curves for radial and backward-curved blades for constant fan speed and air density.

The fan energy consumption E (in W) may be calculated by means of the following relationship between the volumetric airflow rate \dot{V} (m³/s) and the total pressure drop that must be overcome, ΔP (Pa):

$$E = \frac{\Delta P \dot{V}}{\eta} \quad (10.17)$$

where η is the fan efficiency, usually 0.6–0.75.

10.2.4.2 Powder Separators

Dry powder: Cyclones, bag filters, electrostatic precipitators

Wet powder: Wet scrubbers, wet cyclones, irrigated fans

Powder separators must separate dry product from the drying air at the highest possible efficiency and collect the powder. Dry separators are used for the principal dry product separation and collection; wet separators are used for the final air cleaning and hence are situated after dry collectors.

In a dry cyclone (see Figure 10.23), centrifugal force is employed to move the particle toward the wall and separate them from the air core around the

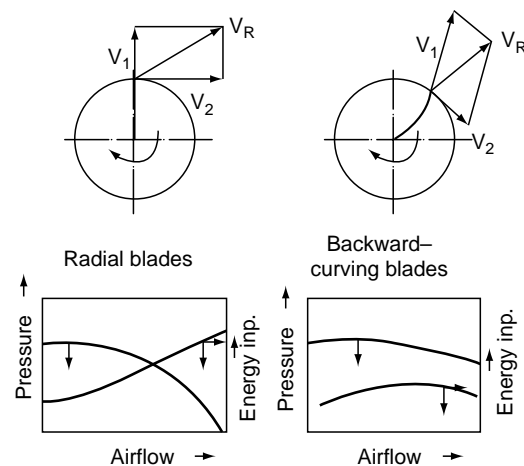


FIGURE 10.22 Typical characteristic curves for radial and backward-curved blades.

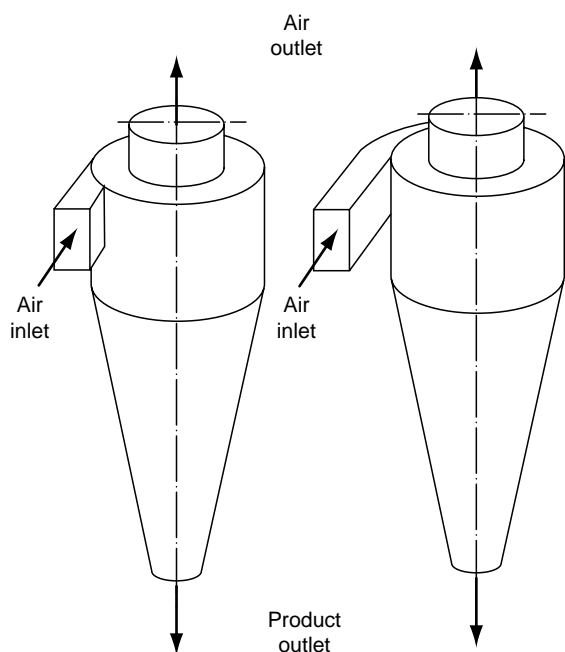


FIGURE 10.23 Dry cyclone designs with different types of air inlet.

axis. Air and particle swirl in a spiral down the cyclone, where the particles collect and leave the cyclone. The clean air flows upward and leaves from the top.

In a spray drying process a simple cyclone system or a multicyclone system may be used. The choice between them depends upon the following factors:

1. Size characteristics to determine the smallest particle size that can be separated in unit

2. Overall cyclone efficiency
3. Pressure drop over the unit

Two characteristics are used to define cyclone performance. They are the critical particle diameter (particle size that is completely removed from the air stream) and the cut size (the particle diameter for which 50% collection efficiency is achieved). A typical example of theoretically and experimentally obtained efficiency curves is shown in Figure 10.24. It is evident from this diagram that particles above 15 μm are removed with high efficiency in the cyclone. The pressure drop across the cyclone unit ranges between 700 and 2000 Pa.

A bag filter is widely used in spray drying processes. The airflow containing dry particles passes through a woven fabric; powder is collected on one side of the fabric and the air leaves on the other. A modern unit consists of several bags installed in a baghouse. A typical collection efficiency curve is shown in Figure 10.25. Very high efficiency is obtained even with 1- μm particles. Bag filters must be carefully maintained to avoid any leakage and regularly cleaned to maintain high operating efficiency. The fabric used in a bag filter is selected in accordance with the characteristics of the dry product and the air temperature. Only rarely are electrostatic precipitators used in spray drying processes because of their high initial cost.

Wet scrubbers are very commonly used following dry collectors. The particles are separated from air by contacting it with a liquid, usually water. The well-known venturi scrubber is preferred in spray drying systems because it offers easy cleaning and maintenance. It can be used for food and pharmaceutical materials that require hygienic handling. A venturi

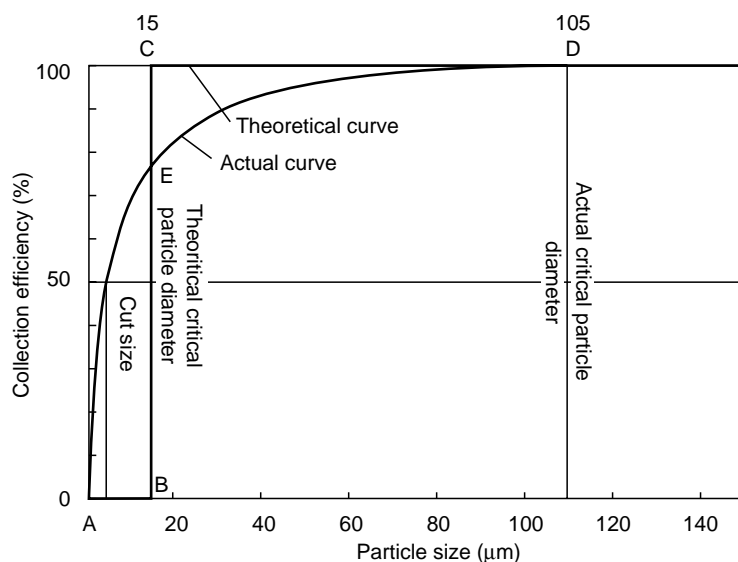


FIGURE 10.24 Cyclone efficiency curves, theoretical and actual.

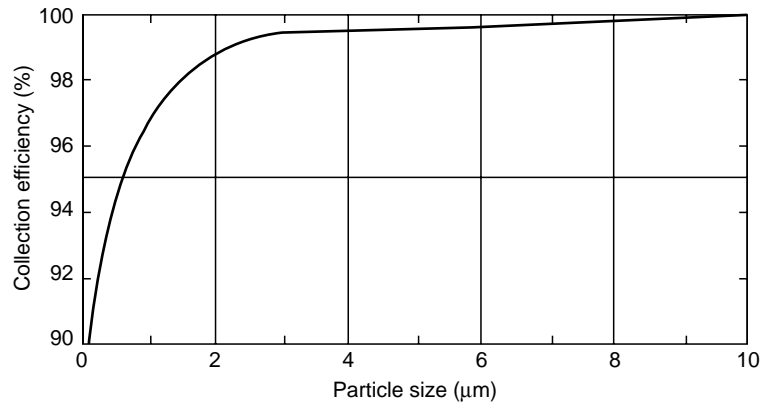


FIGURE 10.25 Bag filter efficiency curve.

scrubber is shown in Figure 10.26 and its typical efficiency curve in Figure 10.27.

The air carrying fine particles flows through a venturi; water is injected at the throat of the scrubber to form a spray. The scrubbing liquid containing the product is separated out and discharged from the scrubber base either to sewage or for recirculation. The pressure drop over the scrubber is usually between 2000 and 5000 Pa.

10.2.5 THERMAL EFFICIENCY

The efficiency of the spray drying operation is defined as the ratio of the heat used in evaporation to the total heat input. Thus

$$\eta = \frac{\dot{M}_{CH}\lambda}{L_A(T_A - T_{WB})C_{PA} + \dot{M}_F(T_F - T_{WB})C_{PF}} \quad (10.18)$$

where \dot{M}_{CH} (kg H₂O/s) represents the chamber evaporation capacity, λ (J/kg) the latent heat of evaporation, L_A (kg/s) the airflow rate, \dot{M}_F (kg/s) the feed flow rate, C_{PA} and C_{PF} (J/kg K) the heat capacity of the air and feed, respectively, T_A and T_F (°C) the temperature of the air and feed, respectively, and T_{WB} (°C) = wet bulb temperature.

If the drying process is supposed to be adiabatic, that is, the heat losses are negligible, Equation 10.18 can be approximated to the relation

$$\eta = \left(\frac{T_{Ai} - T_{Ao}}{T_{Ai} - T_{AMB}} \right) \times 100(\%) \quad (10.19)$$

where T_{Ai} and T_{Ao} are the inlet and outlet air temperatures, respectively, and T_{AMB} is the ambient air temperature. Equation 10.19 is illustrated graphically in Figure 10.41.

As the drying chamber capacity is proportional to the temperature difference of the inlet and outlet air over the chamber, it is desirable to achieve the highest possible value, which are the highest inlet air temperature and the lowest outlet air temperatures. However, there are some limitations to be considered. For many products an increase in the inlet temperature causes serious product damage and a decrease in the outlet air temperature leads to higher moisture content in the dried product.

On the other hand, a decrease in heat input would also cause an increase in the thermal efficiency. The heat consumption is proportional to the evaporation rate, and for a given rate, it depends on the concentration of the dryer feed. Increase in feed solid content from 10 to 25% will result in a 66.6% reduction in heat

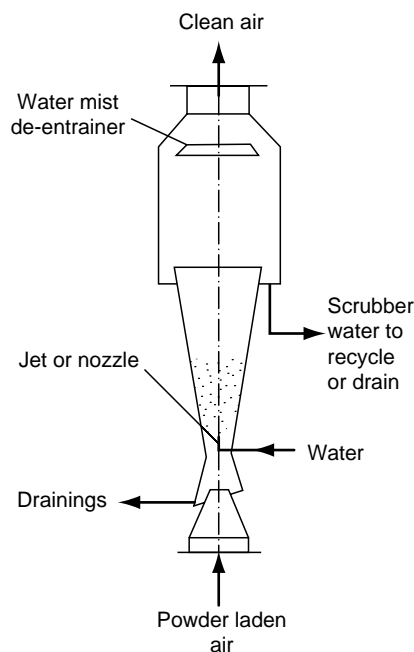


FIGURE 10.26 Typical design of venturi wet scrubber.

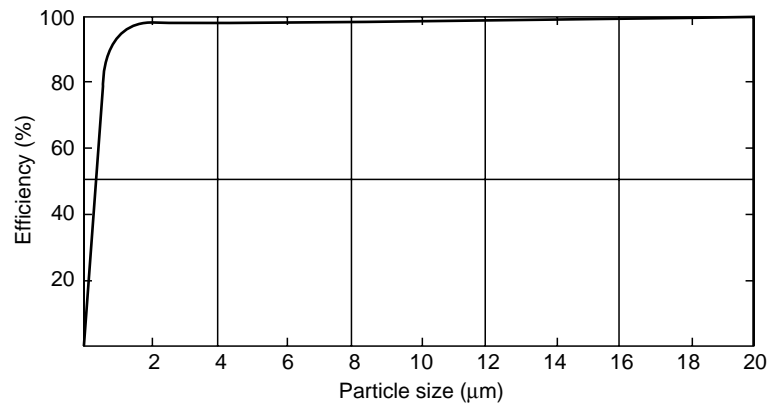


FIGURE 10.27 Venturi scrubber efficiency curve.

consumption for a given rate. A spray dryer has the highest heat consumption for evaporation of 1-kg water in comparison with any dehydration equipment (see Table 10.5).

The feed must be concentrated as much as possible before spray drying. An example of heat consumption in a spray dryer is given in Figure 10.28, in which two spray drying processes are compared, operating at different temperature ranges and employing different heat sources.

10.3 SPRAY DRYING SYSTEMS

10.3.1 PROCESS LAYOUTS AND APPLICATIONS

The characteristics of different process layouts are listed in Table 10.6 [22]. The open-cycle layout, which represents the majority of spray dryer systems, is shown in Figure 10.29. The drying air is taken from the atmosphere and the exhaust air is discharged to the atmosphere. Three types of dry collectors and a wet air cleaner can be used in this layout. A relatively large amount of dry powder (in comparison with other layouts) is lost with the exhaust air. Some applications are shown in Figure 10.30 through Figure 10.35 [33].

With reference to Figure 10.30, the drying medium is a combustion gas that is produced at a temperature of about 550°C in a combustion chamber by burning natural gas. The gas outlet temperature is about 120°C. Dryer capacity is 12 ton/h, its specific evaporation rate is in the range from 8 to 9 kg/(m³ h).

Figure 10.31 shows an example of a spray dryer very commonly used in food and pharmaceutical industries. The drying air is heated by steam. Because of the usually high cost of the final product, more complex separation equipment is justified or necessary. A spray drying system with the atomizer located in the middle of the drying chamber is shown in Figure 10.32. With two air inlets it is possible to control the spray characteristics very well. The secondary air entering the bottom of the chamber decreases the moisture content of the product.

The spray dryer shown in Figure 10.33 enables drying of high-consistency paste-like feeds, such as plastic materials, some salts, and dyestuffs. Here, the feed is transported into the atomizer by means of a screw feeder and is sprayed using a two-fluid atomizer. At very high temperatures (up to 850°C) the evaporation rate is about 25 kg/(m³ h).

One of the applications of the spray drying process is the drying of waste sludge, which is shown in

TABLE 10.5
Heat Consumption for Evaporation of 1 kg Water (kJ)

Membrane process	140
Evaporator 1 stage	2600
Evaporator 2 stages	1300
Evaporator 6 stages	430
Evaporator 6, with thermocompression	370
Evaporator 6, with mechanical compression	400
Spray drying process	Up to 6000

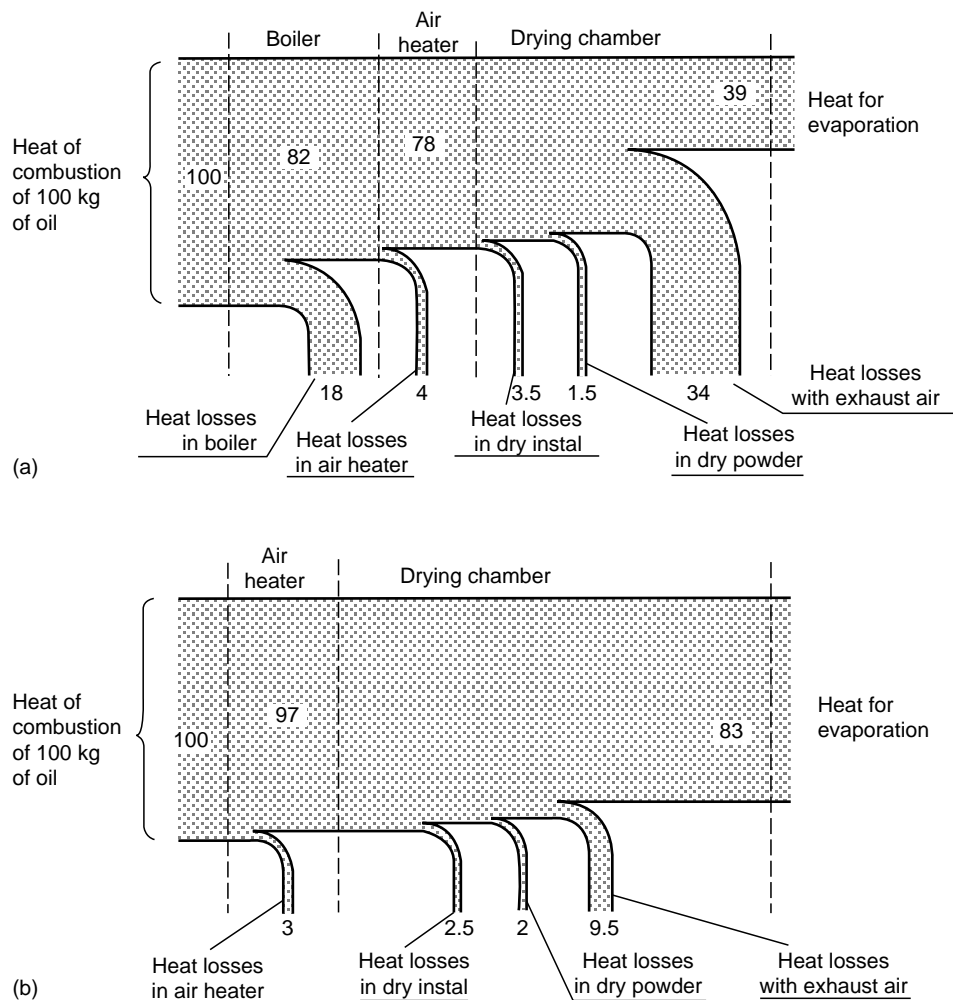


FIGURE 10.28 Thermal efficiency for two spray dryers operating with different energy sources and temperature ranges: (a) $\eta = 39\%$, steam, $T_{A\text{ INLET}} = 180^\circ\text{C}$, $T_{A\text{ OUTLET}} = 90^\circ\text{C}$; (b) $\eta = 83\%$, oil, $T_{A\text{ INLET}} = 500^\circ\text{C}$, $T_{A\text{ OUTLET}} = 65^\circ\text{C}$. (From Strumillo, C., *Podstawy Teorii i Techniki Suszenia*, 2nd Ed., Wydawnic-two Naukowo-Techniczne, Warsaw, Poland, 1983.)

TABLE 10.6
Spray Dryer Layouts

Layout	Drying Medium and Feed	Heating	Application
Open cycle	Air and aqueous	Direct or indirect	Exhaust air to atmosphere
Closed cycle	Inert gas and nonaqueous	Indirect (liquid phase or steam)	For evaporation, recovery of solvents; prevention of vapor emissions; elimination of explosion; fire hazards
Semiclosed (standard)	Air and aqueous	Indirect	For handling materials that cannot contact flue gases; elimination of atmospheric emissions
Semiclosed (self-inertizing)	Air with low O_2 content and aqueous	Direct	Products with explosion characteristics; elimination of powder and odor emissions
Two stage	Any of the above layouts plus fluid-bed, agglomerators, flash dryers as second stage		For improved powder properties, improved heat economy
Combination	Spray and fluidizer	Direct	Improved energy consumption, reduced capital costs

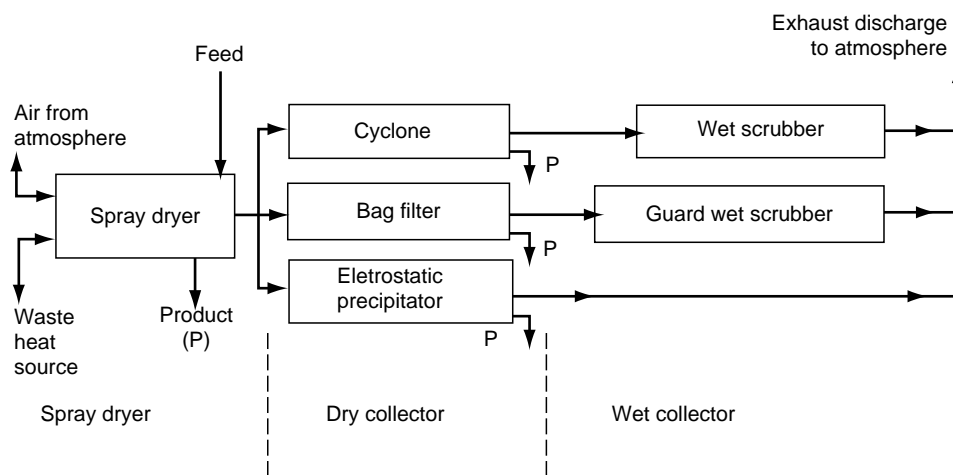


FIGURE 10.29 Open-cycle layout.

Figure 10.34. Solid wastes are burned in a furnace at a temperature of 800–1000°C. The heat of the exhaust gas is used in a spray dryer, where the sludge is atomized and dried. The outlet gas temperature is about 200°C. The typical drying chamber arrangement is shown in Figure 10.35.

A closed-cycle layout shown schematically in Figure 10.36 is used for drying materials containing flammable organic solvents. In such cases it is not possible to discharge the solvent vapor into the atmosphere

because of toxicity and odor problems. The closed-cycle layout prevents leakage of vapor or powder and minimizes explosion and fire hazards, ensuring total solvent recovery at the same time. The system is operated at a slight pressure to prevent any inward leakage of atmospheric air; open-cycle plants are usually operated at a slight vacuum. An example of a closed-cycle layout is shown in Figure 10.37.

A semiclosed-cycle layout is shown in Figure 10.38 in three types [22]:

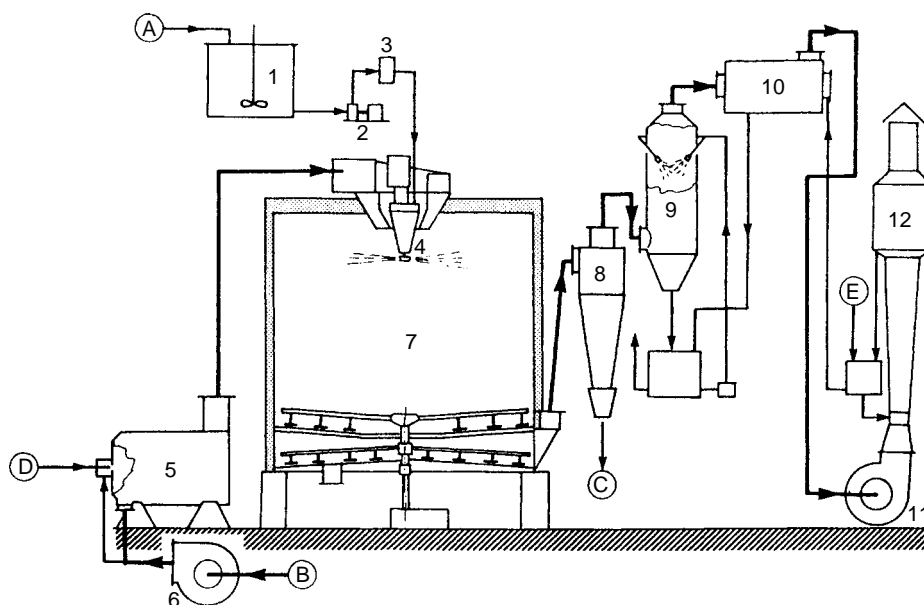


FIGURE 10.30 Spray drying of superphosphate: (1) feed tank; (2) feed pump; (3) filter; (4) atomizer; (5) hot gas generator; (6) fan; (7) drying chamber; (8) cyclone; (9) wet scrubber; (10) absorber; (11) exhaust fan; (12) gas outlet; (A) feed; (B) air; (C) product; (D) fuel; (E) water. (From Strumillo, C., *Podstawy Teorii i Techniki Suszenia*, 2nd Ed., Wydawnic-two Naukowo-Techniczne, Warsaw, Poland, 1983.)

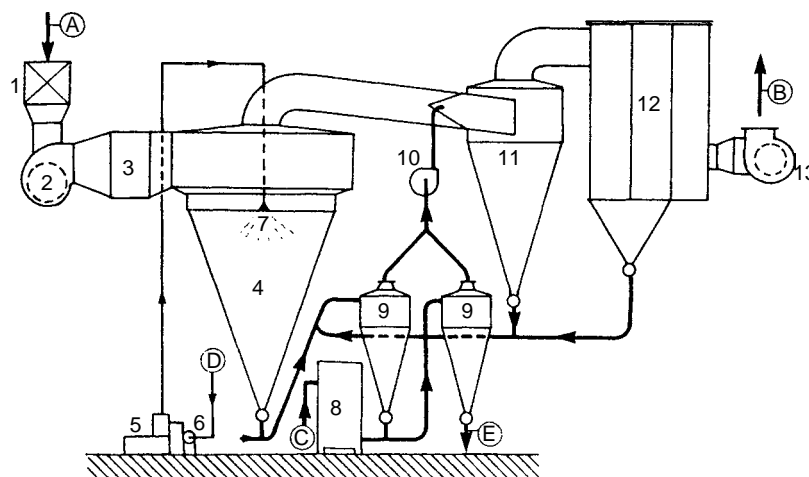


FIGURE 10.31 Spray drying of a heat-sensitive material: (1) filter; (2) fan; (3) air heater; (4) drying chamber; (5) high-pressure pump; (6) secondary pump; (7) pressure nozzle atomizer; (8) air conditioner for transporting air; (9) cyclone; (10) fan; (11) secondary cyclone; (12) bag filter; (13) exhaust fan; (A) air; (B) exhaust air; (C) secondary air; (D) feed; (E) product. (From Strumillo, C., *Podstawy Teorii i Techniki Suszenia*, 2nd Ed., Wydawnic-two Naukowo-Techniczne, Warsaw, Poland, 1983.)

1. Partial recycle cycle
2. Vented closed cycle with indirect heater
3. Vented closed cycle with direct heater

In a semiclosed cycle, only a part of the drying air is exhausted to atmosphere, thus maintaining powder emission. Most of the drying medium is recycled. Partial recycle (Figure 10.38a) is an old idea that is now used again to utilize the heat in the outlet air and thus reduce the fuel consumption. A reduction of up

to 20% in fuel consumption is possible when the outlet air temperature exceeds 120°C. The amount of recycled air is about 50% of the total drying air fed to the dryer.

The vented closed cycle with indirect heating (Figure 10.38b) operates under a slight vacuum. The amount of air exhausted to atmosphere is very small and corresponds to the air that enters into the system through normal leakage, as the dryer is not manufactured to be gastight. This layout can be successfully used to handle toxic materials and those having odor problems, or when contact with combustion gases must be avoided.

The direct-heater layout (Figure 10.38c) can be used for non-heat-sensitive materials. A special application of this layout is the so-called self-inertizing system for materials at high risk for fire and explosion. Any excess air entering the system by leakage is used as additional combustion air in the heater, thus passing through a flame zone where it is deactivated prior to exhaust to the atmosphere. A self-inertizing system is shown in Figure 10.39.

The advantages of this system are lower fuel consumption and lower cost in comparison with the closed-cycle layout, which must be gastight. It is used, for example, in the pharmaceutical industry for drying of fermentation residues.

The layouts described so far represent one-stage spray drying processes, in which dryings are carried out in a single unit. Developments are oriented to achieve higher product quality at higher thermal efficiency. This has resulted in the design of a two-stage

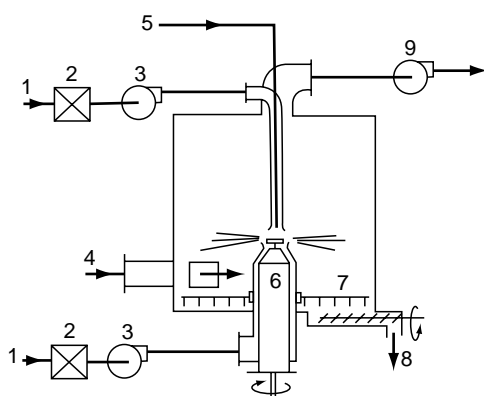


FIGURE 10.32 Spray dryer with atomizer located in the middle of chamber: (1) air inlet; (2) heater; (3) fan; (4) secondary air inlet; (5) feed; (6) rotary atomizer; (7) powder discharge system; (8) powder outlet; (9) exhaust fan. (From Strumillo, C., *Podstawy Teorii i Techniki Suszenia*, 2nd Ed., Wydawnic-two Naukowo-Techniczne, Warsaw, Poland, 1983.)

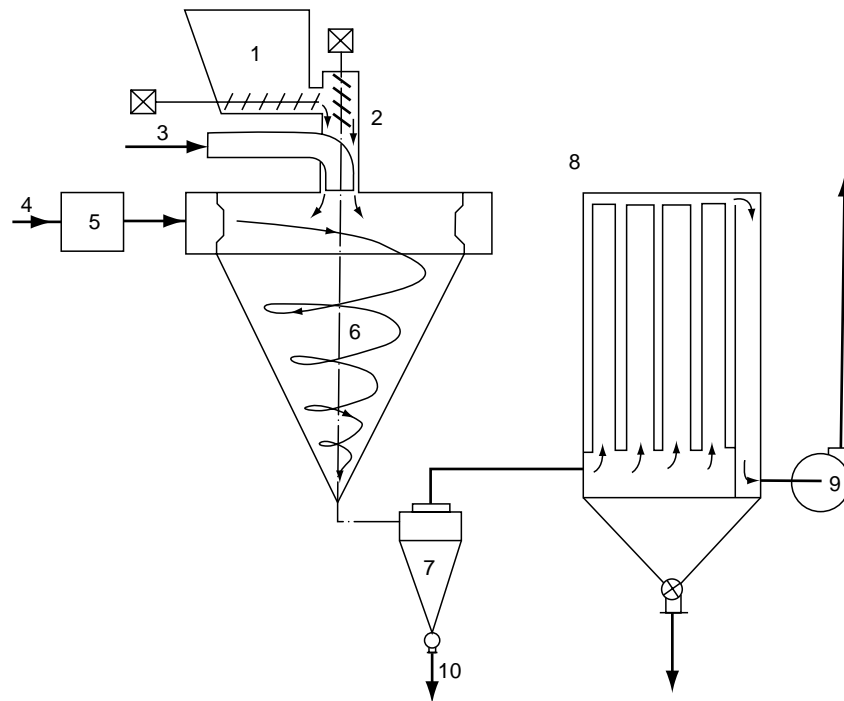


FIGURE 10.33 Spray dryer for paste-like materials: (1) feed tank; (2) atomizing device; (3) compressed air inlet; (4) drying gas; (5) heater; (6) drying chamber; (7) cyclone; (8) bag filter; (9) exhaust fan; (10) product outlet. (From Strumillo, C., *Podstawy Teorii i Techniki Suszenia*, 2nd Ed., Wydawnic-two Naukowo-Techniczne, Warsaw, Poland, 1983.)

spray drying process. This system uses lower energy (~20% less) and produces a powder with “instant” characteristics, which are often required in the food

industry. An example of a two-stage drying process used to dry milk is shown in [Figure 10.40](#).

In a two-stage process, the spray dryer is the first stage; the material is dried up to 10% (in the case of milk) instead of up to the final 3–5% moisture. The final moisture content is achieved in the second stage, which is usually a vibro-fluidized bed dryer. The second stage consists of two parts. In the first part, the powder is dried and in the second it is cooled. The product is agglomerated to achieve instantizing.

The thermal efficiency of a two-stage system is better than that of a single unit because the temperature of the outlet air is lower, about 80°C, instead of 100°C or more in a single unit. This allows the use of a higher inlet air temperature without degradation of the dried material.

Three-stage drying was introduced several years ago to further improve the thermal efficiency of the drying process. A typical three-stage dryer consists of the spray drying chamber as the first stage in which the conical bottom is placed. The second stage is a static fluid bed. The third stage represents an external fluid bed for final drying and cooling [18].

[Figure 10.41](#) shows another three-stage dryer used in the food industry [30]. In the first stage, the feed is dried to a moisture content of 10–20%, depending on

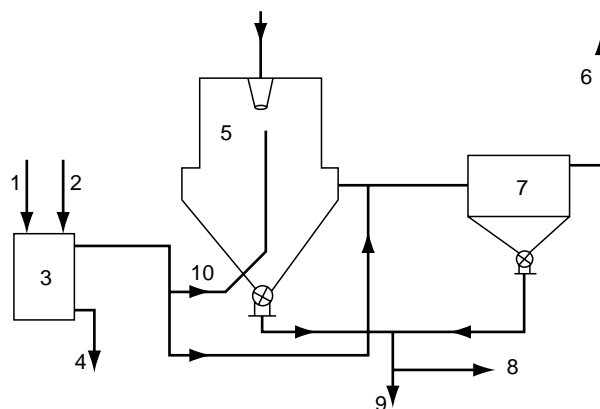


FIGURE 10.34 Drying of sludge: (1) dry material from spray dryer, see Reference 9; (2) solid wastes; (3) combustion chamber; (4) ashes discharge; (5) spray dryer with rotary atomizer; (6) exhaust gas; (7) electric filter; (8) dry product; (9) dry material to combustion chamber; (10) highly consistent sludges. (From Strumillo, C., *Podstawy Teorii i Techniki Suszenia*, 2nd Ed., Wydawnic-two Naukowo-Techniczne, Warsaw, Poland, 1983.)

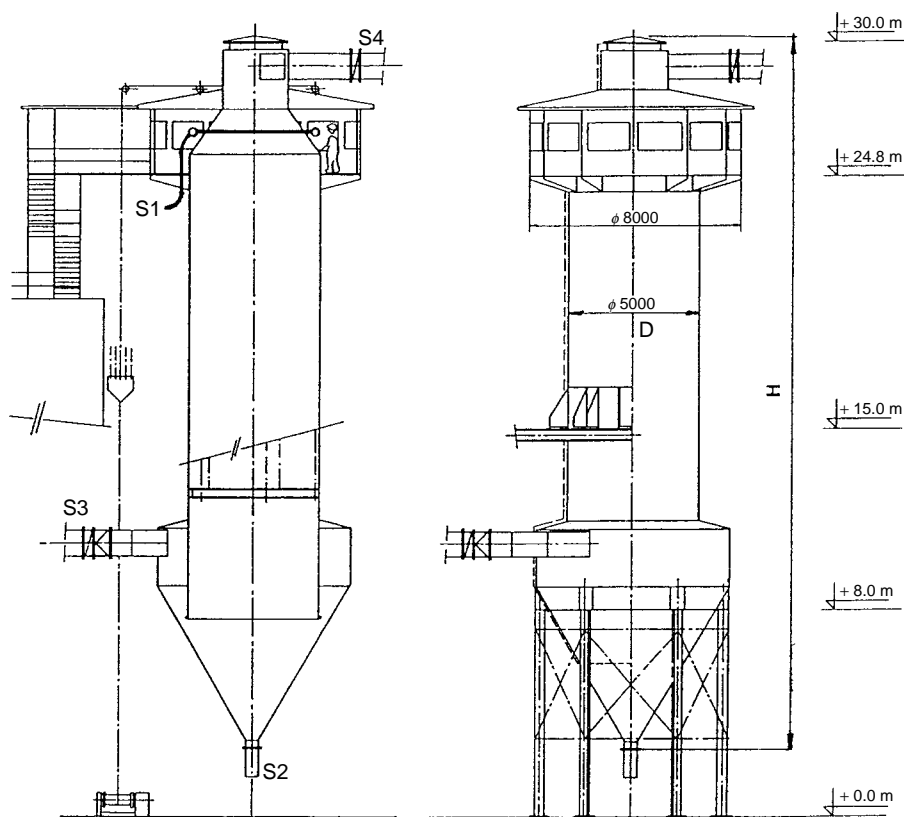


FIGURE 10.35 Drying chamber arrangement.

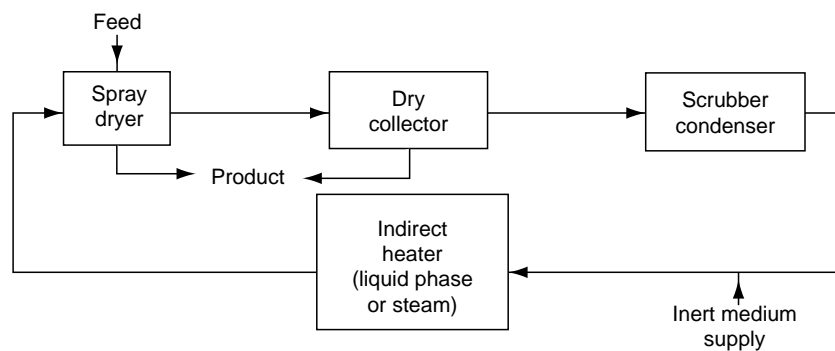


FIGURE 10.36 Closed-cycle layout.

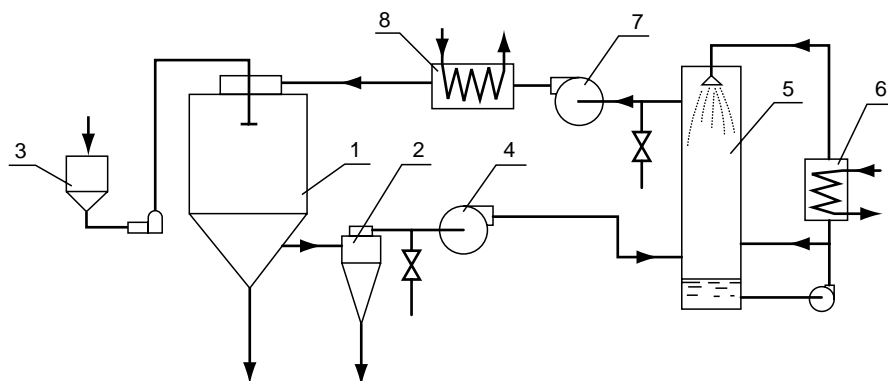


FIGURE 10.37 Example of a closed-cycle layout: (1) spray dryer; (2) cyclone; (3) feed tank; (4) exhaust fan; (5) wet scrubber; (6) scrubber cooling system; (7) fan; (8) heater. (From Strumillo, C., *Podstawy Teorii i Techniki Suszenia*, 2nd Ed., Wydawnictwo Naukowo-Techniczne, Warsaw, Poland, 1983.)

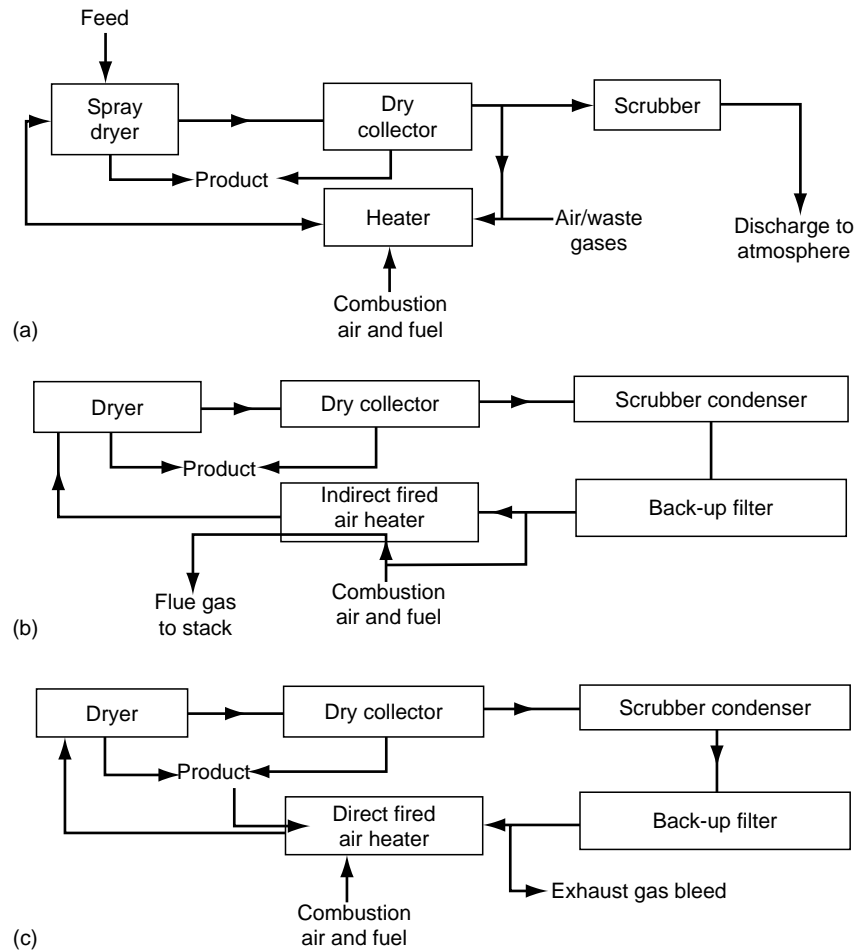


FIGURE 10.38 Semiclosed-cycle layout: (a) partial recycle cycle; (b) vented closed cycle with indirect heater; (c) vented closed cycle with direct heater. (From Masters, K., *Spray drying, Advances in Drying*, Vol. 1, Ed. A.S. Mujumdar, Hemisphere/McGraw-Hill, New York, 1980, pp. 269–298.)

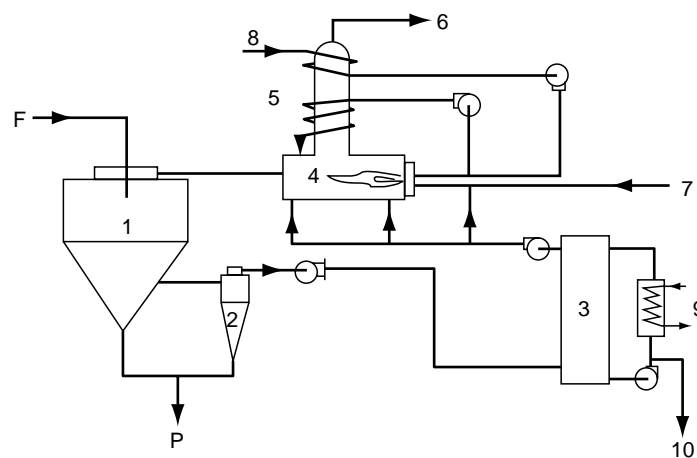


FIGURE 10.39 Self-inertizing spray drying system: (1) drying chamber with rotary atomizer; (2) cyclone; (3) scrubber condenser; (4) direct-fired heater; (5) heat exchanger for waste-heat recovery; (6) exhaust to atmosphere; (7) fuel; (8) combustion air; (9) cooler; (10) condensate; F, feed; P, product.

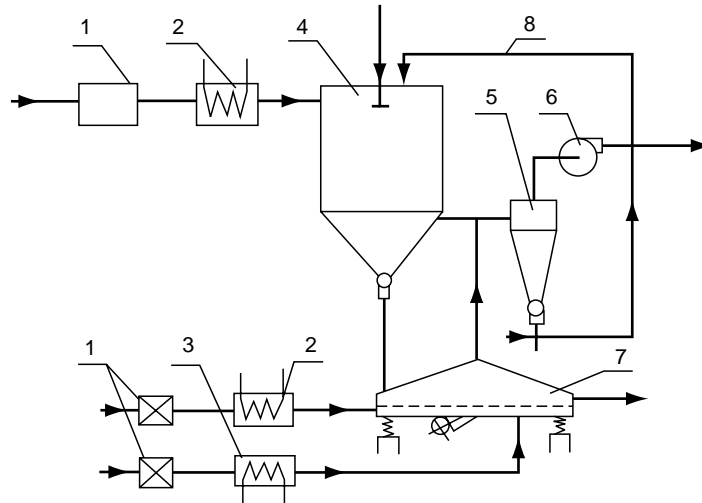


FIGURE 10.40 A two-stage spray drying process for milk: (1) air filter; (2) heater; (3) cooler; (4) spray dryer; (5) cyclone; (6) exhaust fan; (7) fluidized-bed dryer; (8) return line of fine powder.

the product. The semidried powder is deposited on a moving belt (woven polyester filament) situated at the bottom of the primary spray chamber. The drying air is distributed through the belt and the powder with a relatively high velocity. This is the second stage of drying (crossflow drying). After a short stabilization period in a retention section, the powder is conveyed to the third drying stage in which drying is completed by low-temperature air. The final section is a cooling stage. The final product consists of agglomerates with “instant” properties. The temperature of the exhaust air is even lower than in two-stage systems ($65\text{--}70^{\circ}\text{C}$), thus higher thermal efficiency is obtainable.

10.3.2 ENERGY SAVINGS

Spray drying is a very energy-intensive process. There are three main reasons for this:

1. It is necessary to supply the specific heat of evaporation in a short time.
2. The temperature difference across the drying chamber is relatively small because the heat-sensitive materials, which are mostly spray-dried, do not permit the use of high-temperature inlet air. The required quality of final product does not permit the use of low-temperature outlet air, either.
3. An appreciable amount of heat is lost with the exhaust air. Nonideal performance of the powder collection system causes loss of dry product.

Hence, the below-listed energy saving strategies for spray drying processes correspond to the three main categories that are mentioned above.

1. Remove as much water from the feed as possible before drying. It is shown in [Table 10.7](#) that preconcentration of the feed results in significant energy savings. The most effective is mechanical dehydration followed by membrane processes or centrifugal separation. The only requirement is that the feed to the spray dryer must be pumpable. The effect of feed concentration on the heat consumption in a spray dryer is shown in [Table 10.7](#) [22].
2. Increase the inlet air temperature or decrease the outlet air temperature. The effect of both the temperatures on the thermal efficiency is evident in [Figure 10.42](#) [14]. This requirement can be met by using two or more stages when possible.
3. If a wet scrubber is used to clean the exhaust air and to avoid high losses of dry product, the hot scrubber liquid can be used as feed process water or the feed should be concentrated in the scrubber. This is not always possible because of hygienic problems in the food industry. In such cases, a good method of energy savings is to lead the exhaust air from the cyclone into a bag filter, where the product is recovered. The exhaust air then passes through a countercurrent exchanger where the inlet air is preheated up to 75°C . The heat exchanger must be carefully designed so that it is self-cleaning, as the exhaust air always contains fine particles.

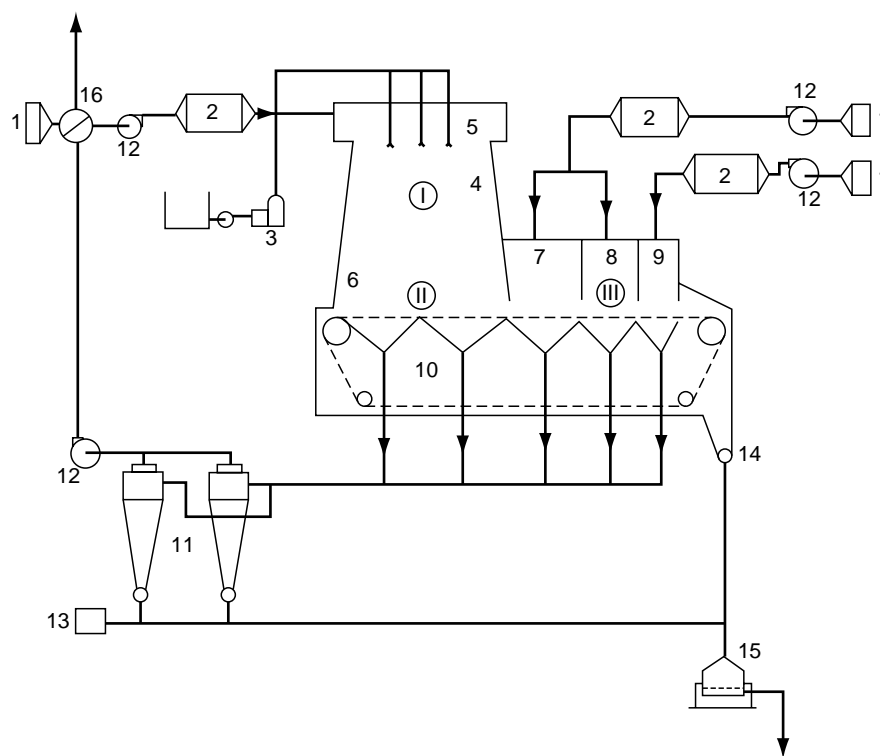


FIGURE 10.41 A three-stage spray drying process (FILTERMAT): (1) air filter; (2) heater-cooler; (3) high-pressure pump; (4) nozzle system; (5) air distributor; (6) primary drying chamber; (7) retention chamber; (8) final drying chamber; (9) cooling chamber; (10) FILTERMAT belt assembly; (11) cyclones; (12) fan; (13) fine recovery system; (14) FILTERMAT powder discharge; (15) sifting system; (16) heat recovery system; I, first drying stage, II, second drying stage; III, third drying stage.

10.3.3 SAFETY ASPECTS

In spray drying operations a potential danger of explosion and fire can exist under certain conditions [7].

Fire hazards exist if any of the following conditions occur:

- The temperature of the air-product mixture reaches a flammability limit (see Table 10.8).
- The oxygen content in the drying medium is high.

(See relevant chapter of this handbook for details on fire and explosion hazards.)

Dry skim milk has the lowest inflammation temperature in a layer; sugar and some detergents have the lowest temperature when in a cloud in a spray dryer. Temperature data for the drying medium in a spray dryer for three typical materials are shown in Figure 10.43. The temperatures are given at the following locations: T1, heater outlet; T2, drying chamber inlet; T3, drying chamber outlet; T4, exhaust air; T5, pneumatic transport location. Useful data for

TABLE 10.7
Heat Consumption for Various Feed Concentrations

Feed Solids (%)	Approximate Heat Consumption (kJ/kg Powder)
10	23.65×10^3
20	10.46×10^3
30	6.17×10^3
40	3.97×10^3
50	2.68×10^3

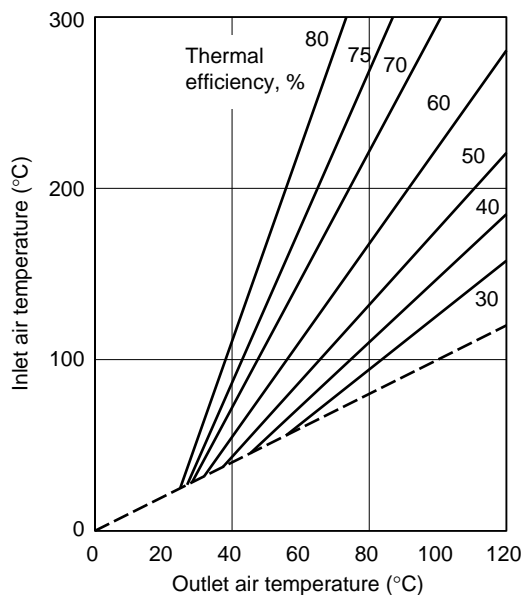


FIGURE 10.42 The effect of both inlet and outlet air temperature on thermal efficiency. (From Kessler, H. G., *Heat conservation in concentration and spray drying of milk products*, *Drying '80*, Vol. 1, Ed. A. S. Mujumdar, Hemisphere/McGraw-Hill, New York, 1980, pp. 339–342.)

minimum ignition temperatures of milk products can be found in Ref. [34].

It is evident that the drying air temperature nearly always exceeds the inflammation temperature in a layer. This means that ignition is possible whenever

dry powder deposits are formed. In the case of detergents, ignition is also possible in the cloud.

Ignition may be initiated for any one of the following reasons:

- Spontaneous combustion in product deposits
- Hot solid particles entering the dryer with drying gas
- Spark generation through friction
- Electrical failure
- Static electricity discharge

Fire prevention methods are based on proper operation and cleaning of the dryer. An important safety device is an automatic control system (ACS), as it can maintain the temperature within $\pm 1^\circ\text{C}$ range whereas the manually controlled temperature may vary $\pm 10^\circ\text{C}$. The control system should include a detection system of parameter deviation, deposit detection, drying gas analysis, and continuous monitoring of the humidity and temperature at selected locations.

The explosion hazard depends on the value of the critical particle concentration, oxidation velocity, and the explosion pressure. If any one of these conditions is reached, there is danger of ignition. A comparison of both explosive and process concentration for three products is shown in Figure 10.44. The powder concentrations are given in the following locations: C1, drying chamber; C2, cyclones; C3, exhaust fan; C4, wet scrubber; and C5, pneumatic transport. Figure 10.44 shows that even under normal operating

TABLE 10.8
Fire and Explosion Data for Some Spray-Dried Materials

Powder	Inflammation Temperature		Minimum Explosion Concentration (g/m ³)	Explosion Pressure
	Layer (°C)	Cloud (°C)		
Wheat starch	—	410–460	7–22	High
Pudding powder	—	—	20	High
Sugar powder	—	360–410	17–77	High
Cream topping	—	—	6.3	High
Monoglyceride	290	370	16	High
Monoglyceride + skim milk	282	435	32	High
Baby food	205	450	36	High
Milk concentrate	190–203	440–450	22–32	High
Skim milk	134	460	52	High
Milk	142	420	54	High
Buttermilk	194	480	56	High
Skim milk + whey + fat	183–240	460–465	20–24	High
Coffee extract	160–170	450–460	50	
Cocoa	170	460–540	103	High
PVC	—	595	40	High
Detergents	160–310	360–560	170–700	Low

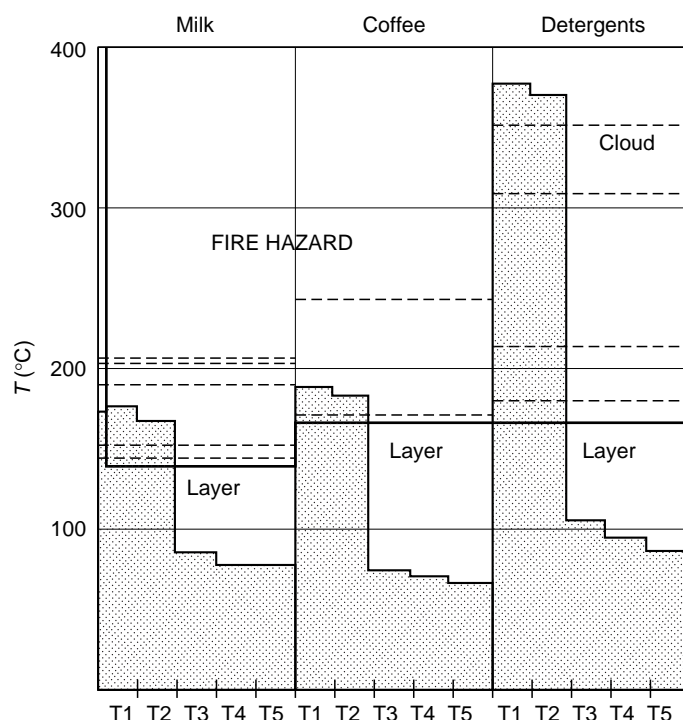


FIGURE 10.43 Temperature data for the drying air in a spray dryer. (From Filka, P., Safety aspects of spray drying, *Drying '84*, Ed. A. S. Mujumdar, Hemisphere/McGraw-Hill, New York, 1984.)

conditions the danger of explosion does exist in a normal spray dryer. The explosion hazard is high above the solid line. The zone between the thick and dotted lines is potentially dangerous as the powder concentration there exceeds 50%. All process equipment in which the operating dust concentration exceeds 50% of the lower explosive concentration must be provided with a safety device. The spray drying chamber, cyclones, and pipes should be equipped with safety doors and explosion vents [19].

When highly inflammable materials are dried it is necessary to use inert atmosphere. A gastight closed-cycle spray dryer with nitrogen as the drying medium may be used in such cases. When explosion hazard exists, a self-inertizing system is recommended. Good maintenance practice cannot be overemphasized.

For additional information and further references on fire and explosion hazards in drying of solids, the reader is referred to the appropriate chapter in this handbook.

10.3.4 CONTROL SYSTEMS

Spray dryers can be controlled either by manual control system (MCS) or by automatic control system (ACS). MCS is restricted to small spray dryers or when the handled material dries easily. In other

cases ACS should be adopted. The outlet air temperature from the drying chamber is the parameter that is controlled as it represents the quality of final product and is easy to monitor. When required for environmental reasons, the control of powder content in exhaust air is also monitored.

Two basic types of control systems commonly employed are shown in Figure 10.45 and Figure 10.46 [23]. Control system A is used mostly with wheel atomizer processes and consists of two circuits. The first circuit controls outlet air temperature by feed rate regulation. The inlet air temperature, controlled by the second circuit, is corrected by the fuel combustion rate. The control system is provided with a safety system that prevents any damage in case of failure in the feed system followed by rapid increase in the outlet air temperature.

Control system B is used usually with nozzle dryers, where the feed rate is to be constant. The outlet air temperature is controlled by the regulation of combustion rate in the gas or oil heater or steam pressure, when a steam heater is used. Similar safety arrangements, as in the previous control system, are applied here.

Any spray dryer control system can be run either automatically or semiautomatically. A fully automatic control system is recommended when the product quality should meet very stringent requirements and

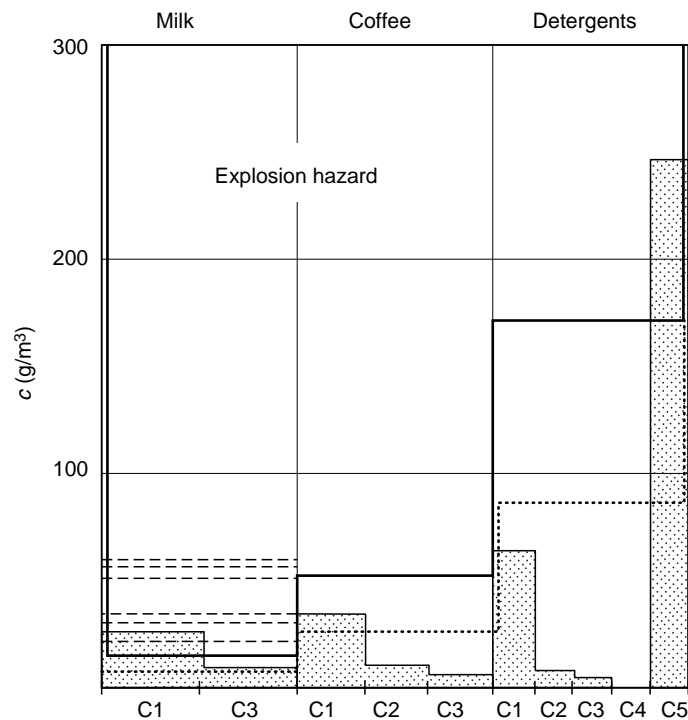


FIGURE 10.44 Powder concentration data for a spray dryer. (From Filka, P., Safety aspects of spray drying, *Drying '84*, Ed. A. S. Mujumdar, Hemisphere/McGraw-Hill, New York, 1984.)

when lower operating costs are essential. A timing device starts up the dryer in a predetermined sequence. In case of failure in the control equipment it is always possible to employ manual control and to continue production without interruption. The shut-down and cleaning operations are programed and controlled by means of timing equipment. The ACS must include a control system for fire detection, a

system for fire or explosion prevention, and a programable system for countermeasures.

10.3.5 SELECTION OF SPRAY DRYERS

Several considerations are critical in the assessment of a material to be spray-dried. Some questions may be asked, for example, what is the material? Is there any

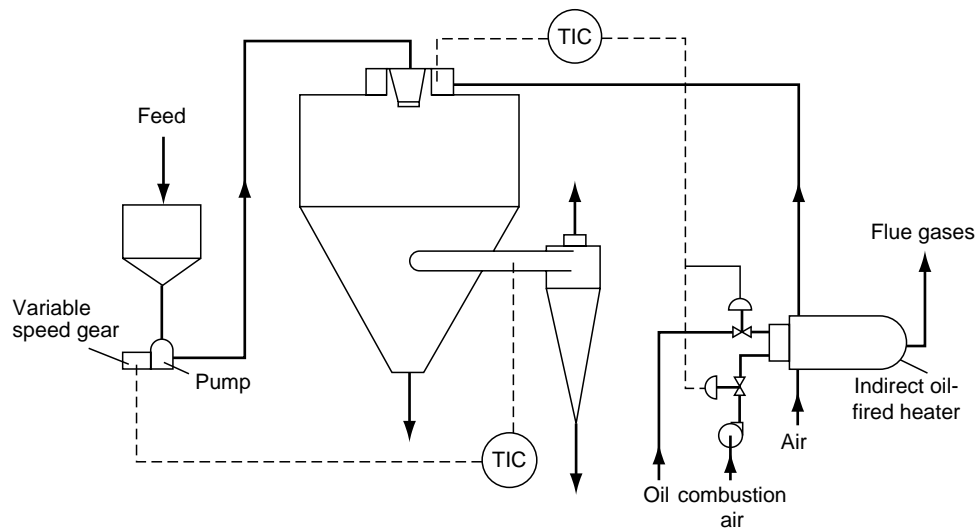


FIGURE 10.45 Control system A. (From Masters, K., *Spray Drying*, Leonard Hill Books, London, 1979.)

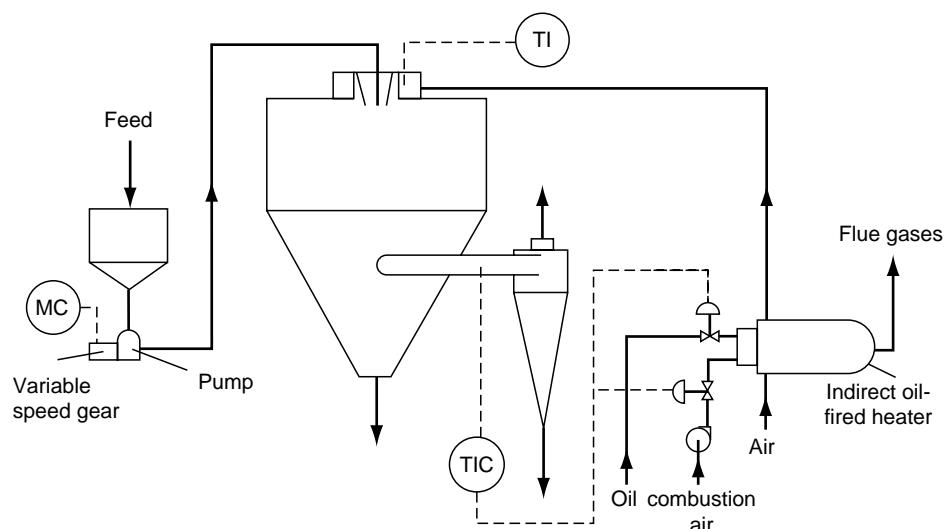


FIGURE 10.46 Control system B. (From Masters, K., *Spray Drying*, Leonard Hill Books, London, 1979.)

historical data on spray drying this material? Is it temperature-sensitive? Is the material hazardous in any way? Is it aqueous or solvent-based? Does the powder present an explosion hazard? The feed physical properties and product requirements, e.g., viscosity, solid concentration, particle size, residual moisture, bulk density etc., also are important considerations. The next step is evaluating the product in a small laboratory-scale dryer to determine starting parameters for plant scale.

On the basis of their operating characteristics, Huang and Mujumdar [51,52] generated a selection tree for various types of spray dryers. Note that selection of the spray drying system also must consider product quality and pilot-scale test results.

10.4 DESIGN CONSIDERATIONS AND NUMERICAL MODELS

10.4.1 EMPIRICAL DESIGN AND PROCEDURE

Masters [61,62] and Gauvin and Katta [12] have suggested a simplified method to design a spray dryer. Huang et al. [53] provided an empirical method to design a centrifugal spray dryer. They reported an empirical correlation for the drying intensity of a centrifugal spray dryer as follows:

$$q = \frac{(T_1 + 273)^{3.4287}}{(T_2 + 273)^{3.34}} \quad (10.20)$$

where q is the drying intensity (kg/m^3) viz. evaporation rate per unit drying chamber volume, T_1 and T_2

are the inlet drying air temperature and outlet air temperature ($^{\circ}\text{C}$), respectively.

Here, we summarize their design procedure as follows:

- Collect design information, such as solid content in feed, feed temperature, evaporation rate, product rate, residual moisture in product.
- Check feed properties, such as latent heat of volatile, viscosity, heat sensitivity, boiling point, solid property (abrasive, corrosive, etc.) in feed.
- Identify the hazard conditions of feed, such as solvent and its property, flammable point, toxicity
- List the dry product properties, such as powder size distribution, particle shape, bulk density, specific heat, glass transition temperature, hygroscopicity, powder flammability.
- Determine air conditions, such as relative humidity, ambient temperature (annual minimum and maximum) and ambient pressure.
- Survey patents, published papers, existed experience and pilot test to decide reasonable operating parameters, such as inlet air temperature, outlet air temperature. Finally select the inlet air temperature as high as possible and outlet air temperature as low as possible if the product requirements are satisfied, based on pilot experiment, published papers, patents and experience.
- Choose the spray drying system, e.g., process layout (open, closed, etc.), type of air-spray contact, atomization way. [51,52]

- Carry out drying chamber selection, based on the atomizer and process layout.
- Select fuel type and its calorific value; decide the air heating way, such as steam heater, electric heater, oil furnace, coal furnace.
- Select the product collecting way, such as cyclone, bag filter, wet scrubber.
- Carry out heat and mass balances, airflow rate for drying, heater capacity. Computational fluid dynamics (CFD) simulations for such a spray dryer may be performed to determine trends as accuracy of CFD models are still not well defined.
- Other selections, such as control systems, collector ways, pumps, fans, pipe size and pipe fittings follow.
- Estimate and decide the droplet residence time needed in the drying chamber depending on the pilot test or empirical data. We summarize the empirical data for air residence time in the drying chamber in Table 10.9. It should be noted that the drying time of droplet is different from the residence time of particle and air in the drying chamber.
- Draw spray drying system layout.
- Estimate investment and operating costs.
- Provide relative requirements to civil engineering, power supply, water supply, and fuel supply if needed, etc.
- Finalize flow sheet.

10.4.2 COMPUTATIONAL FLUID DYNAMICS SIMULATION OF SPRAY DRYING

We are currently unable to express precisely the spray–air contact, the powder trajectories, and airflow patterns. The kinetics of the system and optimum performance of the drying in a given drying chamber cannot be predicted on theoretical basis alone. Hence, specification and design of spray dryers are still based on the pilot tests and existing industrial experience. However, CFD technique is developing very rapidly because of its advantages, such as low cost (relative to experiments); high speed; detailed information obtained etc. Some authors [27–29,41,

48–50,56,59,71,74], have used the CFD technique and commercial codes to simulate the spray drying system. CFD procedures are becoming part of both testing new spray dryer design and investigating problem areas of performance in an existing spray dryer. We provide our CFD simulation results in a cylinder-on-cone in Figure 10.47 [48] as a sample.

The predicted airflow pattern (Figure 10.47a) shows that there is a large recirculation zone in the top of chamber. From Figure 10.47b (velocity vector profiles at different levels in the drying chamber), we find that there is a nonuniform velocity in the core region of the chamber. A reverse velocity results in the recirculation zone. The particle trajectory in Figure 10.47c shows that the particles are tracing in a very complex way. We cannot read them clearly in the total trajectory chart. It also tells us that the particle tracing history in the drying chamber is very difficult to be identified. Such simulations can help us choose new chamber geometries that avoid wall deposits and use dryer volume more effectively.

The following diffusion equation can be used to simulate a single droplet drying (falling rate period) in a hot moving airflow:

$$\frac{\partial C}{\partial t} = D \left(\frac{\partial^2 C}{\partial r^2} + \frac{2}{r} \frac{\partial C}{\partial r} \right) \quad (10.21)$$

where C is the molar concentration of droplet (mol/m^3) and D is diffusion coefficient in droplet which is assumed as a constant (m^2/s).

10.5 NEW DEVELOPMENTS IN SPRAY DRYING

Numerous innovative designs and operational modifications have been proposed in the literature [1–3,32] although very few are readily available commercially due to incomplete knowledge about the new systems.

10.5.1 SUPERHEATED STEAM SPRAY DRYING

Although superheated steam drying was proposed one century ago, the potential of superheated steam

TABLE 10.9
Residence Time Requirements for Spray Drying of Various Products

Residence Time in Chamber	Recommended for
Short (10–20 s)	Fine, non-heat-sensitive products, surface moisture removal, non hygroscopic
Medium (20–35 s)	Fine-to-coarse spray ($d_{\text{mean}} = 180 \mu\text{m}$), drying to low final moisture
Long (>35 s)	Large powder (200–300 μm); low final moisture, low temperature operation for heat-sensitive products

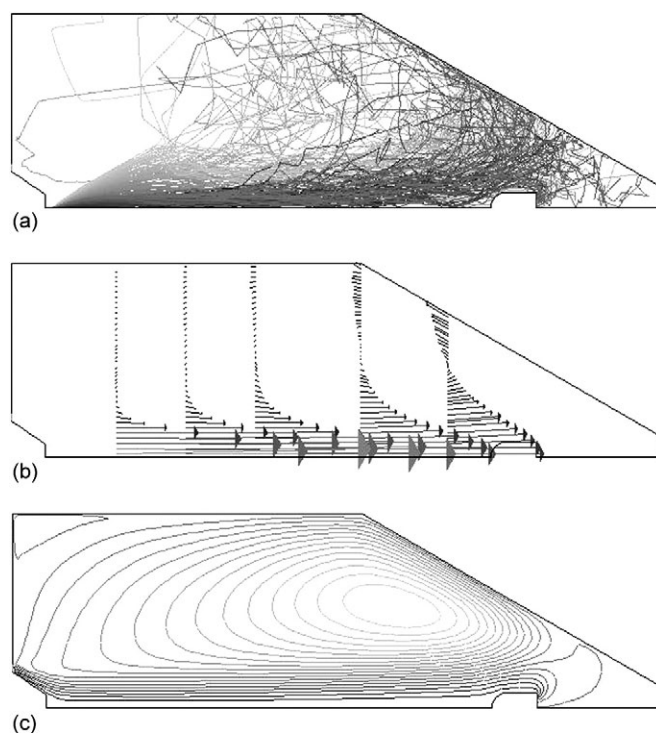


FIGURE 10.47 CFD simulation results in a cylinder-on-cone geometry: (a) airflow pattern; (b) velocity vector (different levels); (c) particle trajectories.

as a drying medium was not exploited industrially for half a century. Some advantages of steam drying are summarized by Mujumdar [64] as follows:

- No fire and explosion hazards
- No oxidative damage
- Ability to operate at vacuum and high operating pressure conditions
- Ease of recovery of latent heat supplied for evaporation
- Better quality product under certain conditions
- Closed system operation to minimize air pollution

Limitations:

- Higher product-temperature
- Higher capital costs compared to hot-air drying
- Possibility of air infiltration making heat recovery from exhaust steam difficult by compression or condensation

Gauvin [13,47] made the first proposal of a superheated steam spray dryer. Raehse and Bauer [67] were issued a patent on superheated steam spray drying. Superheated steam spray drying has been investigated using CFD technique by Frydman et al. [46].

10.5.2 TWO-STAGE HORIZONTAL SPRAY DRYER

To reduce difficulties in scale-up, a horizontal spray dryer has been suggested as an alternative to the conventional vertical types [65]. Kwamya and Mujumdar [58] first did a mathematical model to evaluate this novel concept. Recently, at least two U.S. companies produce horizontal spray dryers, especially for heat-sensitive food and pharmaceutical materials [40,44]. Low drying-temperatures allow flavor retention, high solubility, controlled porosity and density, and fine quality agglomerated products. They also claim substantial energy savings relative to the traditional vertical spray dryer with lower electrical load for given capacity. Both manufacturers use integral bag filters for product collection. Any fines in the exhaust air stream are returned to the dryer. In one of the systems the product is collected at the bottom of the horizontal chamber. However, it is not clear from the product description whether the lower wall of the chamber is also used for drying or cooling of the product. The manufacturers offer the dryer with stainless construction that complies with requirements for sanitary applications. Commercial applications include dryers for egg, albumin, whole egg powder, cheese powder, skim milk, whey protein, etc. Other applications that are feasible include

variety of dairy products, fish products, meat products (chicken or beef broth, beef blood plasma, etc.), and vegetable products (soya milk, soya protein, chocolate, enzymes, glucose, etc.). Both designs use multijet atomizers with indirect-fired burners. Among the advantages claimed are minimum floor-space and building height, low installation and maintenance cost, ability to automatically control the dryer operation, etc. Unfortunately, details of the atomizer layout, droplet size distributions, product quality, etc. are not available. Further research and development (R&D) is needed in this area.

The proposed layout is shown in Figure 10.49. The feed is pumped by a high-pressure pump to spray nozzles. These nozzles can be arranged in various configurations. Here, the spray from the nozzles travels horizontally and then turns down due to gravity. The dried powders are then transported by a conveying belt at the bottom and also conveyed to cyclones or bag filters.

Cakaloz et al. [39] have presented a model for a horizontal spray dryer ($6.0\text{ m} \times 3.0\text{ m} \times 3.0\text{ m}$). Simulation of the moisture-time history of an α -amylase droplet in their spray dryer agreed well with their experiment results. Their results are preliminary, however. Here, we used a validated CFD model to simulate such a horizontal spray dryer. It was seen that the flow pattern in Cakaloz et al. design is not optimal as the main inlet is arranged at the left corner of the chamber. This arrangement makes the spray more likely to hit the top wall and deposit there.

From the process of single-stage horizontal spray dryer in Figure 10.48, it is seen that the dwell time may be too short to let the droplets to dry totally. Hence, to overcome this problem, Mujumdar [65]

proposed a new two-stage, two-dimensional horizontal spray dryer concept, which is designed to allow longer drying times needed for heat-sensitive products and large droplet sizes. The proposed layout is shown in Figure 10.49. The only difference from the single-stage horizontal spray drier (SD) is that there is a fluid-bed dryer, which is installed at the bottom of the horizontal chamber.

This two-stage horizontal SD is yet to be commercialized. The CFD technique can be used to simulate the horizontal spray dryer which can be coupled to a fluid-bed drying model as well. Surface moisture of droplets can be removed rapidly in the spray chamber. Internal moisture takes longer; the latter can be removed in a thin bed, through circulation dryer at the bottom of the chamber.

10.5.3 LOW HUMIDITY SPRAY DRYING

Because spray drying uses high temperature gas as the drying medium and the deposits on the wall cannot be avoided in practical spray dryers, the products are subject to degradation by overheating. Sometimes users have to select freeze dryers rather than spray dryers, especially for drying biochemicals and pharmaceuticals [65]. However, the freeze dryer has high energy consumption, high operating and capital costs than the spray dryer. Freeze drying costs 5–10 times higher than hot-air drying, such as spray drying [45,68].

To reduce degradation of active components in dried powders, we propose a new type of spray dryer, i.e., low dew point (LDP) spray dryer, which uses LDP air at near ambient temperature to 80°C , but with very low humidity. The schematic layout is shown in Figure 10.50. It consists of four sections:

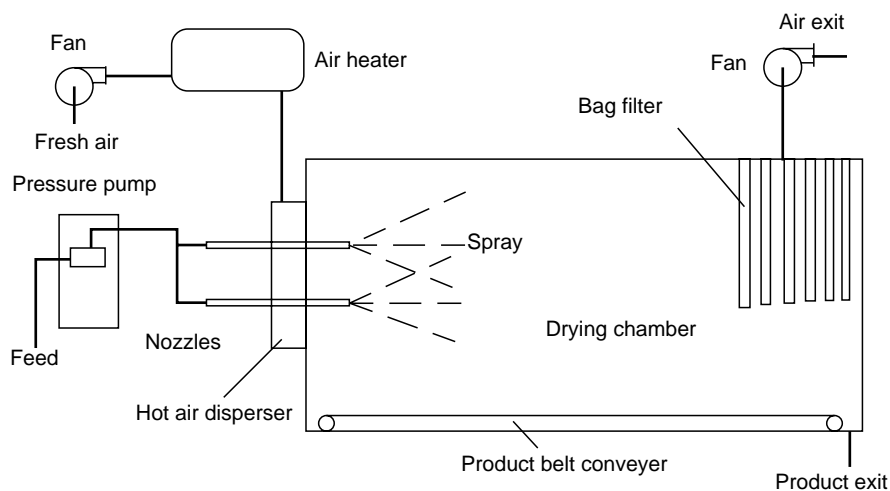


FIGURE 10.48 Schematic layout of the horizontal spray dryer.

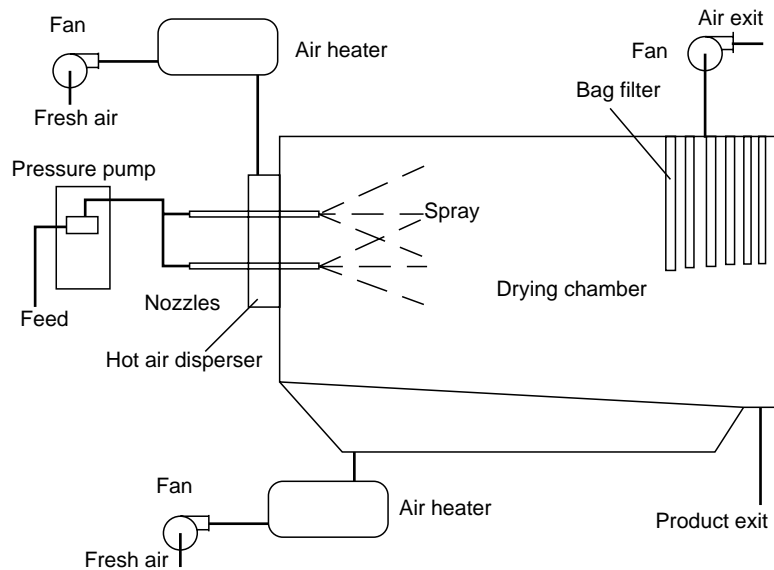


FIGURE 10.49 Proposed layout of a two-stage horizontal spray dryer.

drying air preprocessing system including the dehumidification and heating; feed preparation system; atomization of feed and drying system including the atomizer, air disperser, and drying chamber, etc., and the product collection system.

The main difference of LDP spray dryer from the normal spray dryers is that here we use a dehumidifier to reduce the inlet gas humidity. The mass transfer rate between gas and droplet can be expressed as

$$\frac{dm_p}{dt} = k_g A (H_w - H_a) \quad (10.22)$$

where dm_p/dt is mass transfer rate from the droplet surface (kg/s); k_g is mass transfer coefficient ($\text{kg}/\text{m}^2 \cdot \text{s}$); H_w and H_a represent humidity at the droplet surface and drying medium (kg/kg), respectively. By lowering humidity we can maintain high mass transfer potential at a given temperature.

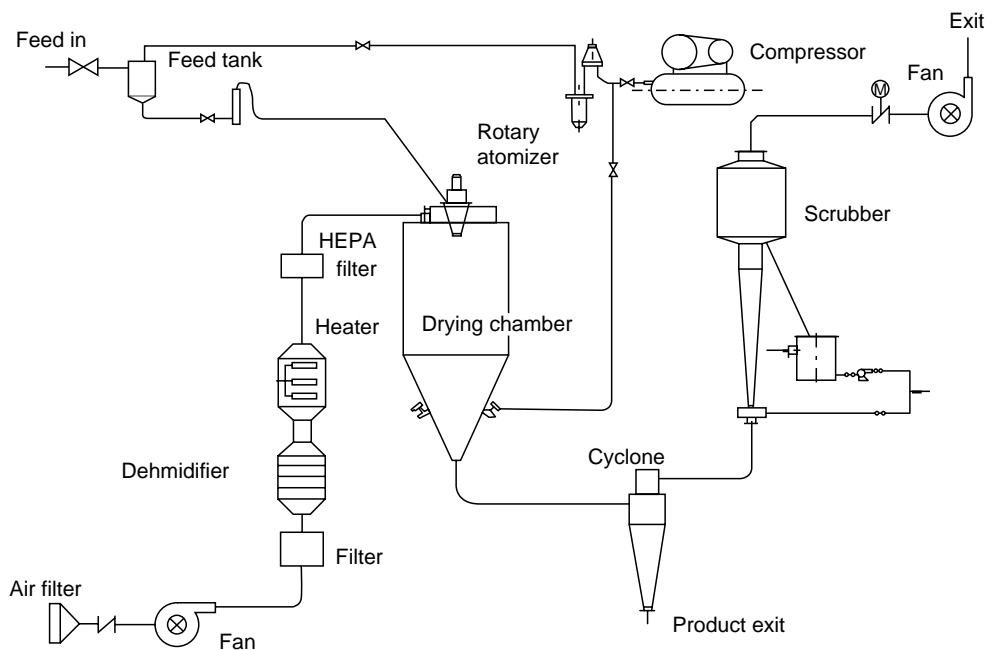


FIGURE 10.50 Schematic layout of the low dew point spray dryer.

Our CFD simulation results show that the LDP spray dryer can be used for drying pharmaceuticals and biochemicals at low temperature. Further work is needed to validate our simulation of this new type of spray dryer.

10.5.4 SPRAY FREEZE DRYING

Freeze drying (FD) is a significant drying technology for foods, pharmaceuticals, and biochemicals, as it can preserve biological activity, flavor, aroma, etc. in highly heat-sensitive materials. Freeze drying is a drying process, in which the solvent in a material is solidified at rather low temperatures and then sublimed from the solid state directly into the vapor state under vacuum. Freeze-drying operation of an aqueous solution involves three steps: freezing of the product, ice sublimation, and solvent vapor removal. However, compared with spray drying, freeze drying is an order-of-magnitude costlier drying process because of need of refrigeration, vacuum, and long running times [38].

Recently, a combination of these two drying processes, i.e., spray freeze drying (SFD), has been found to be very interesting. Such a process is carried out as a batch process as follows:

The solution is atomized in a cryogenic medium, e.g., liquid nitrogen, to freeze sprayed droplets. The dispersion of frozen droplets in the cryogenic medium is taken out and dried in a freeze dryer under vacuum. A typical flow sheet is shown in Figure 10.51. Because of the long freeze-drying time required, this process is

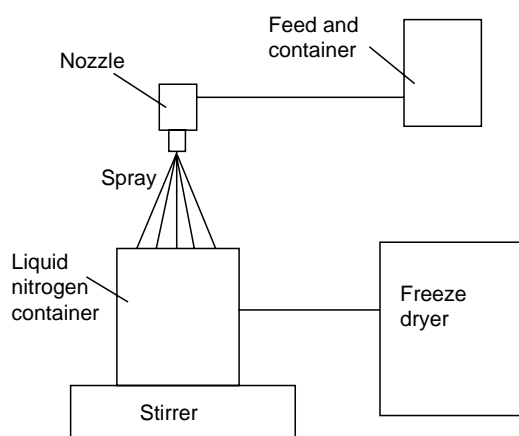


FIGURE 10.51 A schematic flowchart of the conventional spray freeze drying.

carried out in batches rather than continuously. Also, it is suited for high value products of low tonnages only.

Such an SFD process was used to study the protein powders by Sonner et al. [70]. They compared freeze spray-dried powders with those produced by spray drying. A higher quality product is obtained by SFD technology. Sonner et al. also found that the SFD powders were highly porous. They compared the SFD products with those obtained by FD and SD using some model proteins, such as lysozyme and bovine pancreas trypsinogen. It was found that the three drying methods gave similar loss in protein activity, but SD trypsinogen resulted in strongly agglomerated and irregular powders without and visible porous and FD trypsinogen gave a compact cake. Only SFD trypsinogen produced discrete spherical powders.

Yokota et al. [75] used SFD to produce the magnesium sulfate powders. Open pores were found at the dried powder surface. The porosity was about 87–90% with pore size less than 100 nm. Fine powders of several other materials, e.g., calcium phosphate, tetragonal zirconia polycrystals, etc., can also be produced by SFD [55].

We suggest a new process, i.e., a combined atmospheric spray and fluidized-bed freeze drying (ASFBFD) process. The proposed schematic flow diagram is shown in Figure 10.52. It basically consists of a drying or freezing chamber, an internal bag filter, an internal fluidized bed, a liquid nitrogen cooler, a fan, nozzles, a pump, valves, pipes, etc.

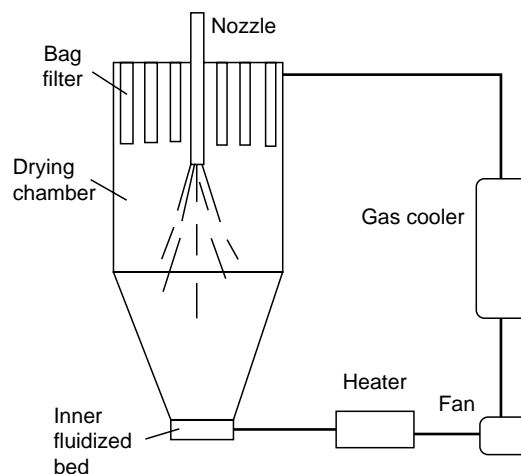


FIGURE 10.52 Proposed flowchart for a batch type atmospheric pressure combined spray and fluidized-bed dryer.

The process is as follows. The feed is transported from the top into the nozzle by a pump, where it is atomized by a nozzle. The fine spray is contacted with very cold air or nitrogen and frozen immediately. Depending on the feed material, this drying medium (air or nitrogen) is cooled to about -90°C by liquid nitrogen and blown by the fan through the fluidized bed perforated plate. The fluidized bed is located at the bottom of the chamber. Because the cold air temperature is low enough to freeze the spray, the frozen particles keep their original physical conditions and fall down to the bottom or fluidized bed due to gravity. Some fine frozen particles may be trapped by the air or nitrogen, but they are separated from the air by the internal bag filter. During this stage, the spray freezing is carried on continuously till the feed is exhausted in this batch process.

After the spray-freezing stage is finished, suitable drying conditions are selected, i.e., the liquid nitrogen cooler is adjusted to meet freeze-drying conditions for the product. The drying time and actual operation conditions need to be determined by testing in the laboratory.

Some researchers have already investigated and showed the feasibility of the atmospheric freeze drying [57,72]. Meryman [63] was the first to demonstrate potential of freeze drying without vacuum. He showed that the drying rate of a material undergoing freeze drying is a function of ice temperature and the vapor pressure gradient between the site of water vapor formation and the drying media, rather than the total pressure in the drying chamber. His suggestion was that a very cold air stream that was kept dry by a refrigerated condenser can be circulated in such a system.

The investigations by Dunoyer and Larouse [42] and Woodard [73] indicated furthermore that atmospheric freeze-drying rates of small particles can be equivalent to those of vacuum freeze drying. They provided an economic analysis of the energy costs for moisture removal as well.

A kinetic study of atmospheric pressure freeze-drying was carried out by Boeh-Gcansey [37]. They showed that sublimation time is far shorter in the atmospheric technique than for drying under vacuum for products with large surface area and small thickness. Mumenthaler and Leuenberger [66] investigated an apparatus and a technique for spray-freezing aqueous solutions *in situ* at very low temperatures (-90°C) and for subsequent dehydration of the resulting frozen particles in a stream of cold, dehumidified air.

Most basic studies have confirmed that atmospheric freeze drying and spray freeze drying are feasible processes. If we append a fluidized bed in such a system, improved heat and mass transfer between the drying medium and the frozen powders would be obtained. It also avoids the need to transfer the frozen particles from LN_2 to a freeze dryer. A continuous process can thus be achieved. The system with the combination of spray dryer and fluidized bed, i.e., multistage drying system, was used here as well. Boeh-Gcansey [37] compared two different freeze-drying methods: a vacuum dryer and an atmospheric fluidized-bed dryer containing an adsorbent. Carrot disks are selected as drying samples. Dry activated alumina granules (0.4 mm average diameter) were used as the adsorbent material. He found that the heat transfer coefficient was some 20–40 times greater in the fluidized-bed freeze dryer than in the vacuum dryer.

Lombrana and Villaran [60] carried out an investigation of freeze-drying (lyophilization) by immersion in an adsorbent medium (LIAM) both at atmospheric pressure and under vacuum. They used the spherical particles of cereal food pastes of 2-mm diameter as test samples.

All these studies show that the atmospheric spray and fluidized-bed freeze drying combination is a viable new technology which requires further R&D. Finally, a summary of comparison among four drying operations, i.e., SD, SFD, FD, and ASFBFD, are given in Table 10.10.

TABLE 10.10
Comparison among Four Drying Operations

	SD	FD	SFD	ASFBFD
Drying time	Short	Long	Long	Medium
Powders	Agglomerated or irregular	Cake	Spherical and porous, mono	Spherical and porous
Product quality	Medium	Good	Good	Good
Energy consumption	Low	High	Highest	Medium
Product capacity	High	Medium	Low	Medium
Operation	Continuous	Batch	Batch	Semicontinuous
Invest cost	Low	High	High	High

TABLE 10.11
Optional Measures for Improving Thermal Efficiency in Spray Dryers

Area		Measures
Housekeeping	Heater	<ul style="list-style-type: none"> • Monitor burner efficiency • Check steam leakage or leaking tubes and fouled surfaces in indirect heaters • Improve the steam traps operation
	Air leakage	<ul style="list-style-type: none"> • Check pipe seal after dismantling or maintenance • Check welded points near motion machine, such as centrifugal atomizer, hammer, etc. • Replace worn seals, gaskets, etc.
	Insulation	<ul style="list-style-type: none"> • Replace lagging that has fallen off • Replace water-laden insulation
Process modifications	Inlet air temperature	<ul style="list-style-type: none"> • As high as possible for products
	Outlet air temperature	<ul style="list-style-type: none"> • As low as possible for reaching the allowed residual moisture in products
	Feed concentration	<ul style="list-style-type: none"> • As high as possible if the pump is allowed to transport
	Feed temperature	<ul style="list-style-type: none"> • High temperature is better if the feed quality is not affected
Heat recovery	Multistage	<ul style="list-style-type: none"> • It leads to low outlet temperature and high thermal efficiency
	Direct use of waste heat	<ul style="list-style-type: none"> • Recycle a portion of the exhaust gas to inlet as a portion of drying gas
	Indirect use of waste heat	<ul style="list-style-type: none"> • Use exhaust gas to preheat the inlet air

Source: From Masters, K., Spray drying in practice, *SprayDryConsult International ApS*, Denmark, 2002, 464 p; Mujumdar, A.S., Superheated steam drying, *Handbook of Industrial Drying*, 2nd edition, Ed. A.S. Mujumdar, Marcel Dekker, New York, 1995, pp. 1071–1086; Huang, L.X., Wang, Z., and Tang, J., *Chem. Eng. (China)*, 29(2), 51, 2001.

10.5.5 ENCAPSULATION

Spray drying and fluid-bed drying lead to another popular food application of these technologies, i.e., microencapsulation. Microencapsulation is defined as a process by which one material or a mixture of materials is coated or entrapped within another material or system [69]. This process is commonly used to protect a core material from degradation, to control the release of a core material, or to separate reactive components within a formulation.

In the food industry, because the spray dryer is commonly available, economical, fast, and produces good-quality material [16], it becomes the most common means of encapsulation. The encapsulation process is simple and similar to the one-stage spray drying process. The coated material is called the active or core material, and the coating material is called the shell, wall material, carrier, or encapsulant [43]. The active material to be encapsulated, such as an oil or flavor in an oil base, is dispersed in a hydrocolloid carrier, e.g., gelatin, modified starch, dextrin or maltodextrin, or gum arabic. After the emulsifier is added, the mixture must be homogenized to form an oil-in-water emulsion, and then it is fed to the atomizer for spray drying. In the dryer chamber, the aqueous phase dries and the active material is entrapped as particles within the hydrocolloid or protein film. The active material from the capsule is released under specified conditions. The main controlling factors of

its release are temperature, moisture, and pressure. The microencapsulation technique is not limited to the food domain. It is also used in the pharmaceutical, biotechnology, and chemical industries [36].

10.5.6 ENERGY EFFICIENCY ENHANCEMENT

Spray drying is an energy-intensive process. With energy costs continuing to increase and with changes in overall production levels, spray-dryer users have to look for some ways to improve the thermal efficiency of the spray dryers. For standard single-stage drying, the best way to control energy usage is raising the inlet temperature as high as possible, keeping outlet temperature as low as possible, and taking full advantage of the energy introduced. However, the disadvantage of this method is potential degradation or discoloration of the product in food spray-drying applications.

There are some other optional measures to improve it, such as good housekeeping measures, process modifications, and heat recovery options [54]. We summarize them in Table 10.11. See Mujumdar [64] for more ideas used in drying.

10.6 CLOSING REMARKS

Spray drying process is one of the most common methods adopted in many industries. In this chapter, a summary is provided on the fundamentals of spray

drying, the selection of spray dryers as well as usages of spray dryers in the industry. Spray dryers, both conventional and innovative, will continue to find increasing applications in various industries, but almost all industries need or use or produce powders starting from liquid feedstocks. Therefore, although it is very difficult to generate rules for the selection of spray dryers in different areas because of numerous possible exceptions and new developments, it is important for the users of spray dryers to understand the typical and main characteristics of spray drying process.

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11 Freeze Drying

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11.1 INTRODUCTION

Certain biological materials, pharmaceuticals, and foodstuffs, which may not be heated even to moderate temperatures in ordinary drying, may be freeze-dried. The substance to be dried is usually frozen. In freeze drying, the water or another solvent is removed as a vapor by sublimation from the frozen material in a vacuum chamber. After the solvent sublimates to a

vapor, it is removed from the drying chamber where the drying process occurs.

As a rule, freeze drying produces the highest quality food product obtainable by any drying method. A prominent factor is the structural rigidity afforded by the frozen substance at the surface where sublimation occurs. This rigidity to a large extent prevents collapse of the solid matrix remaining after drying. The result is a porous, nonshrunken structure in the

dried product that facilitates rapid and almost complete rehydration when water is added to the substance at a later time.

Freeze drying of food and biological materials also has the advantage of little loss of flavor and aroma. The low processing temperatures, the relative absence of liquid water, and the rapid transition of any local region of the material dried from a fully hydrated to a nearly completely dehydrated state minimize the degradative reactions that normally occur in ordinary drying processes, such as nonenzymatic browning, protein denaturation, and enzymatic reactions. In any food material, some nonfrozen water, which is called *bound* or *sorbed water*, will almost unavoidably be present during freeze drying, but there is very often a rather sharp transition temperature for the still wet region during drying [1], below which the product quality improves markedly. This improvement shows that sufficient water is frozen to give the beneficial product characteristics of freeze drying.

However, freeze drying is an expensive form of dehydration for foods because of the slow drying rate and the use of vacuum. The cost of processing is offset to some extent by the absence of any need for refrigerated handling and storage.

Increasingly, freeze drying is used for dehydrating foods otherwise difficult to dry, such as coffee, onions, soups, and certain seafoods and fruits. Freeze drying is also increasingly employed in the drying of pharmaceutical products. Many pharmaceutical products when they are in solution deactivate over a period of time; such pharmaceuticals can preserve their bioactivity by lyophilization soon after their production so that their molecules are stabilized.

Systematic freeze drying is a procedure mainly applied to the following categories of material [1–89]:

1. Nonliving matter, such as blood plasma, serum, hormone solutions, foodstuffs, pharmaceuticals (e.g., antibiotics), ceramics, superconducting materials, and materials of historical documents (e.g., archaeological wood)
2. Surgical transplants, which are made nonviable so that the host cells can grow on them as the skeleton, including arteries, bone, and skin
3. Living cells destined to remain viable for longer periods of time, such as bacteria, yeasts, and viruses

Freeze drying requires very low pressures or high vacuum to produce a satisfactory drying rate. If the water was in a pure state, freeze drying at or near 0°C at an absolute pressure of 4.58 mmHg could be performed. But, since the water usually exists in a

combined state or a solution, the material must be cooled below 0°C to keep the water in the solid phase. Most freeze drying is done at –10°C or lower at absolute pressures of about 2 mmHg or less.

In short, freeze drying is a multiple operation in which the material to be stabilized is

1. Frozen hard by low-temperature cooling
2. Dried by direct sublimation of the frozen solvent and by desorption of the sorbed or bound solvent (nonfrozen solvent), generally under reduced pressure
3. Stored in the dry state under controlled conditions (free of oxygen and water vapor and usually in airtight, opaque containers filled with inert dry gas)

If correctly processed, most products can be kept in such a way for an almost unlimited period of time while retaining all their initial physical, chemical, biological, and organoleptic properties, and remaining products available at any time for immediate reconstitution. In most cases this is done by the addition of the exact amount of solvent that has been extracted, thus giving to the reconstituted product a structure and appearance as close as possible to the original material. However, in some instances, reconstitution can be monitored to yield more concentrated or diluted products by controlling the amount of solvent.

Vaccines and pharmaceutical materials are very often reconstituted in physiological solutions quite different from the original but best suited for intramuscular or intravenous injections. Freeze-dried organisms, such as marine animals, plants, or tissue extracts, can also be the starting point of an extraction process [5] using nonaqueous solvents with the purpose of isolating bioactive substances. Freeze drying allows dehydration of the systems without impairing their physiological activity so that they can be prepared for appropriate organic processing.

Another example is the freeze drying of nuclear wastes, which results in the manufacture of dry powders of medium radioactivity. Mixed with appropriate chemicals, they can be fused into glass bricks or molded to provide low-cost, high-energy radiation sources.

The freeze drying method has also been used in the synthesis of superconducting materials, and produces homogeneous, submicron superconductor powders of high purity [4].

In the chemical industry, catalyzers, adsorbing filters, and expanded plastics can be used in the dry form and placed in the path of appropriate fluids or gases. Freeze-dried dyes may also be dispersed in other media, such as oils and plastics.

These examples are not exhaustive; detailed presentations on the uses of the freeze drying process and of freeze-dried products are given in Refs. [1–6,8,14,15,63,72,84].

11.2 FREEZE DRYING PROCESS

Freeze drying is a process by which a solvent (usually water) is removed from a frozen foodstuff or a frozen solution by sublimation of the solvent and by desorption of the sorbed solvent (nonfrozen solvent), generally under reduced pressure. The freeze drying separation method (process) involves the following three stages: (a) the freezing stage, (b) the primary drying stage, and (c) the secondary drying stage.

In the *freezing stage*, the foodstuff or solution to be processed is cooled down to a temperature at which all the material is in a frozen state.

In the *primary drying stage*, the frozen solvent is removed by sublimation; this requires that the pressure of the system (freeze dryer) at which the product is dried must be less than or near to the equilibrium vapor pressure of the frozen solvent. If, for instance, frozen pure water (ice) is processed, then sublimation of pure water at or near 0°C and at an absolute pressure of 4.58 mmHg could occur. But, since the water usually exists in a combined state (e.g., foodstuff) or a solution (e.g., pharmaceutical product), the material must be cooled below 0°C to keep the water in the frozen state. For this reason, during the primary drying stage, the temperature of the frozen layer (see Figure 11.1) is most often at -10°C or lower at absolute pressures of about 2 mmHg or less. As the solvent (ice) sublimates, the sublimation interface (plane of sublimation), which started at the outside surface (see Figure 11.1), recedes, and a porous shell of dried material remains. The heat for the latent heat

of sublimation (2840 kJ/kg ice) can be conducted through the layer of dried material and through the frozen layer, as shown in Figure 11.1. The vaporized solvent (water) vapor is transported through the porous layer of dried material. During the primary drying stage, some of the sorbed water (nonfrozen water) in the dried layer may be desorbed. The desorption process in the dried layer could affect the amount of heat that arrives at the sublimation interface and therefore it could affect the velocity of the moving sublimation front (interface). The time at which there is no more frozen layer (that is, there is no more sublimation interface) is taken to represent the end of the primary drying stage.

The *secondary drying stage* involves the removal of solvent (water) that did not freeze (this is termed sorbed or bound water). The secondary drying stage starts at the end of the primary drying stage, and the desorbed water vapor is transported through the pores of the material that is dried.

11.2.1 FREEZING STAGE

The freezing stage represents the first separation step in the freeze drying process, and the performance of the overall freeze drying process depends significantly on this stage [64,65]. The material system to be processed (e.g., gel suspension, liquid solution, or foodstuff) is cooled down to a temperature (this temperature depends on the nature of the product) that is always below the solidification temperature of the material system. For instance, if the material to be freeze-dried is a solution with an equilibrium phase diagram that presents a eutectic point (e.g., the solution of NaCl and water presents a eutectic point at -21.6°C), then the value of the final freezing temperature must be below the value of the eutectic temperature; in this case, the material becomes wholly crystalline.

In practice, materials display one of two different types of freezing behavior: (a) the liquid phase suddenly solidifies (eutectic formation) at a temperature that depends on the nature of solids in the sample, or (b) the liquid phase does not solidify (glass formation), but rather it just becomes more and more viscous until it finally takes the form of a very stiff, highly viscous liquid. In case (b), there is no such thing as a eutectic temperature, but a minimum freezing temperature.

At the end of the freezing step there already exists a separation between the water to be removed (frozen water in the form of ice crystals) and the solute. In many cases, at the end of the freezing stage about 65–90% of the initial (at the start of the freezing stage) water is in the frozen state and the remaining

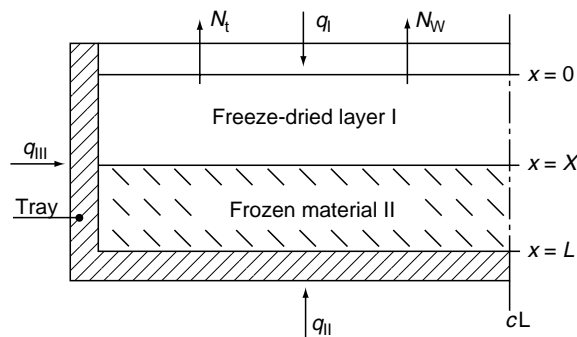


FIGURE 11.1 Diagram of a material on a tray during freeze drying. The variable X denotes the position of the sublimation interface (front) between the freeze-dried layer (layer I) and the frozen material (layer II).

10–35% of the initial water is in the sorbed (nonfrozen) state. The shape of the pores, the pore size distribution, and pore connectivity [6,9,11,16–18,64–72] of the porous network of the dried layer formed by the sublimation of the frozen water during the primary drying stage depend on the ice crystals that formed during the freezing stage; this dependence is of extreme importance because the parameters that characterize the mass and heat transfer rates in the dried layer are influenced significantly by the porous structure of the dried layer. If the ice crystals are small and discontinuous, then the mass transfer rate of the water vapor in the dried layer could be limited. On the other hand, if large dendritic ice crystals are formed and homogeneous dispersion of the pre- and posteutectic frozen solution can be realized, the mass transfer rate of the water vapor in the dried layer could be high and the product could be dried more quickly. Thus, the method and rate of freezing, as well as the shape of the container of the solution and the nature of the product, are critical to the course of lyophilization because they affect the drying rate and the quality of the product.

In industrial freeze dryers, the freezing of the product is usually made in the same plant where the drying also occurs. In the vacuum-spray freeze dryer, the solvent evaporation aut freezes the small particles of product (evaporative freezing), and the freezing stage begins at the same time as the drying stage. In certain food freeze dryers, the freezing of the product is also accomplished by spraying liquid nitrogen in the drying chamber in which the product is placed. In tray and pharmaceutical freeze dryers, the freezing stage is realized by contact between cooled plates and product-supporting containers. The exergy analysis [65] of the freezing stage of the freeze drying process indicates that very substantial reductions in the magnitudes of the total exergy loss and of the exergy input because of the heat that must be removed during the freezing stage, can be obtained when the freezing stage is operated by a rational distribution in the magnitude of the temperature of the cooling source. The rational distribution in the magnitude of the temperature of the cooling source should provide significant savings in the utilization of energy during the freezing stage of the freeze drying process as well as satisfactory freezing rates that form ice crystals that are continuous and highly connected and their shape and size are such that the pores of the porous matrix of the dried layer generated by sublimation during the primary drying stage, have a pore size distribution, pore shape, and pore connectivity [64–69] that are appropriate to allow high rates for mass and heat transfer during the primary and secondary drying stages of the lyophilization process.

11.2.2 PRIMARY DRYING STAGE

After the freezing stage, the drying chamber where the product is placed is evacuated and the chamber pressure is reduced to a value that would allow the sublimation of solvent (water) to take place in the primary drying stage. When the water molecules sublime and enter the vapor phase, they also keep with them a significant amount of the latent heat of sublimation (2840 kJ/kg ice) and thus the temperature of the frozen product is again reduced. If there is no heat supplied to the product by a heat source, then the vapor pressure of the water at the temperature of the product reaches the same value as that of the partial pressure of the water vapor in the drying chamber; therefore, the system reaches equilibrium and no additional water sublimation from the product would occur. Thus, in order to have continuous sublimation of water from the product, the latent heat of sublimation must be provided to the material from a heat source. The heat is supplied to the product usually by conduction, convection, or radiation; conduction is realized by contact between heated plates and product-supporting containers.

The amount of heat that can be supplied to the product cannot be increased freely because there are certain limiting conditions that have to be satisfied during the primary drying stage. One of the constraints has to do with the maximum temperature that the dried product (freeze-dried layer in Figure 11.1) could tolerate without (a) loss of bioactivity, (b) color change, (c) the possibility for degradative chemical and biochemical reactions to occur, and (d) structural deformation in the dried layer [67]. The maximum temperature that the dried product could tolerate without suffering any of the above-mentioned deleterious effects is denoted, for a given product, by T_{scor} (T_{scor} is often called, by convention, the temperature of the scorch point of the dried product).

Another constraint has to do with the maximum temperature the frozen layer could tolerate so that it remains a frozen layer. If the material has a eutectic form and if the temperature of the lowest eutectic is exceeded during the primary drying stage, then melting in the frozen layer (Figure 11.1) can occur. The melting at the sublimation interface, or any melting that would occur in the frozen layer, can cause gross material faults such as puffing, shrinking, and structural topologies filled with liquid solution. When melting has occurred at some point in the frozen layer, then the solvent at that point cannot be removed by sublimation. Therefore, there is process failure in the drying of the frozen material because the frozen solvent (water) cannot be removed any more from the frozen layer (Figure 11.1) only by

sublimation, and there has also been, at the least, loss in structural stability.

If the material has a glass form and if the minimum freezing temperature is exceeded during the primary drying stage, then the phenomenon of collapse can occur; this makes the product collapse with a loss of rigidity in the solid matrix. Again in this case, there is process failure in the drying of the frozen material because the water cannot be removed any more from the frozen layer only by sublimation, and there has also been at least a loss in structural stability.

The structural stability of a material relates to its ability to go through the freeze drying process without change in size, porous structure, and shape. The maximum allowable temperature in the frozen layer is determined by both structural stability and product stability (e.g., product bioactivity) factors; that is, the maximum value of the temperature in the frozen layer during the primary drying stage must be such that the drying process is conducted without loss of product property (e.g., bioactivity) and structural stability.

Sometimes the product stability factors are related to structural stability factors (as in melting). There are systems in which the product stability factors do not depend on structural stability factors, as is the case for many vaccines, viruses, and bacteria, for which the temperature of the frozen layer during the primary drying stage must be kept well below the melting temperature so that there is a good level of bioactivity and organism survival after drying.

In general, product stability is related to the temperature of the frozen layer during the primary drying stage. The maximum allowable temperature that the frozen layer could tolerate without suffering melting, puffing, shrinking, collapse, and loss of product property or stability is denoted, for a given product, by T_m . (T_m is often called, by convention, the melting temperature of the sublimation interface of the frozen layer.)

The water vapor produced by the sublimation of the frozen water in the frozen layer and by the desorption of sorbed (nonfrozen) water in the dried layer during the primary drying stage travels by diffusion and convective flow through the porous structure of the dried layer and enters the drying chamber of the freeze dryer. (It should be noted that most of the water removed during the primary drying stage is produced by sublimation of the frozen water in the frozen layer.) The water vapor must be continuously removed from the drying chamber in order to maintain nonequilibrium conditions for the drying process in the system. This is usually accomplished by fitting a refrigerated trap (called an ice condenser) between the drying chamber and the vacuum pump; the water vapor is collected on the cooled surface of the

condenser in the form of ice. The time at which there is no more frozen layer is taken to represent the end of the primary drying stage.

11.2.3 SECONDARY DRYING STAGE

The secondary drying stage involves the removal of water that did not freeze (sorbed or bound water). In an ideal freeze drying process, the secondary drying stage starts at the end of the primary drying stage. The word ideal is used here to suggest that in an ideal freeze drying process only frozen water should be removed during the primary drying stage, whereas the sorbed water should be removed during the secondary drying stage. But, as we discussed above, in real freeze drying systems a small amount of sorbed water could be removed by desorption from the dried layer of the product during the primary drying stage and thus there could be some secondary drying occurring in the dried layer of the product during the primary drying stage.

In real freeze drying processes, the secondary drying stage is considered to start when all the ice has been removed by sublimation (end of primary drying stage). It is then considered that during the secondary drying stage most of the water that did not freeze (bound water) is removed. The bound moisture is present due to mechanisms of (a) physical adsorption, (b) chemical adsorption, and (c) water of crystallization. Whereas the amount of bound water is about 10–35% of the total moisture content (65–90% of the total moisture could be free water that was frozen and then removed by sublimation during the primary drying stage), its effect on the drying rate and overall drying time is very significant. The time that it takes to remove the sorbed water could be as long or longer than the time that is required for the removal of the free water.

The bound water is removed by heating the product under vacuum. But, as in the case of primary drying, the amount of heat that can be supplied to the product cannot be increased freely because there are certain constraints that have to be satisfied during the secondary drying stage. The constraints have to do with the moisture content and the temperature of the product; these two variables influence the structural stability as well as the product stability during and after drying.

For structural stability, the same phenomena, as in the case of the primary drying stage, have to be considered: collapse, melting (if temperature is increased at constant moisture), or dissolution (if moisture is increased at constant temperature) of the solid matrix can occur. Product stability (e.g., bioactivity) is a function of both moisture content and temperature

in the sample, and during secondary drying the moisture concentration and temperature in the sample could vary widely with location and time. This implies that the potential for product alteration to occur in the sample will vary with time and location. The moisture concentration profile is related to the temperature profile in the dried layer; thus, the moisture content in the sample cannot be controlled independently. Since many products are temperature-sensitive, it is usual to control product stability by limiting the value of the temperature during the secondary drying process and then the final moisture content is checked before the end of the cycle [6,19,63,64,67,72–80].

In the secondary drying stage, the bound water is removed by heating the product under vacuum; the heat is supplied to the product usually by conduction, convection, or radiation. The following product temperatures are usually employed: (a) between 10 and 35°C for heat-sensitive products and (b) 50°C or more for less-heat-sensitive products.

The residual moisture content in the dried material at the end of the secondary drying stage, as well as the temperature at which the dried material is kept in storage, are critical factors in determining product stability during its storage life. Some vaccines can remain stable for many years when they are stored at –20°C, whereas a significant loss of titer can be found after 1 y if they are stored at 37°C [20]. Furthermore, certain vaccines such as live rubella and measles can be damaged by overdrying (final moisture content of about 2% is required for best titer retention), whereas other materials such as chemotherapeutics and antibiotics must be dried to a residual moisture content as low as 0.1% for best results.

11.3 MICROWAVE FREEZE DRYING

The limitations on heat transfer rates in conventionally conducted freeze drying operations have led early to the attempt to provide internal heat generation with the use of microwave power [21,22]. Theoretically, the use of microwaves should result in a very accelerated rate of drying because the heat transfer does not require internal temperature gradients and the temperature of ice could be maintained close to the maximum permissible temperature for the frozen layer without the need for excessive surface temperatures.

If, for instance, it is permissible to maintain the frozen layer at –12°C, then it has been estimated [5] that the drying time for an ideal process using microwaves for a hypothetical 1-in. slab would be 1.37 h. It should be noted that this drying time compares very favorably with the 8.75 h required for the case of heat input through the dry layer, 13.5 h for heat input

through the frozen layer without dry layer removal, and even with the relatively short drying time of 4 h for the case in which the dry layer was continuously removed. In laboratory tests on freeze drying of a 1-in.-thick slab of beef, an actual drying time of slightly over 2 h was achieved, compared with about 15 h for conventionally dried slabs [23].

In spite of these apparent advantages, the application of microwaves to industrial freeze drying has not been successful [5,24,72,86,87]. The major reasons for the failures are the following:

1. Energy supplied in the form of microwaves is very expensive. It was estimated that it may cost 10 to 20 times more to supply 1 Btu from microwaves than it does from steam [24].
2. A major problem in the application of microwaves is the tendency to glow discharge, which can cause ionization of gases in the chamber and deleterious changes in the food, as well as loss of useful power. The tendency to glow discharge is greater in the pressure range of 0.1–5 mmHg and can be minimized by operating the freeze dryers at pressures below 50 μ m. Operation at these low pressures, however has a double drawback: (a) it is quite expensive, primarily because of the need for condensers operating at a very low temperature and (b) the drying rate at these low pressures is much slower.
3. Microwave freeze drying is a process that is very difficult to control. Since water has an inherently higher dielectric loss factor than ice, any localized melting produces a rapid chain reaction, which results in runaway overheating.
4. Economical microwave equipment suitable for the requirements of industrial freeze drying of foods and pharmaceuticals on a large continuous scale is not yet available.

In view of all of these limitations, microwave freeze drying is at present only a potential development [25] and is not considered in the following sections of this chapter.

11.4 FREEZE DRYING PLANTS AND EQUIPMENT

In the freeze drying plant, three process sections are especially energy consuming. Process section 1 involves the freezing of the wet product. As this is normally considered one of the preparatory steps before the freeze drying proper, we will concentrate on the other two that take place in the freeze drying

cabinet [1,8]. Process section 2 involves the controlled supply of heat to the product to cover requirements for the sublimation and desorption processes (primary and secondary drying stages). Process section 3 involves the removal from the freeze drying chamber of the vast volumes of water vapor released during the sublimation and desorption processes. Of these three process sections, removal of the water vapor always consumes the largest amount of energy. The efficiency of water vapor removal, the vapor trap system, therefore has a decisive effect on the total energy consumption of the freeze drying plant.

The vapor trap is placed in a chamber communicating with the freeze-drying cabinet. The water vapor condenses to ice on its refrigerated surfaces. When in operation the efficiency of the vapor trap is shown by a small total temperature difference ΔT between the saturation temperature for water vapor at the pressure in the freeze-drying cabinet and the evaporation temperature of the refrigerant (Figure 11.2). This total temperature difference ΔT results mainly from each of the following three resistances:

1. Pressure difference ΔP equivalent to the pressure drop caused by the resistances to the vapor flow from the freeze-drying cabinet to the cold surfaces of the vapor trap.

2. The temperature difference ΔT_{ice} with the layer of ice on the cold surface.
3. The temperature difference ΔT_{refr} between the cold surface and the evaporating refrigerant. For an efficient vapor trap it is necessary to have a combination of a large cross-sectional area for the vapor flow (low ΔP), an efficient deicing system (low ΔT_{ice}), and an efficient refrigerating system (low ΔT_{refr}).

A less efficient vapor trap means a higher ΔT , thus demanding a lower evaporation temperature of the refrigerating plant to maintain the required vacuum in the freeze-drying cabinet. Lower evaporation temperature means higher operation costs. In this temperature range, an evaporation temperature 10°C lower means 50% increased energy consumption.

When evaluating industrial freeze drying plants, the following characteristics are of prime importance:

1. Operation reliability
2. Ease and quality of process control
3. Product losses
4. Vapor trap efficiency

In [Section 11.4.1](#) and [Section 11.4.2](#), some of the commonly used types of pilot and industrial freeze

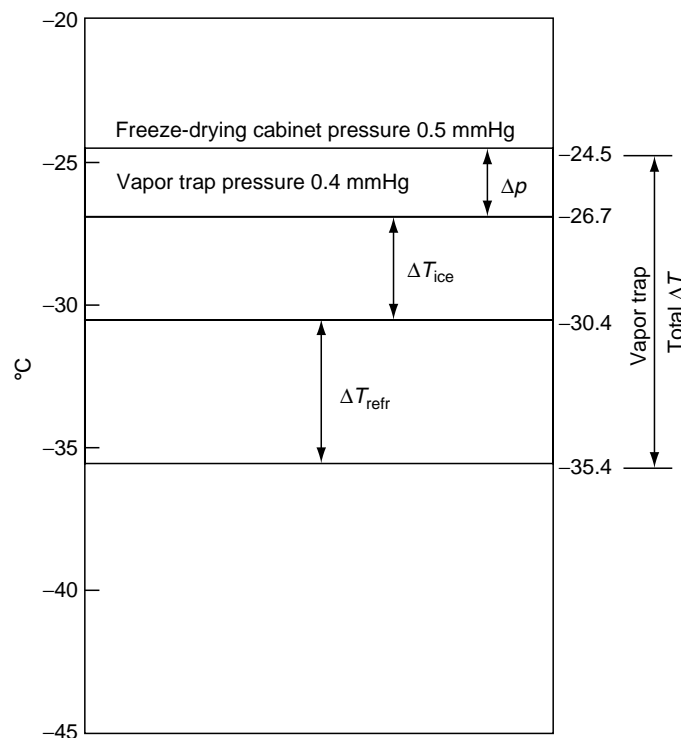


FIGURE 11.2 Graphic presentation of the variables ΔP , ΔT_{ice} , ΔT_{refr} , and ΔT .

drying plants are presented and their most important technical features are discussed. The sections of the plants where the product pretreatment inclusive of the freezing operation takes place are very different from one plant to another, depending on the type of products handled [1,5]. In Section 11.4.3, however, we concentrate on the freeze drying installation proper, that is, the equipment in which sublimation of frozen solvent and desorption of bound solvent takes place. When we later refer to freeze drying plants, it will be in this more restricted meaning of the word.

11.4.1 PILOT FREEZE DRYERS

Freeze drying pilot units appropriate for use in the pharmaceutical and food industries, as well as in the laboratory, are in high demand because they are used to explore possibilities for the preservation of labile products, especially with those of biological origin. These units are portable and are of convenient size for developmental work on freeze-dried products in laboratories and factories around the world. A large number of designs incorporate self-contained facilities for refrigeration, heating, and vacuum pumping, and they can freeze-dry batches consisting of from 2 to 20 kg of frozen product. Because of the large variety of pilot freeze dryers that are employed in industries and laboratories and because of the limitation of space to

describe all of them, a pilot freeze dryer is described here with characteristics that are very close to the characteristics of the industrial large-scale lyophilizers.

A schematic diagram of the pilot unit (Crioforma model C5-2) is shown in Figure 11.3a. The unit consists of (a) a freezing fluid system (R13B1) that can be sent to the heat exchanger in the section of the condenser or into the refrigeration coils for product freezing, (b) a heating circuit (silicon oil is the heating fluid) for plate heating and defrosting of the condenser, and (c) a vacuum system for evacuating air from the apparatus before and during drying.

The rectangular drying chamber shown in Figure 11.3b is mounted on top of the section of the condenser and the dimensions are 0.4 m \times 0.4 m with 0.6 m of depth. Viewing windows are incorporated in the sections of drying and condensation. The refrigeration and vacuum systems are in the internal part of the apparatus with complete dimensions of 1.1 m \times 0.8 m \times 1.8 m (the dimension 1.8 m represents the height of the apparatus).

For its pilot use, the freeze dryer offers full control of the process variables and is able to achieve conditions of pressure and temperature beyond the limits of production units. The shelf and ice condenser temperatures of the pilot unit can be -50 and -70°C , respectively, and the pressure in the drying chamber can be as low as 1 Pa or less. The pilot freeze dryer has a

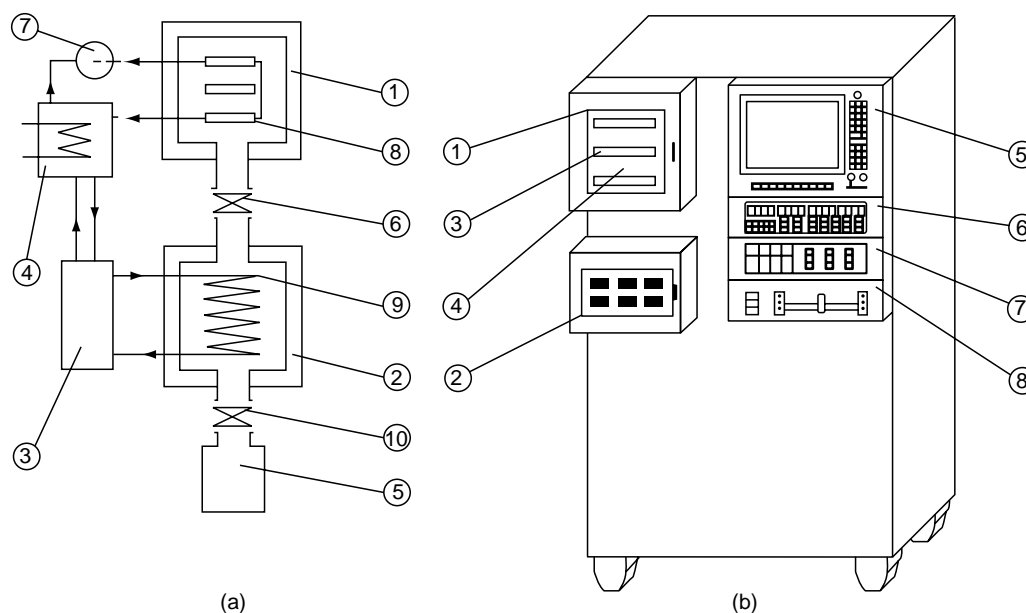


FIGURE 11.3 Pilot freeze dryer: (a) diagram of Crioforma model C5-2; 1, drying chamber; 2, ice condenser chamber; 3, refrigeration unit; 4, cooling and heating system for the plates; 5, vacuum unit; 6, isolation butterfly valve; 7, silicon oil pump; 8, cooling and heating plate; 9, refrigerated coil; 10, condenser vacuum valve; (b) frontal view of model C5-2; 1, drying chamber; 2, ice condenser chamber; 3, cooling and heating plate; 4, inspection window; 5, computer system; 6, vacuum indicator and regulator; 7, temperature control panel; 8, printer. (Model C5-2 courtesy Crioforma.)

control panel fully accessorized with instruments that record and display (a) the temperature inside the product, (b) the temperature on the plates, (c) the temperature of the coils of the condenser, (d) the pressure in the drying chamber, (e) the pressure in the vacuum unit, and (f) the pressure in the section of the condenser. The use of a personal computer, in this pilot unit, with programmable temperature during the freeze drying cycle and programmable input–output logic in the different freeze drying stages, offers a wide variety of drying cycles, as well as the capability for the acquisition of many data, so that process optimization could be examined and studied without the risk and cost of investigating the freeze drying system of interest in a large-production freeze dryer.

11.4.2 INDUSTRIAL FREEZE DRYERS

11.4.2.1 Tray and Pharmaceutical Freeze Dryers

By far the largest number of the industrial freeze dryers in operation is of the vacuum batch type with freeze drying of the product in trays. There are two main types, depending on the type of condenser used. In the first type, the condenser plates are alongside the tray-heater assembly and in the same chamber; in the second type the condenser is in a separate chamber joined to the first by a wide, in general, butterfly valve. This latter type of plant is always used in pharmaceutical industries, but it can also be used for the freeze drying of foods. Because of the wide variety and complexity of the problems associated with the production of pharmaceuticals by freeze drying, in the following paragraphs the principal features of an industrial tray freeze dryer for pharmaceuticals are presented.

The principal problem in the freeze drying of pharmaceutical solutions is to operate in sterile conditions. The location of the plant must be able to warrant a sterile condition during the filling, charging before drying, and discharging after drying of the pharmaceutical product. This is realized by facing the drying chamber door in a wall separating the sterile room from the machine or nonsterile room.

In the plant, this separation is accomplished with an isolation valve that separates the ice condenser from the drying chamber; this valve is also able to permit (a) the pressure rise test at the end of the freeze drying cycle, (b) the simultaneous discharging and loading of the product and condenser defrosting, and (c) the reduction of cross-contamination between batches to a minimum. All the internal parts of the freeze dryer are of stainless steel type AISI 304L or 316L with a finished surface of 300 mesh or more. In the modern plants, the internal sterilization of the

equipment is usually made with pressurized steam at 121°C or more; in old plants, sterilization is realized with the use of certain proprietary sanitizing agents.

The product containers (vials or bottles loaded on stainless steel trays) are usually sterilized in a separate unit before the filling and charging in the freeze dryer. These operations require the presence of people in the sterile room with consequent handling of the containers and possible contamination of the batch. For this reason, the human presence in the sterile room is usually reduced to a number of people that are strictly necessary.

For this purpose, a new freeze dryer plant concept has been developed to reduce the risk of product contamination. The plant, as shown in [Figure 11.4](#) (Crioforma model C300-7) has two doors: a small door for loading the product before drying and a full door (located in a position opposite to the small door) for discharging the product after drying. The condenser is placed on the ground floor, which is below the first floor where the drying chamber is located. The shelves of the freeze dryer are lowered to the bottom of the drying chamber and are then lifted one by one to a position in line with the loading machine. The charging of the product is made under laminar flow of sterile air; the small door is opened only for each plate loading and is then immediately closed.

If the product is unstable and must be frozen within a short time after it is filled into its container, then it is possible to load trays of product onto the precooled shelves a half plate at a time. When the product container is a bottle as shown in [Figure 11.4](#), it usually has on the top a silicon plug that is partially introduced into the bottle; the solvent vapor leaves the container from the free space between the inserted portion of the plug and the container. After drying and before product discharge, the bottles are stoppered in the drying chamber with the plugs that are now fully introduced into the bottles. The stoppering operation is done (a) in vacuum conditions or (b) at atmospheric pressure by breaking the drying chamber vacuum with sterile nitrogen, which prevents successive oxidation of the product; case (b) is most often employed in practice. The silicon plug in the stoppered bottles provides a protection from contamination and it may be possible to discharge the product in a less sterile environment from the full door of the freeze dryer in only one operation. The entire process may be fully automated as the bottles are removed from the filling machine; the disadvantage of the automation is that the loading time of the freeze dryer may become as long as the time it usually takes to complete the filling step of the operation; this could reduce the theoretical freeze dryer production for a large installation.

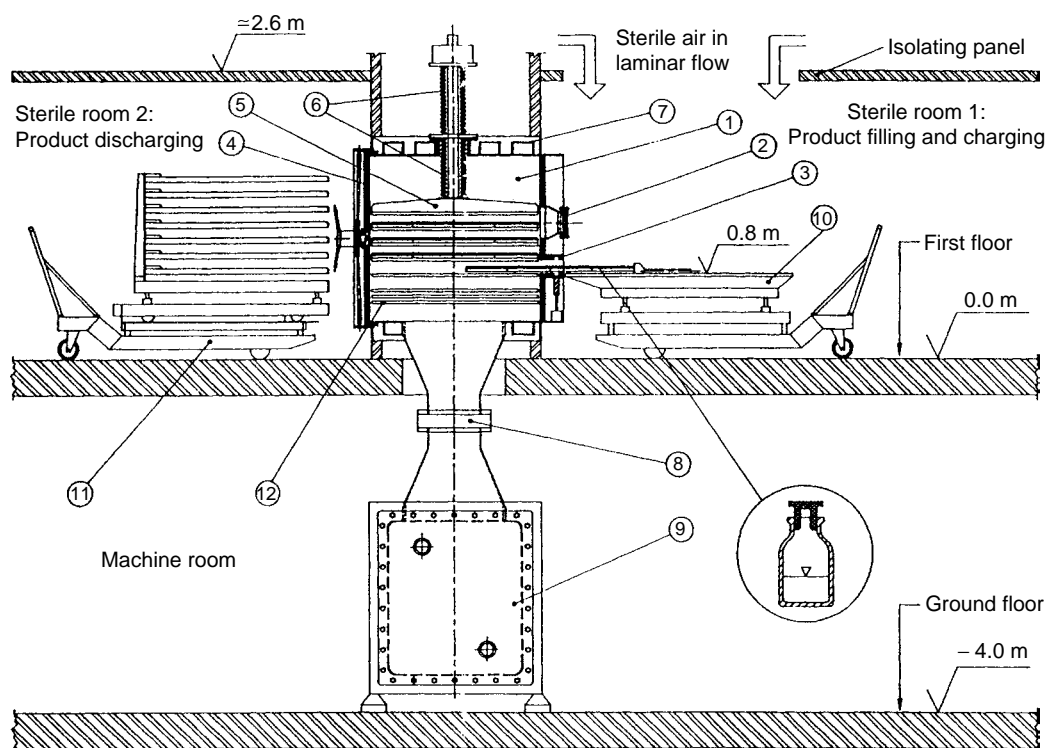


FIGURE 11.4 Layout of an industrial freeze dryer with stoppering device (Criofarma model C300-7): 1, drying chamber; 2, inspection window; 3, automatic small door opening; 4, full door; 5, hydraulic press for stoppering the bottles after drying; 6, PTFE elbows for double sterile condition inside the stoppering plug; 7, reinforcing member and cooling coils after steam sterilization; 8, isolation butterfly valve; 9, ice condenser chamber; 10, loading device; 11, discharging device; 12, unloaded shelves. (Model C300-7 courtesy Criofarma.)

Pharmaceutical freeze dryers are very often used to produce raw materials like ampicillin, cloxacillin, and cefazolin (usually as sodium salt), or other specialty materials like collagen. In these systems, the product is usually charged on stainless steel or polyethylene film trays and the plant is usually a medium or a large unit with a loading surface varying from 15 to 60 m².

If the product to be freeze-dried is not particularly unstable (e.g., collagen) and can withstand a delay of some hours between filling into its tray and freezing, then one can usually accumulate the trays of product on a loading trolley. When the loading trolley is filled, it is placed in front of the freeze dryer and the trays are automatically pushed on the shelves without sliding contact (in order to avoid particle generation) in only one operation. This system is advantageous because it permits maximum utilization of the freeze dryer; the trolley may be loaded ahead of time when the freeze dryer is available for unloading and loading so that the loading operation can be carried out in a few minutes.

If the product is not stable in the liquid state (e.g., ampicillin sodium salt) and must be frozen within a

short time after its preparation, it is common to charge the empty trays on the precooled shelves and then to fill the trays so that the freezing step is very quick and can proceed during the whole loading operation. This approach is also advantageous because it reduces the freeze drying cycle time; this happens because the cooling phase starts at the same time as the loading phase, with a consequent reduction in the total time of these two steps.

If the product is directly charged on the trays (bulk production) of the freeze drying equipment, it is found to be convenient to have an additional small ice condenser or so-called auxiliary ice condenser that is also connected with the drying chamber, together with the principal ice condenser. A typical sketch of this device (Criofarma model C1200-20) in a plant of 60 m² of loading surface is shown in Figure 11.5.

With this device, the plant is working with the principal ice condenser for the removal of free water (frozen water) during the primary drying stage (65–90% of the total moisture content is free water), whereas the plant is working with the unloaded auxiliary ice condenser for the removal of bound water

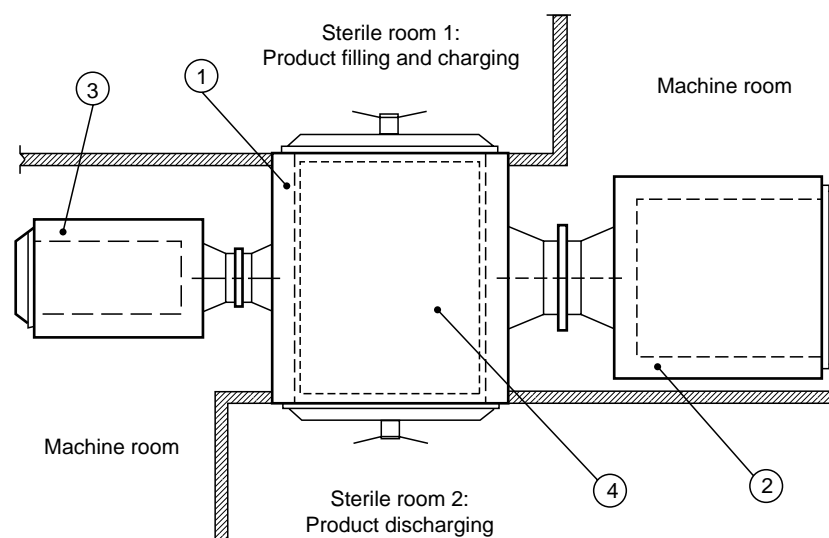


FIGURE 11.5 Top view of an industrial freeze dryer of 60-m² shelf area with auxiliary condenser (Crioforma model C1200-20): 1, drying chamber with double full doors; 2, principal ice condenser chamber; 3, auxiliary ice condenser chamber; 4, product support shelf. (Model C1200-20 courtesy Crioforma.)

(10–35% of the total moisture is bound water) during the secondary drying stage. The time required for the removal of bound water is usually at least as long as the time required for the removal of free water, and for this reason the auxiliary ice condenser has usually a small independent vacuum and refrigeration system with an installed power of one fourth to one sixth of the total refrigeration and vacuum-installed power of the plant. The advantages of this device are: (a) the possibility of the principal ice condenser defrosting before the end of the drying cycle, (b) energy savings that can result in the reduction of the freeze drying running cost, and (c) better performance for the overall drying cycle.

When more freeze dryers than one are used in the production of raw material, then the vacuum line of each plant is connected with the vacuum lines of the other plants by a set of exclusion valves. Thus, if a failure occurs in the vacuum system of one plant, this same plant can end its drying cycle without stopping its operation by using the pumping suction of another plant. A similar device may also be used for the refrigeration units as sometimes one of them is always in a standby condition. The concept that the same vacuum and refrigeration system may operate in different plants is similar to the utilization of industrial multibatch freeze dryers where the simultaneous production of different products is made possible in a freeze drying plant built with a number of batch cabinets programmed to operate with overlapping drying cycles but served by the same central system for (a) tray heating, (b) condenser refrigeration, and

(c) vacuum pumping; for each cabinet, the process is individually controlled from a separate control panel.

A user of a pharmaceutical freeze dryer must observe good manufacturing practice (GMP) for processes and equipment to be validated before and during use. These can be divided into three grouped requirements about different parts and functions of the same plant as follows: (a) plant design and equipment materials, (b) control hardware and software validation, and (c) calibration of instruments. The plant design and the materials of the equipment must be such that they eliminate the potential of dirt traps and ensure successful sterilization (usually with clean pressurized steam at 121°C or more; in older plants they use proprietary sanitizing agents). Also, good cleaning access must be provided, sometimes with a clean-in-place (CIP) system (cleaning the inside part of the plant with sterile water sprayed at high pressure from internal nozzles).

The validation of control hardware and software basically requires the suitability of computer hardware assigned for the task, and that computer programs perform consistently within preestablished operational limits so that analysis of the effects of possible failures can be carried out. The calibration of instruments requires that the supplier of a freeze dryer provides a work certificate of calibration and that the user periodically verifies the performance of the instruments with an external authorized and certified instrument. In Refs. [26–33], useful information for GMP compliance, process, and computer system validation can be found.

11.4.2.2 Multibatch Freeze Dryers

The freeze drying process in a batch plant is normally program controlled to minimize the drying time and to maximize the production of the plant. With a single-batch plant the load on the various systems will be very variable throughout the drying cycle. The material flow and the product handling operations will also be discontinuous because of the batch process characteristic. This means that optimal utilization of resources will not be possible in a single-cabinet batch plant.

To a great extent this disadvantage can be eliminated when an industrial freeze drying plant is built with a number of batch cabinets programmed to operate with staggered, overlapping drying cycles. Each of the cabinets can be charged with products from the same system, and they are served by the same central system for tray heating, for condenser refrigeration, and for vacuum pumping. But, the process is individually controlled for each cabinet from a separate control panel. This makes possible the simultaneous production of different products, which increases the operation flexibility of the plant. With only two cabinets in operation an essential part of the batch disadvantage may be eliminated; for instance, with four cabinets a very good leveling of loads will be achieved. A large number of industrial freeze drying plants operate today in this way as multicabinet batch plants [1,5,14].

11.4.2.3 Tunnel Freeze Dryers

In the tunnel type of freeze dryer (Figure 11.6), the process takes place in a large vacuum cabinet into

which the tray-carrying trolleys are loaded at intervals through a large vacuum lock at one end of the tunnel and discharged similarly at the other end.

The freeze dryer shown in Figure 11.6 consists of a tunnel with vacuum locks at each end, one for loading deep-ribbed aluminum trays containing frozen lumps of food into the tunnel and the other for discharging the freeze-dried product into an air-conditioned room where the dry product is automatically removed by machinery before packaging. The drying conditions are carefully controlled in a number of sections of the tunnel by temperature-pneumatic controllers [1]. Vapor constriction plates, fitting closely inside the walls of the tunnel yet allowing the trolleys to pass through, are at two locations in the main section of the tunnel, and gate valves shut off the locks from the main section. The tunnel is thus separated into five independent process zones.

During the period when the trolley is not moving, a tray-lifting device causes all the trays in each trolley to sit on top of the heaters below. The heaters have flat top surfaces and ribs underneath through which vacuum steam circulates. They are cantilevered in pairs from both sides of the tunnel. Vacuum steam heating has several advantages, including a high latent heat of condensation and temperature control by means of pressure.

The refrigeration system consists of a large aqua-ammonia absorption refrigerator instead of a compression plant, mainly because of the ease with which the refrigeration load can be varied by controlling the oil feed to the boiler that heats the absorber.

The total capacity of a tunnel freeze dryer can be increased as the volume of business increases. Large

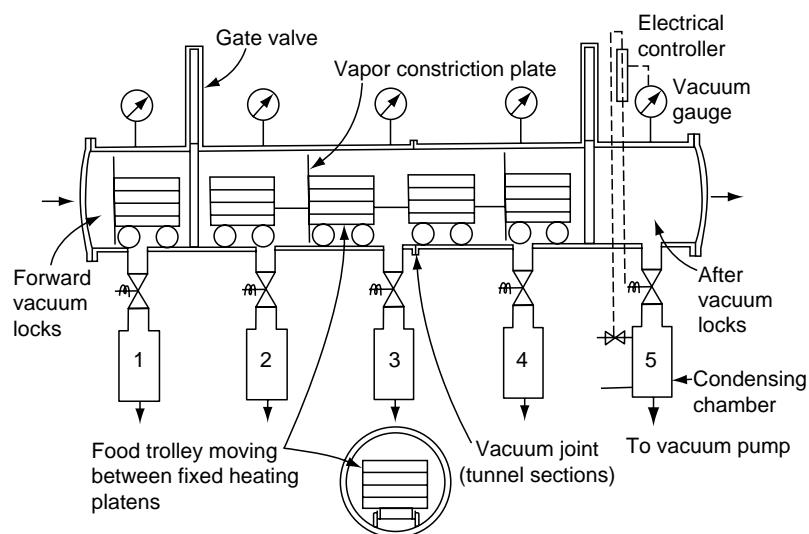


FIGURE 11.6 Schematic diagram of a typical tunnel freeze dryer. (From Mellor, J.D., *Fundamentals of Freeze Drying*, Academic Press, London, 1978.)

commercial plants for processing cottage cheese and coffee have been built up in this way.

The tunnel freeze dryers have the same advantages of plant capacity utilization that can be achieved as in multibatch plants, but the flexibility for simultaneous production of different products or in switching from one product to another is lacking.

11.4.2.4 Vacuum-Spray Freeze Dryers

The vacuum-spray freeze dryer shown in Figure 11.7 has been developed for coffee extract, tea infusion, or milk. The product is sprayed from a single jet upward or downward in a cylindrical tower of 3.7-m diameter by 5.5-m high [1,34]. The liquids solidify into small particles by evaporative freezing. In the tower a refrigerated helical condenser is coiled between the inside wall and a central hopper, the latter collecting the partially dry powder as it falls freely to the bottom of the tower, which in turn is connected to a tunnel where the drying process is completed on a stainless steel belt traveling between radiant heaters. The product passes into a hopper that feeds a vacuum lock, permitting intermittent removal of the product for packing. The whole plant operates under a vacuum of about 67 Pa. Frozen particles obtained by spraying into a vacuum are about 150 μm in diameter and lose about 15% moisture in the initial evaporation. There is no sticking of these particles.

Generally, sprayed freeze-dried coffee has less flavor than normal freeze-dried coffee and the product from this plant is no exception. However, it is hoped retention can be improved in the dried product by concentration before spraying into the tower.

11.4.2.5 Continuous Freeze Dryers

Recent years have shown a growing interest in freeze drying plants operating with a continuous flow of material through the process. Particularly in industries working with a single standardized product and the preparation of the product is by a continuous process, such plants are really profitable. They give continuity in processing throughout and constant operating conditions that are easily controlled, and they require less manual operation and supervision. A particular incentive comes from the prospect of balancing the load imposed on the water vapor condensation system and the vacuum system. In a batch process, the water vapor evolution rate from the foodstuff is quite high at the start of drying and becomes less as drying proceeds. The condenser system must be designed to handle the maximum water vapor removal requirement.

Continuous freeze dryers are used for freeze drying of product in trays and for freeze drying of agitated bulk materials. When handling the product in trays, the most delicate treatment of the product is

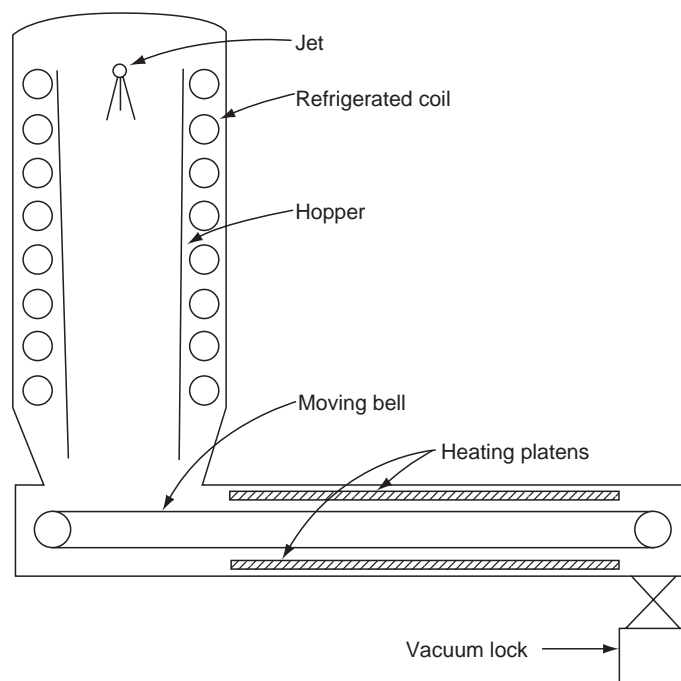


FIGURE 11.7 Layout of a vacuum-spray freeze dryer. (From Mellor, J.D., *Fundamentals of Freeze Drying*, Academic Press, London, 1978.)

achieved. The product is stationary in the tray and therefore is not exposed to abrasion, and it comes in contact only with surfaces that fully meet standards of hygiene.

When agitating a granulated product, more effective heat transfer to the single product particles can be achieved, and thus a considerable reduction of the heating surface is possible. But, both these conditions—abrasion of the product by agitation and increased water vapor production per unit heating surface—tend to carry small product particles with the vapor stream away from the bulk product bed and to cause loss of product. Any complications in the system for water vapor removal to recover the product loss may more than offset the advantage of the higher heater surface load.

The heat transfer to the product and to the trays is by radiation, which in the easiest way safeguarding a correct and an evenly distributed heat transfer to the material during the process. The radiant heat is produced by horizontal heater plates grouped in temperature zones. Each tray remains for a fixed period of time in each temperature zone in such a way that the drying time is minimized.

The Conrad system [5] is a commonly used continuous freeze dryer for treating product in trays. The success of this type of plant is based on the simplicity and reliability of each component that goes to form the total system. The details of this and other continuous systems are given in Refs. [5,8].

11.5 FREEZE DRYING COSTS

Freeze drying is costly because of the long drying times involved; this factor has hindered the application of the technique to drying of materials in bulk. As a result, fixed costs tend to exceed running costs. Anquez [35] has given some estimates of these costs for foodstuffs, and it is usual that fixed costs exceed running costs by 1.5 to 2.5 times [1].

The annual capacity of each freeze dryer is based on a 20-h/d and 250 working days per year, with 4 h/d allowed for loading and unloading operations. Thus,

$$\text{Annual capacity} = 5000 \times \text{rated capacity} \\ (\text{kg ice per hour})$$

11.5.1 FIXED COST

The average fixed cost is given by the ratio of the annual fixed cost to the annual capacity, assuming an annual fixed cost consisting of capital depreciation

at 7.5% per year and loan charges at 8% reducible by the expression

$$C = \frac{R_1(1 + R_1)^n}{(1 + R_1)^{n-1}} \quad (11.1)$$

where C equals the annual charge to repay \$1 loan and interest for n years at a rate R_1 . Thus with $R_1 = 0.08$ for 10 y, $C = 0.149$ and so the loan charge per capital is \$0.049. Keey [36] points out that this expression also gives the capital recovery factor (fraction of the original capital investment set aside each year over the working life of the plant) for comparing drying systems. Mellor [1] shows that a plant would be preferred that is dearer to install but costs less to run. Other fixed capital-dependent charges at 5% of capital cost include maintenance, insurance, and taxes.

11.5.2 RUNNING COST

The running cost consists of labor and utilities costs. Data on the thermophysical properties of foods and biological materials, required for estimating utilities costs, are not always at hand and so the calculation is based only on those properties pertaining to the frozen water content of the material.

Only one person is required to operate any of the dryers, at y dollars per hour. This should increase to $2y$ per hour to cover operating supplies, supervision, payroll overhead, plant overhead, and process control. Then, average labor costs per kilogram ice equals $(\$2y)24/20x$ (plant throughput per hour). Preparation and packing costs are not included as these will depend largely on the nature of the product. The utilities cost can be estimated from a heat and energy balance for 1-kg water undergoing freezing, sublimation, condensation, and melting.

11.5.2.1 Refrigeration

The heat extracted in freezing the water content of the material from 25 to -30°C in 1 h is equivalent to 502 kJ/kg ice. The heat removed during condensation at -40°C is approximately 2840 kJ/kg ice. The compressor power operating on ammonia is about 0.65 kW/kW (refrigeration); thus the energy to be supplied to the compressor is 1840 kJ/kg ice.

11.5.2.2 Heating

The heat required to sublime ice at -20°C equals 2840 kJ/kg ice. The ice collected at the condenser after the completion of the drying cycle has to be melted at -40°C and requires about 419 kJ/kg ice.

11.5.2.3 Vacuum Pumping

Electricity required for two-stage vacuum pumps equals 0.36 kWh/kg ice.

Total Energy	kJ/kg ice	kWh/kg ice
Refrigeration	(502 + 1840)	0.65
Vacuum pumping		0.36
Total		1.01

If the heat of vaporization of steam is 2065 kJ/kg at a pressure of 682 kPa, then $(2840 + 419)/2065 = 1.58$ kg steam/kg ice is required for heating.

Thus, knowing the prices of 1 kg steam and 1 kWh, the utilities cost can be estimated by using the above information. It is usual to allow an increase to the utilities cost of 20% in order to cover thermal losses and other charges. A more detailed economic analysis is given in Ref. [37], which has economic data, analysis, and evaluations that are based on the various operational policies considered in the research studies presented in Refs. [6,37], which have considered the removal of both frozen and bound water.

11.6 PROCESS MODELING: PARAMETERS AND DRYING RATES

11.6.1 SYSTEM FORMULATION

The goal of the process designer and of the processor is to formulate an economical drying system that gives reliably uniform and high product quality [1,2,38–40]. A knowledge of the basic phenomena and mechanisms involved in freeze drying is essential for this purpose. In the following sections, a qualitative description and a mathematical model of the freeze drying process is presented; the model could be used to analyze [6,7,9–12,16–18,37,41] rates of freeze drying. The question of drying rates is important because of the notably long cycle times or residence times that have been required for freeze drying.

In Figure 11.1, a material that is freeze-dried in a tray is shown. The thickness of the sides and bottom of the tray, as well as the material from which the tray is made, are most often in practice such that the resistance of the tray to heat transfer could be considered to be negligible [1,6,13,42]. Heat q_I could be supplied to the surface of the dried layer by conduction, convection, or radiation from the gas phase; this heat is then transferred by conduction to the frozen layer. Heat q_{II} is supplied by a heating plate and is conducted through the bottom of the tray and through the frozen material to reach the sublimation

interface or plane. The magnitude of the amount of heat q_{III} in the vertical sides of the tray is much smaller [6,13,42] than that of q_I or q_{II} ; q_{III} represents the amount of heat transferred between the environment in the drying chamber and the vertical sides of the tray. Since the contribution of q_{III} is rather negligible when compared to the contributions of q_I and q_{II} , the contribution of q_{III} to the drying rate will not be considered further [6,13,42]. The terms N_w and N_t in Figure 11.1 represent the mass flux of water vapor and the total mass flux, respectively, in the dried layer. The total mass flux is equal to the sum of the mass fluxes of water vapor and inert gas, $N_t = N_w + N_{in}$, where N_{in} denotes the mass flux of the inert gas.

11.6.1.1 Mathematical Model for the Primary Drying Stage

In the primary drying stage sublimation occurs as a result of heat conducted to the sublimation interface through the dried (I) and frozen (II) layers. The resulting water vapor is transported by convection and diffusion through the porous dried layer, enters the vacuum chamber, and finally collects upon the condenser plate. The following assumptions are made in the development of the mathematical model: (a) only one-dimensional heat and mass flows, normal to the interface and surfaces, are considered; (b) sublimation occurs at an interface parallel to and at a distance X from the surface of the sample; (c) the thickness of the interface is taken to be infinitesimal [1,5,6,43]; (d) a binary mixture of water vapor and inert gas flows through the dried layer; (e) at the interface, the concentration of water vapor is in equilibrium with the ice; (f) in the porous region, the solid matrix and the gas are in thermal equilibrium; (g) the frozen region is considered to be homogeneous, of uniform thermal conductivity, density, and specific heat, and to contain a negligible proportion of dissolved gases.

Energy balances in the dried (I) and frozen (II) layers can now be made [6,42,73–77,79,81]

$$\frac{\partial T_I}{\partial t} = \alpha_{Ie} \frac{\partial^2 T_I}{\partial x^2} - \frac{C_{pg}}{\rho_{Ie} C_{ple}} \left(\frac{\partial(N_t T_I)}{\partial x} \right) + \frac{\Delta H_v}{\rho_{Ie} C_{ple}} \left(\frac{\partial C_{sw}}{\partial t} \right), \quad 0 \leq x \leq X \quad (11.2)$$

$$\frac{\partial T_{II}}{\partial t} = \alpha_{II} \frac{\partial^2 T_{II}}{\partial x^2}, \quad X \leq x \leq L \quad (11.3)$$

where $\alpha_{Ie} = k_{Ie}/\rho_{Ie} C_{ple}$, $\alpha_{II} = k_{II}/\rho_{II} C_{pII}$, and $N_t = N_w + N_{in}$. In the dried layer, effective parameters are considered that include the physical properties of

both the gas and solid [6,9,11,38,42]. The initial and boundary conditions are

At $t = 0$

$$T_I = T_{II} = T_X = T^o, \quad 0 \leq x \leq L \quad (11.4)$$

at $x = 0$

$$q_I = -k_{le} \frac{\partial T_I}{\partial x} \Big|_{x=0}, \quad t > 0 \quad (11.5)$$

and

$$q_I = \sigma F(T_{up}^4 - T_I^4|_{x=0}), \quad t > 0 \quad (11.6)$$

for radiation heat transfer to the upper dried surface, at $x = X$

$$k_{II} \frac{\partial T_{II}}{\partial x} - k_{le} \frac{\partial T_I}{\partial x} + V(\rho_{II} C_{pII} T_{II} - \rho_I C_{pI} T_I) + N_t C_{pg} T_X = -\Delta H_s N_w, \quad 0 < t \leq t_{X=L} \quad (11.7)$$

at $x = X$

$$T_I = T_X = T_{II}, \quad t > 0 \quad (11.8)$$

at $x = L$

$$q_{II} = k_{II} \frac{\partial T_{II}}{\partial x} \Big|_{x=L}, \quad t > 0 \quad (11.9)$$

The continuity (material balance) equations for the dried (I) layer are [6,42,73–77,79,81]

$$\varepsilon_p \frac{\partial C_{pw}}{\partial t} + \frac{\partial C_{sw}}{\partial t} + \frac{\partial N_w}{\partial x} = 0 \quad (11.10)$$

$$\varepsilon_p \frac{\partial C_{pin}}{\partial t} + \frac{\partial N_{in}}{\partial x} = 0 \quad (11.11)$$

where N_w and N_{in} represent the mass fluxes of water vapor and inert gas, respectively, in the dried layer. The term $\partial C_{sw}/\partial t$ in Equation 11.2 and Equation 11.10 accounts for the change in the concentration of sorbed or bound water with time. The mass fluxes N_w and N_{in} can be obtained from the following constitutive equations:

$$N_w = -D_{win, e} \frac{\partial C_{pw}}{\partial x} + \left(\frac{C_{pw}}{C_{pw} + C_{pin}} \right) N_t \quad (11.12)$$

$$N_{in} = -D_{win, e} \frac{\partial C_{pin}}{\partial x} + \left(\frac{C_{pin}}{C_{pw} + C_{pin}} \right) N_t \quad (11.13)$$

The total mass flux N_t ($N_t = N_w + N_{in}$) is given by

$$N_t = v_p(C_{pw} + C_{pin}) \quad (11.14)$$

where v_p represents the convective velocity of the gas (water vapor and inerts) in the porous dried (I) layer. The convective velocity v_p of the gas in the dried layer is obtained from Darcy's equation as follows:

$$v_p = -\left(\frac{k}{\mu} \right) \frac{\partial P}{\partial x} \quad (11.15)$$

In the above equation, κ is the permeability of the porous dried (I) layer and μ is the viscosity of the gas. By combining Equation 11.14 and Equation 11.15, the following expression is obtained for N_t :

$$N_t = -(C_{pw} + C_{pin}) \left(\frac{k}{\mu} \right) \frac{\partial P}{\partial x} \quad (11.16)$$

Equation 11.12 and Equation 11.13 are then substituted into Equation 11.10 and Equation 11.11. The term $\partial C_{sw}/\partial t$ in Equation 11.2 and Equation 11.10 can be quantified if a thermodynamically consistent mathematical model could be constructed that could describe the change in the concentration of bound water with time. Different rate mechanisms may be considered [6,11,39,44–47]. One of the rate mechanisms could be given by the following expression:

$$\frac{\partial C_{sw}}{\partial t} = k_t C_{pw}(C_T - C_{sw}) - k_2 C_{sw} \quad (11.17)$$

where C_T denotes the maximum equilibrium concentration of sorbed water, and k_1 and k_2 represent the rate constants of the adsorption and desorption steps, respectively. The parameters k_1 and k_2 can be functions of temperature [39,44–48,73,81]. In Equation 11.17, the term $\partial C_{sw}/\partial t$ is negative if $k_2 C_{sw}$ is greater than $k_1 C_{pw}(C_T - C_{sw})$. Of course, if $k_2 C_{sw} \gg k_1 C_{pw}(C_T - C_{sw})$ for all times and everywhere in the dried layer, then the term $\partial C_{sw}/\partial t$ could be set equal to $-k_2 C_{sw}$ without introducing any significant error in the calculations of the drying rate and time.

The initial and boundary conditions of Equation 11.10, Equation 11.11, and Equation 11.14 through Equation 11.17 are as follows:

$$\text{at } t = 0, \quad C_{pw} = 0 \quad \text{for } x > 0 \quad (11.18)$$

$$\text{at } t = 0, \quad C_{pin} = 0 \quad \text{for } x > 0 \quad (11.19)$$

$$\text{at } t = 0, \quad C_{sw} = C_{sw}^o \quad \text{for } 0 \leq x \leq L \quad (11.20)$$

$$\text{at } x = 0, \quad C_{pw} = C_{pw}^o = M_w \left(\frac{p_w^o}{RT_1|_{x=0}} \right), \quad t \geq 0 \quad (11.21)$$

$$\text{at } x = 0, \quad C_{pin} = C_{pin}^o = M_{in} \left(\frac{p_{in}^o}{RT_1|_{x=0}} \right), \quad t \geq 0 \quad (11.22)$$

$$\begin{aligned} \text{at } x = X, \quad C_{pw} &= C_{pwX} = M_w \left(\frac{p_{wX}}{RT_X} \right) \\ &= M_w \left(\frac{g(T_X)}{RT_X} \right), \quad 0 < t \leq t_{X=L} \end{aligned} \quad (11.23)$$

$$\text{at } x = X, \quad \frac{\partial C_{pin}}{\partial x}|_{x=X} = 0, \quad 0 < t \leq t_{X=L} \quad (11.24)$$

$$\text{at } x = 0, \quad P = P^o = p_{in}^o + p_w^o, \quad t \geq 0 \quad (11.25)$$

$$\begin{aligned} \text{at } x = X, \quad N_t &= -((C_{pw} + C_{pin})|_{x=X}) \left(\frac{k}{\mu|_{x=X}} \right) \\ &\times \left(\frac{\partial P}{\partial x}|_{x=X} \right), \quad 0 < t \leq t_{X=L} \end{aligned} \quad (11.26)$$

The total pressure at $x = X$ is given by $P_X = p_{wX} + P_{inX}$, where $p_{wX} = g(T_X)$. The variable p_w^o is the chamber water vapor pressure determined by the condenser design, and the function $g(T_X)$ represents the thermodynamic equilibrium between the frozen product and the water vapor [1,6,11].

The mathematical model is completely specified by a material balance at the interface that defines its velocity as

$$V = \frac{dX}{dt} = - \frac{N_w}{\rho_{II} - \rho_{Ie}} \quad (11.27)$$

where the variable X (position of the interface) is a function of time, $X = X(t)$.

Equation 11.2 through Equation 11.27 represent the mathematical model that could be used to describe the dynamic behavior of the primary drying stage of the freeze drying process [6,11,42]. This model involves a moving boundary (the position of the sublimation interface) and accounts for the removal of frozen water by sublimation, as well as for the removal of bound water by secondary drying in the dried layer during the primary drying stage. External transport resistances can be easily incorporated into this model by including the expressions developed by Liapis and Litchfield [9].

However, in a well-designed freeze dryer the external mass and heat transfer resistances should not be controlling in determining the drying time. A point

to be stressed is that in any freeze drying process it will be desirable to fix the design and operating conditions so that the process is not rate limited by external resistances to either heat or mass transfer. The internal heat and mass transfer resistances are characteristic of the material dried, but the external resistances are characteristic of the equipment. The design conditions refer to having appropriate capacities for the vacuum pump, the water vapor condenser, and the heaters, and that the spacings between trays are such that the external heat and mass transfer resistances are not significant.

Two limits may possibly be reached during the primary drying stage. First, the surface temperature $T_1(t, o)$ must not become too high because of the risk of thermal damage. Second, the temperature of the interface T_X must be kept well below the melting point. If the outer surface temperature limit (T_{scor}) is encountered first as $T_1(t, o)$ is raised, the process is considered to be heat transfer controlled; to increase the drying rate further, the thermal conductivity k_{Ie} of the dried layer must be raised. Many commercial freeze drying processes are heat transfer controlled [1]. If the melting point temperature T_m is encountered first, then the process is considered to be mass transfer limited and, in order to increase the drying rate, the effective diffusivity of water vapor in the dried layer $D_{win, e}$ and the total mass flux N_t must be raised (an increase in N_t implies that the convective velocity of the vapor in the pores of the dried layer is increased); the values of $D_{win, e}$ and N_t could be raised by decreasing the pressure in the drying chamber.

The frozen layer temperature must be maintained below the melting point, which may in some cases be 10°C or more below the melting point of ice for the reasons discussed previously. Typical ice temperatures existing in the freeze drying of foods under conditions in which the total pressure was primarily due to water vapor and the heat transfer took place via the dried (I) layer are shown in Table 11.1 [5]. Typical ice temperatures existing in the freeze drying of pharmaceuticals under conditions in which the total pressure was primarily due to water vapor and the heat transfer took place via the dried (I) and frozen (II) layers are shown in Table 11.2. Liapis and Sadikoglu [77] proposed and developed a novel dynamic pressure rise method as a remote sensing procedure [64,77] for determining at different times during the primary drying stage of the freeze drying process (i) the temperature of the moving interface between the dried and frozen layers of the product, (ii) the temperature close to the upper surface of the dried layer of the product, (iii) the temperature of the bottom surface of the frozen layer of the product, and (iv) the temperature profile of the frozen layer of the product.

TABLE 11.1
Frozen Layer and Maximum Dry Surface
Temperatures in Typical Freeze Drying Operation
Conducted with Heat Input through the Dry Layer

Food Material	Chamber Pressure (mmHg)	Maximum Surface Temperature (°C)	Frozen Layer Temperature (°C)
Chicken dice	0.95	60	−20
Strawberry slices	0.45	70	−15
Orange juice	0.05–0.1	49	−43
Guava juice	0.05–0.1	43	−37
Shrimp	0.1	52	−29
Shrimp	0.1	79	−18
Salmon steaks	0.1	79	−29
Beef, quick frozen	0.5	60	−14
Beef, slow frozen	0.5	60	−17

Source: From Goldblith, S.A., Rey, L., and Rothmayr, W.W., *Freeze Drying and Advanced Food Technology*, Academic Press, London, 1975.

Furthermore, by knowing the temperature of the heating plate and determining the value of the temperature of the moving interface from the dynamic pressure rise method, the value of the position of the moving interface could be determined by an expression developed by Liapis and Sadikoglu [77].

It should be noted that in the above model, diffusion of sorbed water on the surface of the pores of the dried layer (surface diffusion) and diffusion of sorbed water in the solid material of the dried layer (solid diffusion) were not considered. The data of Pikal et al. [10] suggest that the contribution of solid diffusion in the removal of bound water is not significant. The

surface diffusion mass flux of bound water could be incorporated [73,81] in the continuity equation for the bound water, and this would increase the complexity of the model; furthermore, the value of the surface diffusion coefficient has to be estimated. The data of Pikal et al. [10] appear to indicate that the desorption (evaporation) of bound water represents the rate-limiting mass transfer process in secondary drying. This model accounts for the desorption of bound water in secondary drying.

The equations of the model presented earlier can be solved by the numerical method developed by Liapis and Litchfield [49] and Millman [42]. This method immobilizes the moving boundary and transforms the problem of the freeze drying process into a problem of fixed extent; then the numerical solution of the partial differential equations is obtained by the method of orthogonal collocation [39,42,49]. This model has been found [11,16,73,75,79,81] to provide theoretical predications that agree well with the experimental freeze drying rate and time data. It should be mentioned at this point that, in certain pharmaceutical products, because of their processing origin or for freezing processing purposes, solvents other than water are used together with water. In this case, the mass flux of the solvent, the material balance equation for the solvent, and the rate expression for the removal of bound solvent have to be introduced in the structure of the mathematical model presented earlier.

For the lyophilization of a pharmaceutical product in vials, the mathematical model in Refs [78,90–92] should be used. This model accounts for the removal of frozen and bound water, and the temperature and concentration variables vary with time and with two space variables (one space variable is along the length of the cylindrical vial and the other is along the radial coordinate of the vial).

TABLE 11.2
Frozen Layer and Maximum Dry Surface
Temperatures in Typical Freeze Drying
Operation Conducted with Heat Input through
the Dry and Frozen Layers

Pharmaceutical Material	Chamber Pressure (mmHg)	Maximum Surface Temperature (°C)	Frozen Layer Temperature (°C)
Ampicillin sodium salt	0.15	40	−24
Cloxacillin sodium salt	0.20	45	−20
Cephalosporin sodium salt	0.15	40	−25
Collagen	0.30	70	−20

11.6.1.2 Mathematical Model for the Secondary Drying Stage

In the secondary drying stage, there is no frozen (II) layer, and thus there is no moving sublimation interface. The secondary drying stage involves the removal of bound water. The thickness of the dried (I) layer is L , and the energy balance in this layer (it has the same form as Equation 11.2) is as follows [6,42,73–77, 79,81]:

$$\frac{\partial T_I}{\partial t} = \alpha_{Ie} \frac{\partial^2 T_I}{\partial x^2} - \frac{C_{pg}}{\rho_{Ie} C_{ple}} \left(\frac{\partial(N_I T_I)}{\partial x} \right) + \frac{\Delta H_v}{\rho_{Ie} C_{ple}} \left(\frac{\partial C_{sw}}{\partial t} \right), \quad 0 \leq x \leq L \quad (11.28)$$

The initial and boundary conditions of Equation 11.28 are

$$\text{at } t_s = 0, \quad T_I = \Psi(x), \quad 0 \leq x \leq L \quad (11.29)$$

$$\text{at } x = 0, \quad q_I = -k_{Ie} \frac{\partial T_I}{\partial x} \Big|_{x=0}, \quad t_s > 0 \quad (11.30)$$

and,

$$q_I = \sigma F(T_{up}^4 - T_I^4|_{x=0}), \quad t_s > 0 \quad (11.31)$$

for radiation heat transfer to the upper dried surface,

$$\text{at } x = L, \quad q_{II} = k_{Ie} \frac{\partial T_I}{\partial x} \Big|_{x=L}, \quad t_s > 0 \quad (11.32)$$

The material balance equations for the water vapor and the inert gas are given by Equation 11.10 and Equation 11.11, and the constitutive expressions for N_w and N_{in} are obtained from Equation 11.12 and Equation 11.13. Equation 11.17 represents one possible form for the rate expression of the removal of bound water (see Section 11.2.2), and this equation (or its simpler form, as discussed above) could be used to describe the change in the concentration of sorbed water with time. The total mass flux N_t through the porous dried layer is obtained from Darcy's law, and is given by Equation 11.16.

The initial and boundary conditions of Equation 11.10, Equation 11.11, Equation 11.16, and Equation 11.17 in the secondary drying stage, are given by the following expressions:

$$\text{at } t_s = 0, \quad C_{pw} = \gamma(x), \quad 0 \leq x \leq L \quad (11.33)$$

$$\text{at } t_s = 0, \quad C_{pin} = \delta(x), \quad 0 \leq x \leq L \quad (11.34)$$

$$\text{at } t_s = 0, \quad C_{sw} = \theta(x), \quad 0 \leq x \leq L \quad (11.35)$$

$$\text{at } x = 0, \quad C_{pw} = C_{pw}^o = M_w \left(\frac{p_w^o}{RT_I|_{x=0}} \right), \quad t_s > 0 \quad (11.36)$$

$$\text{at } x = 0, \quad C_{pin} = C_{pin}^o = M_{in} \left(\frac{p_{in}^o}{RT_I|_{x=0}} \right), \quad t_s > 0 \quad (11.37)$$

$$\text{at } x = 0, \quad P = P^o = p_{in}^o + p_w^o, \quad t_s > 0 \quad (11.38)$$

$$\text{at } x = L, \quad \frac{\partial C_{pw}}{\partial x} \Big|_{x=L} = 0, \quad t_s > 0 \quad (11.39)$$

$$\text{at } x = L, \quad \frac{\partial C_{pin}}{\partial x} \Big|_{x=L} = 0, \quad t_s > 0 \quad (11.40)$$

The functions $\Psi(x)$, $\gamma(x)$, $\delta(x)$, and $\theta(x)$ provide the profiles of T_I , C_{pw} , C_{pin} , and C_{sw} at the end of the primary drying stage or at the beginning of the secondary drying stage; these profiles are obtained by the solution of the model equations for the primary drying stage. The total pressure at $x = L$ is given by $P_L = p_{wL} + p_{inL}$.

Equation 11.10 through Equation 11.17 and Equation 11.28 through Equation 11.40 represent the mathematical model that could be used to describe the dynamic behavior of the secondary drying stage of the freeze drying process; this model has been found [75] to provide theoretical predictions that agree well with the experimental freeze drying rate and time data. The numerical solution of the partial differential equations of this model can be obtained by the method of orthogonal collocation [6,39,42,49]. External transport resistances can be easily incorporated into this model by including the expressions developed by Liapis and Litchfield [9]. But, as it was discussed above (for the model of the primary drying stage), in a well-designed freeze dryer the external resistances should not be controlling in determining the drying time.

For the lyophilization of a pharmaceutical product in vials, the mathematical model in Refs [78,90–92] should be used to describe the dynamic behavior of the secondary drying stage (see Section 11.2.2).

11.6.1.3 Effect of Chamber Pressure on the Heat and Mass Transfer Parameters of the Dried Layer

The effective thermal conductivity k_{Ie} in the dried material has been found to vary significantly with the total pressure and with the type of gas present. At very low pressures the thermal conductivity reaches a lower asymptotic value independent of the surrounding gas. This asymptotic conductivity reflects the geometric structure of the solid matrix itself, with no contribution from the gas in the voids of the material since the gas pressure is so low.

At high pressures the thermal conductivity levels out again at a higher asymptotic value. This higher asymptote is characteristic of the heterogeneous matrix composed of solid material and the gas in the voids. Consequently, the high-pressure thermal conductivity is dependent on the nature of the gas present and specifically increases as the thermal conductivity of the gas increases and, hence, as the molecular weight of the gas decreases.

When the thermal conductivity attains the high-pressure asymptotic value, the mean free path of the gas molecules within the void spaces of the dried layer

has become substantially less than the dimensions of the void spaces. During the transition in thermal conductivity from the low-pressure asymptote to the high-pressure asymptote, the mean free path of the gas molecules rivals the void space dimensions in magnitude, but once the mean free path is reduced to the point at which the gas phase within the solid matrix obeys simple kinetic theory, the thermal conductivity stops rising.

This reflects the fact that the thermal conductivity of a gas obeying simple kinetic theory is independent of the pressure. The transition in thermal conductivity between asymptotes usually occurs between 0.1 and 100 mmHg, which includes the pressures characteristic of freeze drying processes. The pressure range over which the transition in thermal conductivity between asymptotes occurs is characteristic of the pore size distribution of the void spaces within the freeze-dried material [50]. A smaller pore dimension means that the gas must achieve a higher pressure in order for the mean free path of the gas to become comparable to

the pore spacing and, hence, means that the transition between asymptotes will occur at higher pressures.

Since fast freezing before freeze drying leads to smaller pore spacing after freeze drying [51,52], it follows that faster freezing should lead to lower thermal conductivities at a given pressure. If a freeze drying process is rate limited by internal heat transfer, the rate of freeze drying for fast-frozen material should be less than that of a slowly frozen material. Slower freeze drying rates for food pieces frozen more rapidly have been reported [51,52]. Also, Triebes and King [53] and Saravacos and Pilsworth [54] have found that the thermal conductivity of freeze-dried materials is higher at higher relative humidities of the surrounding gas, in rough proportion to the volume fraction of sorbed water present, and weighted in proportion to the thermal conductivity of liquid water.

Table 11.3 and Table 11.4 summarize thermal conductivity measurements of freeze-dried food substances and pharmaceuticals, respectively. The surrounding gases and the pressure range are indicated,

TABLE 11.3
Thermal Conductivities of Freeze-Dried Food Substances

Food Substance	Surrounding Gas	Range of Pressures (mmHg)	Range of Thermal Conductivities (Btu/h ft °F)
Beef	Water vapor	0.5–2.4	0.020–0.032
Mushrooms	Air	0.3–760	0.006–0.021
Cornstarch solutions	Water vapor and air	0.1–2.0	0.008–0.019
Beef	Air	0.001–760	0.022–0.038
Apple	Air	0.001–760	0.009–0.024
Peach	Air	0.001–760	0.009–0.025
Pear	Freon-12, carbon dioxide, nitrogen, neon, hydrogen	0.02–760	0.013–0.108
Apple	Same	0.02–760	0.013–0.115
Beef	Same	0.02–760	0.022–0.116
Apple	Water vapor	0.01–0.3	0.020–0.067
Milk	Water vapor	0.01–0.3	0.013–0.047
Salmon	Water vapor	0.15	0.024–0.077
Haddock	Water vapor	0.08	0.011–0.015
Perch	Water vapor	0.08	0.013–0.020
Beef	Water vapor and air	0.007–80	0.020–0.037
Beef	Water vapor	0.2–3.0	0.030–0.042
Potato starch	Water vapor and air	0.03–760	0.005–0.024
Gelatin	Same	0.03–760	0.009–0.024
Cellulose gum	Same	0.03–760	0.011–0.032
Egg albumin	Same	0.03–760	0.008–0.024
Pectin	Same	0.03–760	0.007–0.024
Tomato juice	Water vapor	0.4–1.5	0.020–0.100
Turkey	Air, water vapor, Freon-12, carbon dioxide, helium	0.01–760	0.008–0.112

Source: From King, C.J., *Freeze-Drying of Foods*, CRC Press, Cleveland, OH, 1971.

TABLE 11.4
Thermal Conductivities of Freeze-Dried
Pharmaceutical Materials

Pharmaceutical Material	Surrounding Gas	Range of Pressures (mmHg)	Range of Thermal Conductivities (Btu/h ft °F)
Ampicillin sodium salt	Water vapor	0.01–0.15	0.012–0.040
Cloxacillin sodium salt	Water vapor	0.01–0.20	0.012–0.040
Cephalosporin sodium salt	Water vapor	0.01–0.15	0.012–0.017
Collagen	Water vapor	0.1–0.4	0.015–0.040

along with the range of thermal conductivities encountered. Thermal conductivities can be measured by the use of a thermopile apparatus or may be inferred from actual freeze drying rate measurements [8]. It will probably be helpful in many cases to make use of the thermal conductivity models for porous media [8,53,55,56] in order to extrapolate and interpolate data to different conditions.

As shown in Table 11.3 and Table 11.4, the thermal conductivities of dry layers of foods and pharmaceuticals are extremely low compared with the conductivities of insulators, such as cork and styrofoam. As a consequence, the temperature drop across the dry layer is large, and with surface temperatures often limited to values below 65°C because of danger of discoloration and in some cases to values below 38°C because of the danger of denaturation, the resultant ice temperature is usually well below –18°C. Except for materials with very low melting points, it is the surface temperature that limits the drying rate.

Because of this limitation, drying rates attainable in practice are much below the maximum rates attainable with ice. Thus, for materials loaded into the freeze dryer at about 8–18 kg/m² of tray surface, which corresponds to industrial practice [1], average drying rates are of the order of 1.5 kg of water removed per square meter. The corresponding drying times are 6–10 h. A much more rapid rate can be achieved by decreasing the particle size and loading rates [6,37]. This corresponds to the reduction of the average thickness of the dry layer and thus of mass and heat transfer resistances. This approach, however, is limited to selected products, since efficient operation requires specialized equipment, such as continuous freeze dryers.

The effective diffusivities $D_{win, e}$ and $D_{inw, e}$ in the dried material are functions of the structure of the

material, Knudsen diffusivity, and molecular diffusivity. Simplified expressions for $D_{win, e}$ and $D_{inw, e}$ are given as [38]

$$D_{win, e} = \frac{\varepsilon_p}{\tau} \left(\frac{1}{(1 - \alpha y_w)/D_{win} + 1/D_{Kw}} \right) \quad (11.41)$$

$$D_{inw, e} = \frac{\varepsilon_p}{\tau} \left(\frac{1}{(1 - \beta y_{in})/D_{inw} + 1/D_{Kin}} \right) \quad (11.42)$$

where τ is the tortuosity factor of the porous dried layer and y_w and y_{in} are the mole fractions of water vapor and inerts, respectively. The expressions for α and β in Equation 11.41 and Equation 11.42 are as follows: $\alpha = 1 + (N_{in}M_w/N_wM_{in})$ and $\beta = 1 + (N_wM_{in}/N_{in}M_w)$. The Knudsen and the molecular diffusivities can be obtained from the following expressions [38,57]:

$$D_{Kw} = 97.0 \bar{r} (T_I/M_w)^{1/2} \quad (11.43)$$

$$D_{Kin} = 97.0 \bar{r} (T_I/M_{in})^{1/2} \quad (11.44)$$

$$D_{win} = \frac{1.8583 \times 10^{-7} T_I^{3/2}}{P \sigma_{win}^2 \Omega_{win}} \left(\frac{1}{M_w} + \frac{1}{M_{in}} \right)^{1/2} \quad (11.45)$$

$$D_{inw} = \frac{1.8583 \times 10^{-7} T_I^{3/2}}{P \sigma_{inw}^2 \Omega_{inw}} \left(\frac{1}{M_{in}} + \frac{1}{M_w} \right)^{1/2} \quad (11.46)$$

In Equation 11.43 and Equation 11.44, the term \bar{r} is the average pore radius in the dried layer (m). In Equation 11.45 and Equation 11.46, the terms σ_{win} and σ_{inw} are the average collision diameters, and Ω_{win} and Ω_{inw} are the collision integrals.

From Equation 11.45 and Equation 11.46, it can be observed that the magnitude of D_{win} and D_{inw} would decrease if the total pressure P is increased. If the value of D_{win} is decreased because of increased total pressure, then Equation 11.41 indicates that the value of the effective pore diffusivity $D_{win, e}$ could decrease as the total pressure is increased. Therefore, if the total pressure in the drying chamber is increased, then the effective diffusivity $D_{win, e}$ could decrease, and thus the diffusional mass flux of water vapor in the dried layer could decrease.

Furthermore, when the total pressure in the drying chamber is increased, the gradient of the total pressure $\partial P/\partial x$ in the dried layer could be reduced, and this could decrease the convective velocity v_p and the total mass flux N_t (Equation 11.14 through Equation 11.16). As the freeze drying process will become internal, mass transfer controlled above a certain pressure ($D_{win, e}$ and N_t decrease with increasing pressure, and k_{te} increases with pressure), the highest rate

under mass transfer control will occur at the pressure of transition from heat transfer control to mass transfer control and the attainable drying rate will decrease at higher pressures.

More elaborate expressions of $D_{\text{win},e}$ and $D_{\text{inw},e}$ can be found in Refs. [6,9,11,16,17,58]. These expressions are more complex than Equation 11.41 and Equation 11.42. In all cases, $D_{\text{win},e}$ and $D_{\text{inw},e}$ decrease with increasing total pressure.

In general, operating conditions in freeze drying of foods include maximum surface temperatures of 38–82°C and chamber pressures of 0.1–2 mmHg. Freeze drying of biological specimens, vaccines, and microorganisms is usually conducted with maximum surface temperatures of 20–32°C and chamber pressures below 0.1 mmHg. It is possible to conduct freeze drying at atmospheric pressure, provided the gas in which the drying is conducted is very dry. In this case, the heat transfer is improved but the external and internal mass transfer rates deteriorate; mass transfer becomes limiting in atmospheric freeze drying. As a consequence, for all but very small particles, the drying rates are very slow.

11.7 CONTROL VARIABLES AND POLICIES IN FREEZE DRYING

Much interest has been focused on ways to reduce the drying times of the freeze drying process so that the proportional amortization and operating costs are minimized. Thus, there has been considerable interest in investigating the factors affecting the batch time of freeze dryers, since this variable is most amenable to control, and efforts have been made to minimize the batch time [1,6,9,12,37,42,59,76,78,79,82,83].

The heat variables q_I and q_{II} from the energy sources and the drying chamber pressure P_{ch} , $P_{\text{ch}} \cong P^o = p_w^o + p_{\text{in}}^o$ ($P_{\text{ch}} \cong P^o$ when the external mass transfer resistance is insignificant, as would be the case with a well-designed freeze dryer) are natural control variables. It should be emphasized at this point that (as the equations of the mathematical models for the primary and secondary drying stages and the material in the preceding section indicate) the effects on the heat and mass transfer rates resulting from changes in the values of q_I , q_{II} , and P_{ch} are coupled.

The variable p_w^o is taken to represent the water vapor pressure in the drying chamber (the external mass transfer resistance is taken to be insignificant) and its value is determined by the design and the operational temperature of the ice condenser. Thus, P_{ch} may be changed by changes in p_w^o (p_w^o could be changed by changes in the temperature of the ice

condenser), and by increasing or decreasing p_w^o . Therefore, changes in the temperature of the ice condenser affect the pressure P_{ch} (through p_w^o) in the drying chamber, and thus the mass transfer rate in the dried layer. The controls q_I , q_{II} , and P_{ch} (the temperature of the ice condenser and the value of p_{in}^o can increase or decrease the value of P_{ch}) must be selected from a set of admissible controllers

$$\begin{aligned} q_I^* &\leq q_I \leq q_I^* \\ q_{II}^* &\leq q_{II} \leq q_{II}^* \\ q_{\text{ch}}^* &\leq q_{\text{ch}} \leq q_{\text{ch}}^* \end{aligned} \quad (11.47)$$

This set of controllers excludes those that would produce an unacceptable product quality. Two important constraints on the product state are that the surface temperature $T_I(t, o)$ must not exceed the scorch point of the dried product

$$T_I(t, o) \leq T_{\text{scor}} \quad (11.48)$$

and that the frozen interface and, in general, the frozen layer must not melt

$$T_X \leq T_m \quad (11.49)$$

and

$$T_{II}(t, x) \leq T_m, \quad X \leq x \leq L \quad (11.50)$$

If during the drying run $T_I(t, o) = T_{\text{scor}}$ and $T_X < T_m$, then the process is called heat transfer controlled, and if $T_I(t, o) < T_{\text{scor}}$ and $T_X = T_m$, then the process is considered to be mass transfer controlled.

The objective is to minimize the total batch time t_b , equivalent to defining a performance index of the form

$$\bar{Q} = \min_{q_I, q_{II}, P_{\text{ch}}} Q = \min_{q_I, q_{II}, P_{\text{ch}}} \int_0^{t_b} dt \quad (11.51)$$

where t_b is defined as the time when a fixed amount of water remains in the product. The problem given in Equation 11.47 and Equation 11.51 along with a mathematical model of the freeze drying process (see Section 11.6.1.1 and Section 11.6.1.2) is the standard time optimal control problem [9,12,76,79,83].

Liapis and Litchfield [9] performed a quasisteady-state analysis for a system where $q_I \neq 0$ and $q_{II} = 0$ and obtained general guidelines about the optimal control policy at the beginning of the drying process (when neither of the state constraints is active), as well as during operation, when the process may be heat or mass transfer limited [6,9].

The complete unsteady-state optimal control problem has been studied by Litchfield and Liapis [12] for a system where $q_I \neq 0$ and $q_{II} = 0$ using turkey meat and nonfat-reconstituted milk as model foodstuffs. The results of the dynamic analysis for nonfat-reconstituted milk confirm the suggested control policies of the quasisteady-state analysis. At low chamber pressures the dynamic analysis with turkey meat showed control results similar to those obtained by the quasi-steady-state analysis. However, at higher pressures the assumed control policy based on the quasisteady-state analysis was not optimal. The optimal control dynamic study of Litchfield and Liapis [12] suggested that the policies of the quasisteady-state analysis may be useful guidelines but they should be interpreted with some caution. To obtain accurate optimal control policies on the heat input and chamber pressure of the freeze drying process, the complete unsteady-state optimal control problem should be solved [76,78,79,82,83].

Millman et al. [6] studied the freeze drying of skim milk under various operational policies that included the case where $q_I \neq 0$ and $q_{II} \neq 0$. They found that the control policy that produced the shortest primary drying stage was also the policy that provided the shortest overall drying time. Their results show that at least 80% of the heat used during the primary drying stage was transferred through the frozen layer of the sample. They also showed that the type of criterion used in terminating the secondary drying stage is of extreme importance, especially for samples of large thicknesses, as it may lead to an undesirable sorbed (bound) water profile that may deteriorate the quality of the dried product.

Mellor [1] has suggested that periodically time-varied chamber pressure will produce improvements in drying time when compared with conventional steady-state pressure operation. The basis of the argument was that heat must be transmitted through an insulating dried layer to the ice interface in order to provide energy for sublimation and that the average effective thermal conductivity of the dried layer can be enhanced by cycling the pressure. Several industrial- and pilot-scale cyclic pressure plants have been constructed, mainly in Australia by CSIRO, and substantial reductions in drying time have been reported [1].

It should be noted that any analysis or evaluation of the cyclic pressure freeze drying process should involve nonsteady-state heat and mass transfer equations like those presented in [Section 11.6.1.1](#) and [Section 11.6.1.2](#). The effectiveness of cyclic pressure freeze drying and the effect of cycle period and shape on drying times have been the subject of a number of investigations [1,60,61]. Litchfield and Liapis [12]

found that optimal policies with respect to pressure could be closely approximated by a constant pressure policy over the entire period of the primary drying stage. This near-optimal constant pressure policy formed the basis for comparison with the cyclic pressure process [1,60]. The results for turkey meat showed that all the cyclical policies tried were inferior (although only slightly in some cases) to a near-optimal constant pressure policy developed by Liapis and Litchfield [9] and Litchfield and Liapis [12]. Since the capital cost of a cyclic pressure process is considerably greater than that of a constant pressure process [1,8], it appears that the latter process would be preferred.

It should be noted that an increase in pressure will increase k_{Ie} but at the expense of resistance to mass transfer, which is also increased. Hence, the mass flux is reduced and consequently the temperature of the sublimation interface and of the frozen layer is increased. Thus, the thermal conductivity increases but the temperature driving force is decreased.

A simple analysis [2,60] has shown that an increase in pressure (which will increase k_{Ie}) will not guarantee an increase in heat transfer and that an optimum pressure may exist that will maximize the heat flux. If, furthermore, such an optimum does exist, then a cyclic policy will have no beneficial effect since perturbations in either direction away from the optimum will be detrimental.

An optimum chamber pressure for turkey meat has been found experimentally by Sandall et al. [62] and has also been established through theoretical analysis [12]. Litchfield et al. [61] compared cyclic pressure and near-optimal constant pressure freeze drying processes in a situation in which operation at a pressure that would minimize drying time was not possible because of an interface temperature constraint; at no time during the entire run did the cycled pressure process prove superior. In view of these considerations, it may be inferred that for materials exhibiting an attainable optimum with respect to pressure, there will be no advantage in cycling the chamber pressure when compared with near-optimal constant pressure operation.

Sadikoglu et al. [76], Sadikoglu [79], Sheehan and Liapis [78], Sheehan et al. [82], and Liapis et al. [83] have studied the optimal control of the primary and secondary drying stages of the freeze drying of solutions in trays and in vials. Their results strongly motivate the aggressive control of freeze drying and they have found that heat input and drying chamber pressure control that runs the process close to the melting and scorch temperature constraints yields (i) faster drying times and (ii) more uniform distributions of temperature and concentration of bound water at the end of the secondary drying stage.

It is important to report at this point that since a batch of pharmaceutical product in an industrial freeze dryer can easily be worth significant amounts of money, it is of paramount importance that the units of the plant and the control systems of the process should always operate under conditions at which there is insignificant loss in the quality of the product that is freeze-dried. For this purpose, the freeze dryer usually has one additional refrigeration or vacuum unit in a standby condition and furthermore the control instrumentation is designed in such a way that the control policies can be implemented either automatically by computer or by manual override.

11.8 CONCLUSION

The evolution of freeze drying in the last 55 years indicates that this separation process (unit operation) is a convenient method for drying those decomposable products (mostly pharmaceuticals, e.g., plasma, vaccines, antibiotics, sera, and growth hormones) that cannot be stabilized in any other way or that show markedly improved quality for a rather high average cost (coffee, mushrooms, diced chicken, and others); however, most food products are still dried by conventional means for obvious economic reasons. The absence of interfacial forces during freeze drying has been exploited to produce highly dispersed, homogeneous, free-flowing, and very reactive powders [15]. This separation method has found uses in the engineering ceramics area [3] and in the synthesis of superconducting powders [4].

The economics of the process indicate that freeze drying can be suitable for high-value products with specific biological or physicochemical properties. The highest cost advantages would be obtained from the processing of concentrated solutions of expensive materials; in this respect, freeze drying could represent a viable alternative to filtration and crystallization. It is certain that freeze drying has a future, and it is likely that this will essentially be in the fields of food, chemistry, materials science, biological sciences, medicine, pharmaceuticals, and biotechnology. Its potential evolution is still great and will, of course, depend upon progress in basic research and upon the level of creativity in the design and operating conditions of plants and instruments.

NOMENCLATURE

C	annual charge to repay \$1 loan and interest
C_{pg}	heat capacity of gas in the dried layer, kJ/kg K
C_{pin}	concentration of inert gas in the dried layer, kg/m ³

C_{pin}^o	concentration of inert gas at $x = 0$, kg/m ³
C_{pw}	concentration of water vapor in the dried layer, kg/m ³
C_{po}^o	concentration of water vapor at $x = 0$, kg/m ³
C_{pwX}	concentration of water vapor at $x = X$, kg/m ³
C_{pie}	effective heat capacity of dried layer, kJ/kg K
C_{pII}	heat capacity of frozen layer, kJ/kg K
C_{sw}	concentration of bound water, kg/m ³ dried layer
C_{sw}^o	initial concentration of bound water, kg/m ³ dried layer
C_T	maximum equilibrium concentration of bound water, kg/m ³ dried layer
D_{inw}	molecular diffusivity of a binary mixture of inert gas and water vapor (Equation 11.46), m ² /s
$D_{inw, e}$	effective pore diffusivity of a binary mixture of inert gas and water vapor in the dried layer (Equation 11.42), m ² /s
D_{Kin}	Knudsen diffusivity for inert gas (Equation 11.44), m ² /s
D_{Kw}	Knudsen diffusivity for water vapor (Equation 11.43), m ² /s
D_{win}	molecular diffusivity of a binary mixture of water vapor and inert gas (Equation 11.45), m ² /s
$D_{inw, e}$	effective pore diffusivity of a binary mixture of water vapor and inert gas in the dried layer (Equation 11.41), m ² /s
$g(T_X)$	functional form of the thermodynamic equilibrium between the water vapor and the frozen layer at the temperature of the sublimation interface, T_x ($p_{wX} = g(T_X)$), N/m ²
k_1	rate constant in Equation 11.17, m ³ /kg s
k_2	rate constant in Equation 11.17, per s
k_{1e}	effective thermal conductivity in the dried layer, kW/m K
k_{II}	thermal conductivity in the frozen layer, kW/m K
L	sample thickness, m
M_{in}	molecular weight of inert gas, kg/kg mole
M_w	molecular weight of water vapor, kg/kg mole
N_{in}	mass flux of inert gas in the dried layer, kg/m ² s
N_t	total mass flux in the dried layer ($N_t = N_{in} + N_w$), kg/m ² s
N_w	mass flux of water vapor in the dried layer, kg/m ² s
P_{inX}	partial pressure of inert gas at $x = X$, N/m ²
p_{in}^o	partial pressure of inert gas at $x = 0$, N/m ²
p_w^o	partial pressure of water vapor at $x = 0$, N/m ²
p_{wX}	partial pressure of water vapor in equilibrium with the sublimation front ($p_{wX} = g(T_X)$), N/m ²

P	total pressure ($P = p_{\text{in}} + p_{\text{w}}$) in the dried layer, N/m^2
P°	total pressure at $x = 0$, N/m^2
P_{ch}	total pressure in the drying chamber, N/m^2
q_{I}	heat flux at $x = 0$, kW/m^2
q_{II}	heat flux at the bottom of the tray, kW/m^2
\underline{Q}	performance index
\bar{Q}	optimum performance index
R	gas law constant
R_{I}	interest rate
t	time, s
t_{b}	batch time, s
t_{s}	time for secondary drying stage, s
$t_{X=L}$	time at which the sublimation front arrives at $x = L$, s
T_{I}	temperature in the dried layer, K
T_{II}	temperature in the frozen layer, K
T°	initial temperature, K
T_{m}	melting temperature, K
T_{scor}	scorch temperature, K
T_X	temperature of the sublimation front, K
T_{up}	temperature of upper plate, K
ν_{p}	convective velocity of the binary mixture of water vapor and inert gas in the porous dried layer (Equation 11.15), m/s
V	velocity of the sublimation front (Equation 11.27), m/s
x	space coordinate, m
X	position of sublimation front (interface), m

GREEK SYMBOLS

ΔH_{s}	heat of sublimation of ice, kJ/kg
ΔH_{v}	heat of vaporization of bound water, kJ/kg
ΔP	pressure drop, N/m^2
ΔT	total temperature difference in Figure 11.2, K
ΔT_{ice}	temperature difference through the layer of ice on the cold surface, K
ΔT_{refr}	temperature difference between the cold surface and the evaporating refrigerant, K
ε_{p}	void fraction in the dried layer
κ	permeability of the porous dried layer, m^2
μ	viscosity of the binary mixture of water vapor and inert gas in the porous dried layer, kg/m s
ρ_{Ie}	effective density of the dried layer, kg/m^3
ρ_{II}	density of the frozen layer, kg/m^3
σ	Stefan–Boltzmann constant
τ	tortuosity factor of the porous dried layer

SUPERSCRIPTS

* maximum value

SUBSCRIPTS

* minimum value
 I dried layer
 II frozen layer

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12

Microwave and Dielectric Drying

Robert F. Schiffmann

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It is sometimes surprising to realize that dielectric and microwave heating have been in use for quite some time. It appears to many engineers that these are new forms of heating when in fact practical applications began during World War II and the home microwave oven was invented shortly after World War II. Yet, these remain small industries, and for the most part, the equipment manufacturers are likewise small companies. The older of the two, dielectric heating, is a “workhorse” heating method used in many industries, including plastics, wood, ceramics, furniture, textiles, and paper. It is also by far the larger of the two industries; however, it is also not very glamorous, and the industrial microwave heating industry has glamor, but limited sales. To try to quantify the relationship, there are probably only 100 to 150 MW of microwave power in use globally for industrial heating purposes, whereas a single large dielectric heating system may employ as much as 2 or 3 MW of power. The annual worldwide sales of industrial microwave heating systems probably amounts to only less than 100 million dollars, but the sales of the home microwave ovens in the United States is of the order of 1.5 to 2.0 billion dollars. The reasons for the relatively small size of these markets are several, but two stand out: first, the heating mechanisms are not familiar to most engineers, and second, they often represent a radical departure from conventional systems and there is generally a tendency to resist real innovation in most industries.

In the past few years, there has been a surge of interest in the applications of microwave and dielectric heating for industrial purposes. This is primarily

due to the worldwide energy crisis and the growing acceptance of and familiarity with microwave ovens. The unique heating mechanisms of microwaves and dielectrics permit dramatic energy savings in many instances, as well as providing other benefits. This is nowhere better seen than in some of the applications in drying. The purpose of this chapter is to provide background into these heating methods and their applications and, it is hoped, thereby stimulate their consideration in new drying systems.

12.1 BACKGROUND

The terms “dielectric” and “microwave” are somewhat confusing and must be defined as best we can. The term “dielectric heating” can be applied logically to all electromagnetic frequencies up to and including at least the infrared spectrum. The lower frequency systems operate at frequencies through at least two bands: high frequency (HF) (3–30 MHz) and very high frequency (VHF) (30–300 MHz). Thus the names HF, dielectric, radio frequency (RF), and RF heating can often be used interchangeably. However, it is generally accepted that dielectric heating is done at frequencies between 1 and 100 MHz, whereas microwave heating occurs between 300 MHz and 300 GHz. This makes the wavelengths in dielectric heating extend to many meters. Microwave wavelengths range from 1 mm to 1 m. [Table 12.1](#) and [Table 12.2](#) show the industrial, scientific, and medical (ISM) bands established by international agreement [1–3]. Note that these frequency allocations are made

TABLE 12.1

Frequencies Designated by the International Telecommunication Union for Use as Fundamental Industrial, Scientific, and Medical Frequencies^a

Center Frequency (MHz)	Frequency Range (MHz)	Maximum Radiation Limit ^b	Number of Appropriate Footnote to the Table of Frequency Allocation to the ITU Radio Regulations
6.780	6.765–6.795	Under consideration	524 ^c
13.560	13.553–13.567	Unrestricted	534
27.120	26.957–27.283	Unrestricted	546
40.680	40.66–40.70	Unrestricted	548
433.920	433.05–434.79	Under consideration	661 ^c , 662 (region 1 only)
915.000	902–928	Unrestricted	707 (region 2 only)
2450	2400–2500	Unrestricted	752
5800	5725–5875	Unrestricted	806
24,125	24,000–24,250	Unrestricted	881
61,250	61,000–61,500	Under consideration	911 ^c
122,500	122,000–123,000	Under consideration	916 ^c
245,000	244,000–246,000	Under consideration	922 ^c

^aResolution No. 63 of the ITU Radio Regulations applies.

^bThe term “unrestricted” applies to the fundamental and all other frequency components falling within the designated band. Special measures to achieve compatibility may be necessary where other equipment satisfying immunity requirements (e.g., EN 55020) is placed close to ISM equipment.

^cUse of these frequency bands is subject to special authorization by administrations concerned in agreement with other administrations with radio communication services that might be affected.

Source: From IEC CISPR Publication 11, 2nd ed., 1990–09, *Limits and Methods of Measurement of Electromagnetic Disturbance; Characteristics of Industrial, Scientific, and Medical (ISM) Radio-Frequency Equipment*. With permission.

by the International Telecommunication Union (ITU) and some frequencies are specific to certain countries. For example, 915 MHz is allowed in the United States but not in European countries.

TABLE 12.2

Frequencies Designated on a National Basis in CENELEC Countries for Use as Fundamental Industrial, Scientific, and Medical Frequencies

Frequency (MHz)	Maximum Radiation Limit ^a	Notes
0.009–0.010	Unlimited	Germany only
3.370–3.410	Unlimited	Netherlands only
13.533–13.553	110 dB(μV/m) at 100 m	United Kingdom only
13.567–13.587	110 dB(μV/m) at 100 m	United Kingdom only
83.996–84.004	130 dB(μV/m) at 30 m	United Kingdom only
167.992–168.008	130 dB(μV/m) at 30 m	United Kingdom only
886.000–906.000	120 dB(μV/m) at 30 m	United Kingdom only

^aDistance measured from the exterior wall outside the building in which the equipment is situated.

Source: From Cenelec European Standard, CISPR 11, modified, *Limits and Methods of Measurement of Radio Disturbance Characteristics of Industrial, Scientific, and Medical (ISM) Radio-Frequency Equipment*, 1991. With permission.

Practical heating applications, including those for drying, are done at 13.56, 27.12, 40.68, 896, 915, and 2450 MHz. Note that a great deal of dielectric industrial heating in the United States is done at frequencies other than these ISM bands, as explained below.

Although the basic principles of heating and drying at dielectric and microwave frequencies are the same, the methods of generation and equipment are different. These will be described separately later in the chapter. In other cases, the two terms may be used interchangeably in the text.

12.2 FUNDAMENTALS OF MICROWAVE AND DIELECTRIC HEATING

12.2.1 ELECTROMAGNETIC WAVES

We are surrounded by electromagnetic waves at all times. Light, x-irradiation, TV, AM, and FM radio waves, ultraviolet, infrared, and microwaves are some of the common manifestations of these waves. All bodies in the universe, above absolute zero temperature, emit electromagnetic waves. The relationship of these waves is found in the electromagnetic spectrum (see Figure 12.1). All electromagnetic waves are characterized by their wavelength and frequency, and an

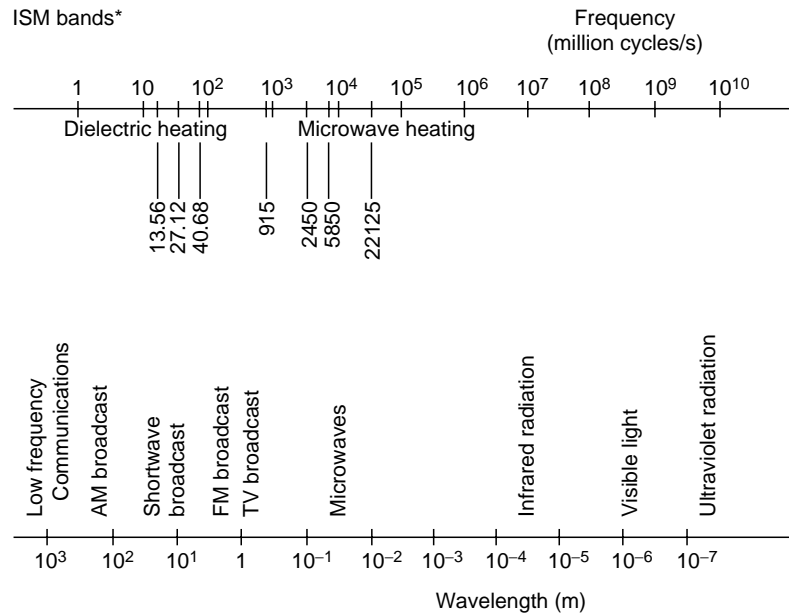


FIGURE 12.1 The electromagnetic spectrum.

illustration of a plane monochromatic electromagnetic wave is seen in Figure 12.2. It is seen that an electromagnetic wave is a blend of an electric component E and a magnetic component H . Note that E and H are perpendicular to each other and both are perpendicular to the direction of travel. This is what makes this a “plane” wave. Note further that the field strength at any point may be represented by a sine or cosine function, which is what makes it “monochromatic.” Further, it is “linearly polarized” as the electric and magnetic field vectors E and H lie in one direction only. The plane of polarization is YX for the E vector and ZX for the H vector.

Figure 12.1 also indicates that the wave is traveling in the X -direction at the velocity C , which is equivalent to the speed of light in air or vacuum but slows as it passes through another medium, as indicated in Equation:

$$V_p = \frac{C}{\sqrt{\epsilon'}} \quad (12.1)$$

where V_p is the velocity of propagation, C is the speed of light in air, and ϵ' is the dielectric constant of the material through which the wave is propagated.

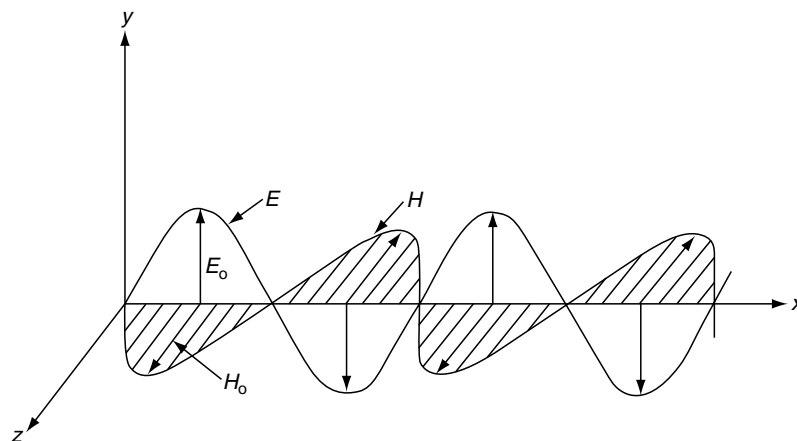


FIGURE 12.2 Diagrammatic illustration of a plane electromagnetic wave. E and H represent the electrical and magnetic components of the wave; E_0 and H_0 are their respective amplitudes.

We further note the distance λ , which is the wavelength. Equation 12.2 shows the relationship between the frequency of the wave f , that is, how many times it goes through a full cycle per second, and the wavelength:

$$f = \frac{V_p}{\lambda} \quad (12.2)$$

Note that as an electromagnetic wave passes through a material its frequency remains the same; therefore, its wavelength changes, and this affects the depth of penetration, and will be discussed later.

A further examination of Figure 12.1 indicates that an electromagnetic wave is an energy wave that changes its energy content and amplitude as it travels through a medium, as seen by a change in the amplitude of the wave. For example, if we trace the E component we see that at some point it is zero; then it builds up to a maximum value, decays to zero, and again builds up to a maximum value with the opposite polarity before again decaying to zero. The same thing happens to the H component. The amplitude of the wave at any point along the X -axis represents the electrical (E) or magnetic (H) field strength, which are measured as volts or amperes per unit distance, respectively. It is this periodic flip-flopping of the wave's polarity and its decay through zero that cause the stress upon ions, atoms, and molecules, which is converted to heat, and the greater the field strength, the greater will be the whole effect.

12.2.2 HEATING MECHANISM

A crucial fact to keep in mind at all times is that microwaves and dielectrics are not forms of heat but rather forms of energy that are manifested as heat through their interaction with materials. It is as if they cause materials to heat themselves. There are many mechanisms for this energy conversion, as can be seen in only a partial listing in Table 12.3 [3].

12.2.3 IONIC CONDUCTION

Since ions are charged units they are accelerated by electric fields. In a solution of salt in water, for example, there are sodium, chloride, hydronium, and hydroxyl ions, all of which will be caused to move in the direction opposite to their own polarity by the electric field. In doing so, they collide with unionized water molecules, giving up kinetic energy and causing them to accelerate and collide with other water molecules in billiard ball fashion, and when the polarity changes the ions accelerate in the opposite fashion. Since this occurs many millions of times per second,

TABLE 12.3
Partial List of Energy Conversion Mechanisms^a

Ionic conduction
<i>Dipole rotation</i>
Entire molecule quantized
Twist
Bend
Interface polarization
Dipole stretching
Ferroelectric hysteresis
Electric domain wall resonance
Electrostriction
Piezoelectricity
Nuclear magnetic resonance
Ferromagnetic resonance
Ferrimagnetic resonance

^aIt is the first two with which we are primarily interested in dielectric heating phenomena.

large numbers of collisions and transfers of energy occur. Therefore, there is a two-step energy conversion: electric field energy is converted to induced ordered kinetic energy, which in turn is converted to disordered kinetic energy, at which point it may be regarded as heat. This type of heating is not dependent to any great extent upon either temperature or frequency. The power developed per unit volume (P_v) through ionic conduction is shown as

$$P_v = E^2 q n \mu \quad (12.3)$$

where q is the amount of electrical charge on each of the ions, n is the ion density, the number of ions per unit volume, and μ is the level of mobility of the ions.

12.2.4 DIPOLAR ROTATION

Many molecules, such as water, are dipolar in nature; that is, they possess an asymmetric charge center. Water is typical of such a molecule. Other molecules may become "induced dipoles" because of the stresses caused by the electric field. Dipoles are influenced by the rapidly changing polarity of the electric field. Although they are normally randomly oriented, the electric field attempts to pull them into alignment. However, as the field decays to zero (relaxes), the dipoles return to their random orientation only to be pulled toward alignment again as the electric field builds up to its opposite polarity. This buildup and decay of the field, occurring at a frequency of many millions of times per second, causes the dipoles similarly to align and relax millions of times per second. This causes an energy conversion from electrical field

energy to stored potential energy in the material and then to stored random kinetic or thermal energy in the material. This temperature-dependent, molecular size-dependent time for buildup and decay defines a frequency known as the “relaxation frequency.” For small molecules, such as water and monomers, the relaxation frequency is already higher than the microwave frequency and rises further as the temperature increases, causing a slowing of energy conversion. On the other hand, large molecules, such as polymers, have a relaxation frequency at room temperature that is much lower than the microwave frequency but that increases and approaches it as the temperature rises, resulting in better energy conversion into heat. This may lead to runaway heating in materials that at room temperature are very transparent to the microwave field. This must be superimposed upon the fact that such liquids as water and monomers are better absorbers of microwave energy than polymers. Since in drying or curing applications it is the liquids and monomers that require heating, not the existing polymeric substrate, it is possible to execute the process well, often at a lower temperature. In fact, it is even possible to dry such materials as foods and medicinals at cold or subfreezing temperatures.

The power formula for dipolar rotation is

$$P_v = kE^2 f \epsilon' \tan \delta$$

or

$$P_v = kE^2 f \epsilon'' \quad (12.4)$$

where k is the constant dependent upon the units of measurement used, E is the electric field strength, in volts per unit distance, f is the frequency, ϵ' is the relative dielectric constant, or relative permittivity, $\tan \delta$ is the loss tangent or dissipation factor, and ϵ'' is the loss factor.

The relative dielectric constant expresses the degree to which an electric field may build up within a material when a dielectric field is applied to the material. The loss tangent is a measure of how much of that electric field will be converted into heat.

A further examination of Equation 12.4 reveals that E and f are functions of the equipment, whereas ϵ' , ϵ'' , and $\tan \delta$ are factors related to the material that is heated. Another important point is that as frequency f is changed it is necessary to increase the electric field strength E in order to maintain a particular power level P_v . Since dielectric heating frequencies are much lower than microwave frequencies, this requires that the field strengths be much higher for comparable power output in a dielectric system. This may lead to voltage breakdown of air (arcing) or

in the process material. The sparking threshold for air is about 30,000 V/cm (75,000 V/in.).

12.2.5 INTERACTION OF ELECTROMAGNETIC FIELDS WITH MATERIALS

We may divide materials and the way they interact with electromagnetic fields into four categories:

1. *Conductors.* Materials with free electrons, such as metals, are materials that reflect electromagnetic waves just as light is reflected by a mirror. These materials are used to contain and direct electromagnetic waves in the form of applicators and waveguides.
2. *Insulators.* Electrically nonconductive materials, such as glass, ceramics, and air, act as insulators, which reflect and absorb electromagnetic waves to a negligible extent and primarily transmit them (that is, they are transparent to the waves). They are therefore useful to support or contain materials to be heated by the electromagnetic field and may take the form of conveyor belts, support trays, dishes, or others. These materials may also be considered “nonlossy dielectrics.”
3. *Dielectrics.* These are materials with properties that range from conductors to insulators. There is within this broad class of materials a group referred to as “lossy dielectrics,” and it is this group that absorbs electromagnetic energy and converts it to heat. Examples of lossy dielectrics are water, oils, wood, food, and other materials containing moisture, and the like.
4. *Magnetic compounds.* These are materials, such as ferrites, that interact with the magnetic component of the electromagnetic wave and as such will heat. They are often used as shielding or choking devices that prevent leakage of electromagnetic energy. They may also be used for heating in special devices.

As indicated earlier, those properties that govern whether a material may be successfully heated by a dielectric or microwave field are the dielectric properties: relative dielectric constant ϵ' , loss tangent or dissipation factor ($\tan \delta$), and the loss factor ϵ'' .

Note that the complex dielectric constant ϵ may be expressed as

$$\epsilon = \epsilon' - j\epsilon'' \quad (12.5)$$

where $j = \sqrt{-1}$, which indicates a 90° phase shift between the real (ϵ') and imaginary (ϵ'') parts of the complex dielectric constant. The loss tangent is

defined as the ratio of dielectric loss to dielectric constant:

$$\tan \delta = \frac{\epsilon''}{\epsilon'} \quad (12.6)$$

These factors are affected by several parameters.

12.2.5.1 Moisture Content

The amount of free moisture in a substance greatly affects its dielectric constant since water has a high dielectric constant, approximately 78 at room temperature; that of base materials is of the order of 2. Thus, with a larger percentage of water the dielectric constant generally increases, usually proportionally. It should be emphasized that very complex phenomena occur when different dielectrics are mixed. However, a few rules of thumb may be applied.

1. The higher the moisture content, usually the higher is the dielectric constant.
2. The dielectric loss usually increases with increasing moisture content but levels off at values in the range of 20 to 30% and may decrease at still higher moisture.
3. The dielectric constant of a mixture usually lies between that of its components.

Various materials, including alcohols and some organic solvents, also exhibit dielectric properties that make them suitable for heating with microwave and dielectric energy and, so, behave similarly to water. Table 12.4 indicates the heating properties of various classes of materials.

Since drying is concerned with the removal of water or a solvent, it is interesting to note that as these liquids are removed the dielectric loss decreases and hence, the material heats less well. In many cases, this leads to self-limitation of the heating as the material becomes relatively transparent at low moisture content. This has great value in obtaining moisture leveling, especially in sheet materials, in which the electromagnetic energy is likely to preferentially dry the wetter areas. Figure 12.3 shows a general graph of the variation in loss factor with moisture content [4]. Water exists in materials in different states, for example, bound or free, and these states may be ascribed to different regions on the graph, as indicated by the change of slope ($d\epsilon''/dm$). Thus at low moisture contents, below the critical moisture content, we are dealing primarily with bound water; above it we encounter primarily free water. (Note that the dielectric loss of bound water is very low since it is not free to rotate under the influence of

TABLE 12.4
Heating Properties of Various Materials

Heat Well	Heat Poorly
Water	Hydrocarbons
Acid anhydrides	Halogenated hydrocarbons (symmetrical)
Alcohols	
Aldehydes	Alkali halides (e.g., salt)
Ketones	Inorganic oxides (e.g., alumina)
Amides	
Amines	Some elements (e.g., sulfur)
Nitrates	Boron nitride
Cyanides	Mica
Proteins	
Halogenated hydrocarbons (unsymmetrical)	
Ferrites	
Ferroelectrics	
Ionic solutions	

Source: From White, J., *Transactions of the International Microwave Power Institute*, 1:40–61 (1973), Manassas, Virginia.

the electromagnetic field. This is seen in an analogous situation with ice, which has a dielectric loss factor of approximately 0.003; that of water is approximately 12.) The change in the slope may be quite gradual for some materials, making positive identification fairly difficult. The critical moisture content for highly hygroscopic materials occurs between 10 and 40% (dry basis); for nonhygroscopic materials it is in the region of about 1%. It is obvious that moisture leveling will be quite effective above the critical moisture content but not so effective below it. Although some materials become quite transparent below the critical

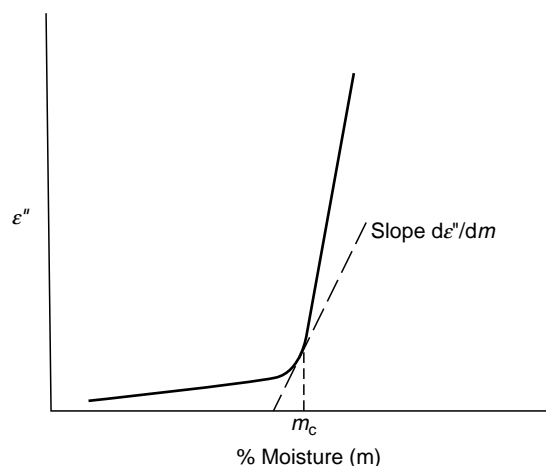


FIGURE 12.3 The critical moisture content m_c . The dielectric loss factor is ϵ'' . The region below m_c is indicative of bound water, whereas above free water is more easily removed.

moisture content, there are others, such as wood and textiles that will continue to heat and may scorch or burn.

12.2.5.2 Density

The dielectric constant of air is 1.0, and it is, for all practical purposes, transparent to electromagnetic waves at industrial frequencies. Therefore, its inclusion in materials reduces the dielectric constants, and as density decreases so do the dielectric properties, and heating is reduced.

12.2.5.3 Temperature

The temperature dependence of a dielectric constant is quite complex, and it may increase or decrease with temperature depending upon the material (see Section 12.2.4). In general, however, a material below its freezing point exhibits lowered dielectric constant and dielectric loss. Above freezing the situation is not clear-cut, and since moisture and temperature are important to both drying and dielectric properties, it is important to understand the functional relationships in materials to be dried. Wood, for example, has a positive temperature coefficient at low moisture content [5]; that is, its dielectric loss increases with temperature. This may lead to runaway heating, which in turn will cause the wood to burn internally if heating continues once the wood is dried.

12.2.5.4 Frequency

Dielectric properties are affected by the frequency of the applied electromagnetic field. However, since industrial heating is restricted to the ISM frequencies, the engineer is limited in making use of this phenomenon. It may, however, be useful in measuring moisture content.

12.2.5.5 Conductivity

Conductivity refers to the ability of a material to conduct electric currents by the displacement of electrons and ions; this effect is described in detail in Section 12.2.3. Suffice it to say that these charged units can have a major effect on heating, and in a drying situation in which the ion concentration increases as the water is removed, this effect can be very complex.

12.2.5.6 Thermal Conductivity

Thermal conductivity often plays a lesser role in microwave and dielectric heating than in conventional heating because of the great speed with which the former heat thus reducing the time in which thermal

conductivity can be effective. There are cases, however, in which it has a major role. For example, when penetration depth of the electromagnetic energy is small in comparison with the volume heated, thermal conductivity may be depended upon to transfer the heat to the interior. Another important case is to even out the nonuniformities of heating that may occur with electromagnetic fields. Sometimes the microwave or dielectric power is pulsed on and off to allow for this evening out of temperature, as in microwave thawing.

12.2.5.7 Specific Heat

The specific heat parameter is often neglected by the researcher or engineer dealing with electromagnetic heating who focuses attention only on the dielectric properties. However, specific heat can have profound effects and may, in fact, be the overriding parameter, causing materials to heat much faster than one would predict by looking only at their dielectric properties.

12.2.5.8 Penetration Depth

Although not a property of a material but rather a result of its various properties, penetration depth is of utmost importance. Since electromagnetic heating is, in effect, bulk heating, it is important that the energy penetrates as deeply as possible. If it does not, then the heating is limited to the surface. Those parameters affecting the depth of the field into the material are the wavelength, the dielectric constant, and the loss factor, as shown in Equation 12.7

$$D = \frac{\lambda_0 \sqrt{2}}{2\pi} \left[\epsilon' \sqrt{1 + (\epsilon''/\epsilon')^2} - 1 \right]^{-1/2} \quad (12.7)$$

where D is the penetration depth at which the available power in the material has dropped to about 37% ($1/e$) of its value at the surface and λ_0 is the free space wavelength. If ϵ'' is low, Equation 12.7 may be simplified:

$$D = \frac{\lambda_0 \sqrt{\epsilon'}}{2\pi \epsilon''} \quad (12.8)$$

This equation is reasonably accurate for most foods even though many have relatively high ϵ'' values.

From these equations, it is obvious that materials with high dielectric constants and loss factors will have smaller depths of penetration than those with lower values. It is also apparent that the depth of penetration is greatly affected by the wavelength (and hence the frequency) of the applied field. This is illustrated in Table 12.5, in which penetration depth

TABLE 12.5
Wavelength and Depth of Penetration in Douglas Fir
at Various Frequencies

	Frequency (MHz)					
	5.0	13.56	27.12	40.0	915	2450
<i>Wavelength</i>						
Meters	60.0	22.1	11.1	7.5	0.328	0.122
Feet	196.9	72.6	36.3	24.6	1.07	0.400
<i>Depth of penetration</i>						
Meters	23.9	8.8	4.4	3.0	0.130	0.049
Feet	78.4	28.9	14.4	9.8	0.425	0.158

Source: From Tinga, W.R., *Proceedings of the International Microwave Power Institute Short Course for Users of Microwave Power*, Manassas, Virginia, 1970, pp. 19–29.

into Douglas fir is shown for various wavelengths [5]. This demonstrates the very clear superiority of dielectric heating of very large materials with substantial dielectric properties.

A peculiarity of this type of heating is the unusual temperature gradients that may be generated. This is due to a number of factors. First, unless an auxiliary heat form is applied, the air in the system remains cold. Hence, the surface will be cooler than a zone somewhat below the surface. This is especially true in a drying system in which evaporative cooling of the surface will occur.

Another circumstance concerns the depth of penetration as it relates to the size of the piece that is heated. If the piece is several times larger than the depth of penetration, then the temperature gradient will resemble conventional gradients, with a cooler interior and a warmer exterior. However, if the piece is small in comparison with the penetration depth, for example only one or two times greater, then there may be a focused accumulation of the electromagnetic field in the center of the piece due to the multiple passes of the waves and internal reflections. In this case, the center may be the hottest place, and in fact, if it is overheated, the center may burn whereas the surface remains cool.

12.3 PROCESS ADVANTAGES OF MICROWAVE AND DIELECTRIC SYSTEMS

12.3.1 ADVANTAGES OF MICROWAVE AND DIELECTRIC HEATING

Heating and drying with microwave and dielectric energy is distinctly different from conventional

means. Whereas conventional methods depend upon the slow march of heat from the surface of the material to the interior as determined by differential in temperature from a hot outside to a cool inside, heating with dielectric and microwave energy is, in effect, bulk heating in which the electromagnetic field interacts with the material as a whole. The heating occurs nearly instantaneously and can be very fast, although it does not have to be. However, the speed of heating can be an advantage, and it is often possible to accomplish in seconds or minutes what could take minutes, hours, and even days with conventional heating methods. The fastest industrial heating system of which this author is aware heats fine plastic thread at the rate of about 30,000°C/s (the material was actually heated about 100°C in about 3 ms) [3]. On the other hand, one can heat at the rate of 1°C per century, if desired. The governing parameters here are the mass of the material, its specific heat, dielectric properties, geometry, heat loss mechanisms, and coupling efficiency, the power generated in the material, and the output power of the microwave–dielectric heating system. If all other things are equal, the speed may be doubled by doubling the output power.

A list of advantages of microwave and dielectric heating includes the following:

1. Process speed is increased, as described above.
2. Uniform heating may occur throughout the material. Although not always true, often the bulk heating effect does produce uniform heating, avoiding the large temperature gradients that occur in conventional heating systems.
3. Efficiency of energy conversion: In this type of heating, the energy couples directly to the material that is heated. It is not expended in heating the air, walls of the oven, conveyor, or other parts. This can lead to significant energy savings. Also, the energy source is not hot and plant cooling savings may be realized.
4. Better and more rapid process control: The instantaneous on–off nature of the heating and the ability to change the degree of heating by controlling the output power of the generator mean fast, efficient, and accurate control of heating.
5. Floor space requirements are usually less. This is because of more rapid heating.
6. Selective heating may occur. The electromagnetic field generally couples into the solvent, not the substrate. Hence, it is the moisture that is heated and removed, whereas the carrier or substrate is heated primarily by conduction. This also avoids heating of the air, oven walls, conveyor, or other parts.

7. Product quality may be improved. Since high surface temperatures are not usually generated, overheating of the surface and case hardening, which are common with conventional heating methods are eliminated. This often leads to less rejected product.
8. Desirable chemical and physical effects may result. Many chemical and physical reactions are promoted by the heat generated by this method, leading to puffing, drying, melting, protein denaturation, starch gelatinization, and the like.

12.3.2 ADVANTAGES OF MICROWAVE AND DIELECTRIC DRYING

The mechanism for drying with microwave and dielectric energy is quite different from that of ordinary drying. In conventional drying, moisture is initially flashed off from the surface and the remaining water diffuses slowly to the surface. Although the potential of energy transfer for heating is the temperature gradient, which results in energy transfer to the interior of the material, the potential for mass transfer is the mass concentration gradient existing between the wet interior and the drier surface. This is often a slow process, diffusion rate limited, which requires high external temperatures to generate the temperature differences required.

With internal heat generation, in microwave and dielectric systems, mass transfer is primarily due to the total pressure gradient established because of the rapid vapor generation within the material [6]. Most of the moisture is vaporized before leaving the sample. If the sample is initially very wet and the pressure inside the sample rises very rapidly, liquid may be removed from the sample under the influence of a total pressure gradient. The higher the initial moisture, the greater is the influence of the pressure gradient on the total mass removal. Thus, there is, in effect, a sort of “pumping” action, forcing liquid to the surface, often as a vapor. This leads to very rapid drying without the need to overheat the atmosphere and perhaps cause case hardening or other surface overheating phenomena. Table 12.6 summarizes the advantages of microwave and dielectric drying.

Of great interest today are the potential energy savings achievable from such a system. This is due to speed of drying, the direct coupling of energy into the solvent, possible lower drying temperatures, far more effective use of conventional heating in combination with the dielectric methods, and less overall heat loss.

A word of caution must be expressed here. These systems can heat and dry quickly, but too rapid heating can be destructive. Care must be taken not to heat so fast that the material may scorch, burn, or

TABLE 12.6
Advantages of Microwave and Dielectric Drying

Efficiency: in most cases, the energy couples into the solvent, not the substrate
Nondestructive: drying can be done at low ambient temperatures; no need to maintain high surface temperatures, leading to lower thermal profiles
Reduction of migration: solvent often mobilized as a vapor; therefore does not transfer other materials to the surface
Leveling effects: coupling tends toward the wetter areas
Speed: drying times can be shortened by 50% or more
Uniformity of drying: by a combination of more uniform thermal profiles and leveling
Conveyorized systems: less floor space, reduced handling
Product improvement in some cases: eliminates case hardening, internal stresses, and other problems

be otherwise damaged or dry so quickly that the steam or other vapors cannot escape quickly enough, leading to internal pressure buildup, which can lead to rupture of the piece or an explosion.

When drying with dielectric heating it is usual to combine hot air with the system, particularly with microwave systems. This is because it usually improves the efficiency and the economics of the drying process [7]. Hot air is, by itself, relatively efficient at removing free water at or near the surface, whereas the unique pumping action of dielectric heating provides an efficient way of removing internal free water as well as bound water. By combining these properly, it is possible to draw on the benefits of each and maximize efficiency and keep the costs of drying down. Note that drying with microwaves or dielectrics alone can be very expensive in terms of both equipment and operating costs.

There are three ways in which microwave and dielectric energy may be combined with conventional drying methods, as illustrated in [Figure 12.4](#).

12.3.2.1 Preheating

By applying the microwave or dielectric energy at the entrance to the dryer, the interior of the load is heated to evaporation temperature, thereby immediately forcing moisture to the surface and immediately permitting the conventional dryer to operate at its most efficient condition, at higher temperatures (Figure 12.4a). The drying curve is steeper, and drying time is shortened.

12.3.2.2 Booster Drying

The microwave or dielectric energy is added to the conventional dryer when the drying rate begins to fall

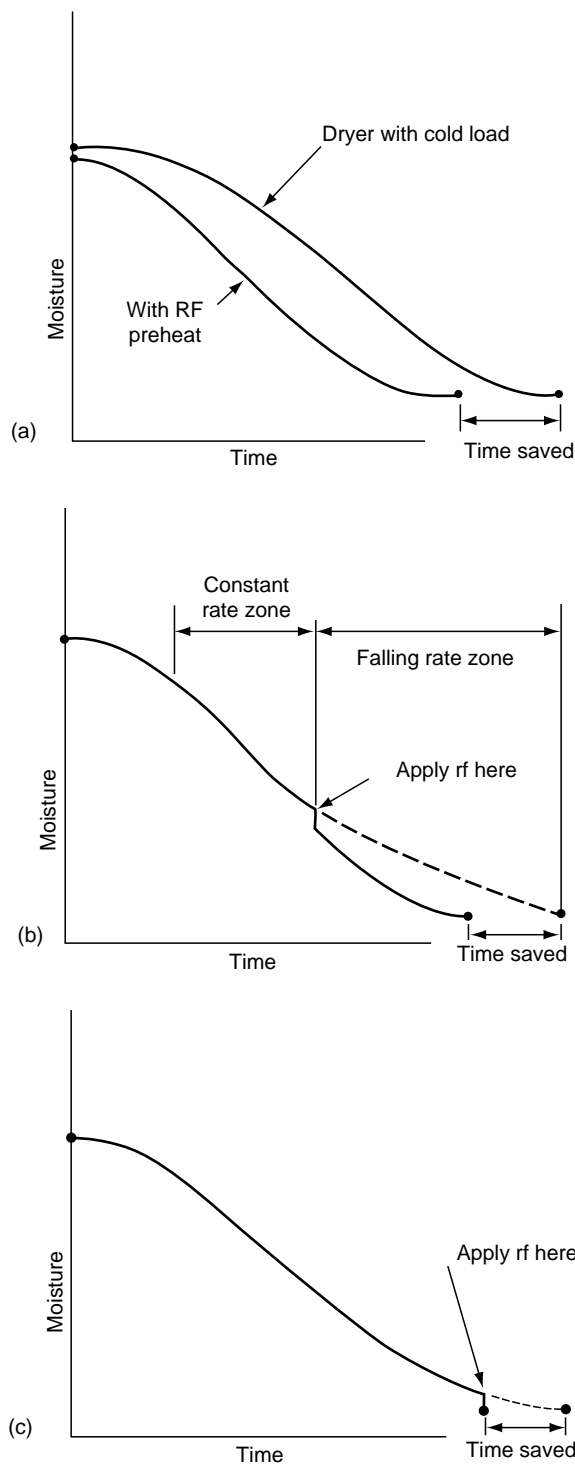


FIGURE 12.4 Typical drying curves for microwave and dielectric drying systems: (a) preheating with microwave or dielectrics; (b) booster drying; (c) finish drying.

off (Figure 12.4b). The surface of the material is dry, and moisture is concentrated in the center. The added electromagnetic energy generates internal heat and vapor pressure, forcing the moisture to the surface,

where it is readily removed. The drying is sharply increased with a leverage of 6:1 or 8:1 in terms of increased drying capacity for each unit of electromagnetic energy added. This is most effective on thick, hard to heat materials.

12.3.2.3 Finish Drying

The least efficient portion of a conventional drying system is near the end, when two thirds of the time may be spent removing the last one third of the water (Figure 12.4c). By adding a microwave or dielectric dryer at the exit of the conventional dryer, this replaces the inefficiency of hot air drying with internal heat generation. The conventional dryer may also be speeded up, thereby increasing the throughput of the dryer whereas presenting the dryer with a wetter load, thus increasing the efficiency. This method also provides close control of the terminal moisture and moisture leveling at the same time avoiding overdrying.

The most common methods of application are booster and finish drying, and in spite of the greater cost for electrical energy than gas, the overall increase in drying efficiency and throughput can bring about large economic savings.

12.4 EQUIPMENT FOR MICROWAVE AND DIELECTRIC HEATING AND DRYING

The heating mechanisms for microwave and dielectric heating are similar, but the means of achieving them are somewhat different. The basic components of these systems are a means of generating the high-frequency energy—the generator—and a means of applying it to the workpiece—the applicator. These are described in the following sections.

12.4.1 GENERATORS

The basic function of the generator is to convert the alternating current of 50 or 60 Hz to the high frequencies desired. The means of doing this are quite different for dielectric and microwave systems.

Dielectric systems usually employ negative grid triode tubes, although some systems operating in the 50 to 100-MHz range use beamed power types. RF circuits are usually simple, self-excited oscillators of the Hartley, Colpitts, or tuned plate-tuned grid type. These circuits usually consist of coils and capacitors or coaxial lines. The load to be heated may actually form part of the tank circuit capacitance, it may be separately tuned or inductively or capacitively coupled to the oscillator, or it may be a combination of these, which is directly connected to the oscillator and also

partially tuned to the oscillator circuit. This last option has a distinct advantage in drying applications, in which the electrodes may be partially tuned to the oscillator circuit on the side of the tuning curve, where they will be detuned when a dry load with a lower dielectric constant is present. Thus, a wet load will cause an increase in electrode voltage, and a dry load will cause a power reduction. In this way, there is a self-limiting or leveling effect [8]. Note that energy not developed in the load can be lost in the coupled circuits, causing less efficiency and wasted power.

A negative grid tube is a variable power device that draws current from the supply line only in sufficient quantity to supply circuit losses plus the power consumed in the load. Thus, most dielectric systems control power to the load by varying the RF electrode voltage, often by a variable capacitor. In many instances, this system is self-limiting and self-regulating, supplying power only as demanded by changes in size or electrical characteristics of the load [8].

Dielectric heating systems utilize a wide range in frequencies, from 3 MHz to more than 150 MHz. Many of these are not confined to the ISM bands and may vary frequency to improve the efficiency of heating. In doing so, however, they must be properly shielded and filtered to comply with Part 18, Subpart D of the FCC Rules and Regulations, to prevent out-of-band radiation leakage.

Microwave systems operate on a nominal fixed frequency of 915 or 2450 MHz, with the frequency controlled by the tube dimensions and geometry. They must also be shielded to prevent excessive radiation of harmonics, as well as for safety.

A microwave generator consists of a dc power supply and a tube—either a magnetron or a klystron. These tubes are constant output power devices, and power to the load may be controlled by sensing the load requirements and controlling the input power accordingly, usually by indirectly varying the dc anode voltage. Although magnetrons and klystrons are capable of withstanding a reasonable degree of mismatching, manifested as power reflected back to the tube, precautions must be taken to avoid overheating or in other ways damaging these tubes (see Section 12.4.3.3).

12.4.2 APPLICATORS

The means of applying microwave and dielectric energy to a workpiece differ in a very significant manner. Microwave energy may be transported through free space and must be focused upon the load. On the other hand, dielectric energy is usually applied by means of electrodes, in which the field oscillates through the load, which is placed between the electrodes. This is described in the following sections.

12.4.2.1 Dielectric Systems

There are many types of electrodes, but they may be described in three basic categories, which are illustrated in Figure 12.5.

The platen type usually consists of flat plates in pairs, between which the workpiece may be held in a batch system or pass on a conveyor belt. Often, a conveyor belt may represent one of the plate electrodes. This is especially useful for bulky objects. A drawback is that as the plates become widely separated the high field strengths required may cause voltage discharge, which can burn the load.

With the stray field type, the load, usually in the form of a thin web, passes over the electrodes of alternating polarity. Since the load represents the path of least electrical resistance, the dielectric field passes through it, causing heating. The staggered type is usually used for sheet materials and thick webs. The distance between the electrodes is kept to a minimum in order to achieve heating without arcing.

As for other types, electrodes may also be shaped to conform to the geometry of the load or may be part of the conveyor or hydraulic or pneumatic press. The entire system is confined within a metal housing to prevent leakage of radiation. At the same time, hot air of controlled temperature, humidity, and velocity may be passed through the applicator.

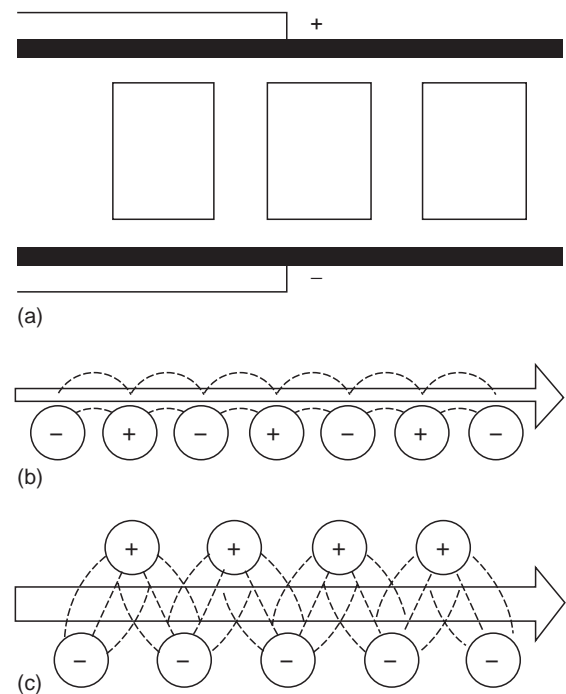


FIGURE 12.5 Electrode configurations for dielectric heating systems: (a) platen type for bulky objects; (b) stray field type for thin webs; (c) staggered type for thick webs or board.

12.4.2.2 Microwave Systems

After generating the microwave energy it must be transported to the applicator. This is usually accomplished by means of waveguides, although coaxial cable is also useful for lower power.

The waveguide is ordinarily a hollow rectangular metallic conduit, usually made of brass or aluminum. Its interior dimensions are carefully chosen to control the nature of the microwave field presented to the applicator. Applicators are of several major types and are always constructed of metal.

Waveguides themselves may be used as applicators. As the electric field may be maximum in the center of a waveguide, it is possible to pass a material through this intense field to obtain very efficient heating. A good example is the heating of filamentary materials.

Traveling wave applicators are also known as slotted, folded, or serpentine waveguides. A slot is cut into the narrow sides of the waveguide, and several waveguides are joined together as shown in Figure 12.6. A thin-sheet material, such as paper or textile, may be passed through the slots. The microwave energy makes several passes through the load, heating it as it travels. These are highly efficient heating systems, although they may cause some side-to-side nonuniformity.

Cavity applicators are a large class of applicators, but they are probably the most common type. Home microwave ovens are a typical example. They consist of metal boxes, which may be used for conveyORIZED systems or in batch operations. The microwave energy may be coupled into this applicator by means of waveguide or coaxial cable through a single port or multiple ports

(see Figure 12.7). There is an industrial system in which over 100 magnetrons are separately introduced into the cavity. In this type of applicator, the load usually represents only a small fraction of the volume of the applicator and is subjected to the microwave field reflected from the sides of the applicator and passing through it from all sides. This causes a three-dimensional bulk heating effect that is unique and is of great use.

A major problem with cavity applicators is uniformity of the microwave field in the load. In order to ensure uniformity of heating, a number of steps may be taken, usually in combination: moving or turning the load in the applicator by means of conveyors or turntables; providing mode stirrers, which often resemble slowly rotating fans and increase the number of modes in the oven, causing reflective scattering of energy; using multiple inputs for the microwave energy; using multiple microwave sources with slight differences in frequency that cause different mode patterns; and choosing the cavity dimensions to support the maximum number of modes, the so-called resonant multimode cavity.

12.4.3 OTHER DEVICES

A number of auxiliary devices and systems should be mentioned at this point.

12.4.3.1 Control Systems

Since the output power of the microwave or dielectric heating system is governed by electrical energy, unique control systems can be designed utilizing feedback

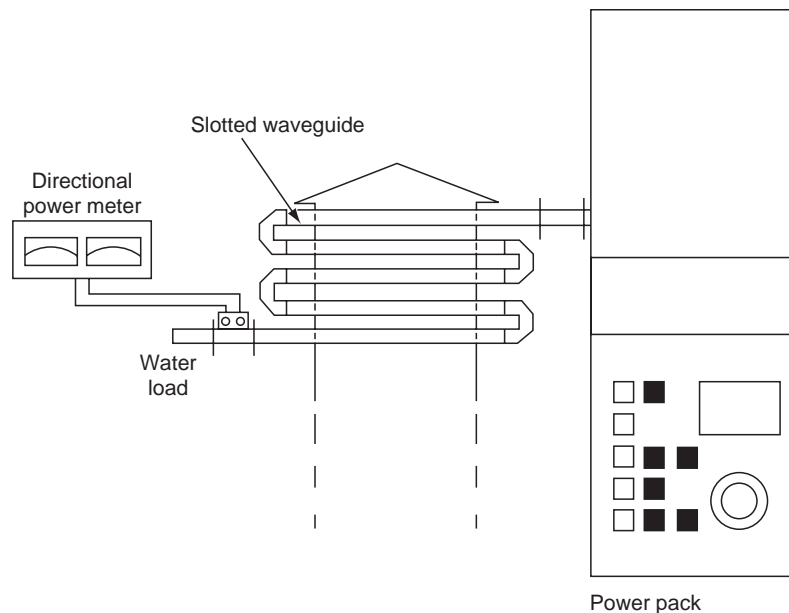


FIGURE 12.6 Slotted or serpentine waveguide. The material to be dried, usually a thin web, is passed through slots in the sides of the waveguide and exposed by multiple passes to the microwave field.

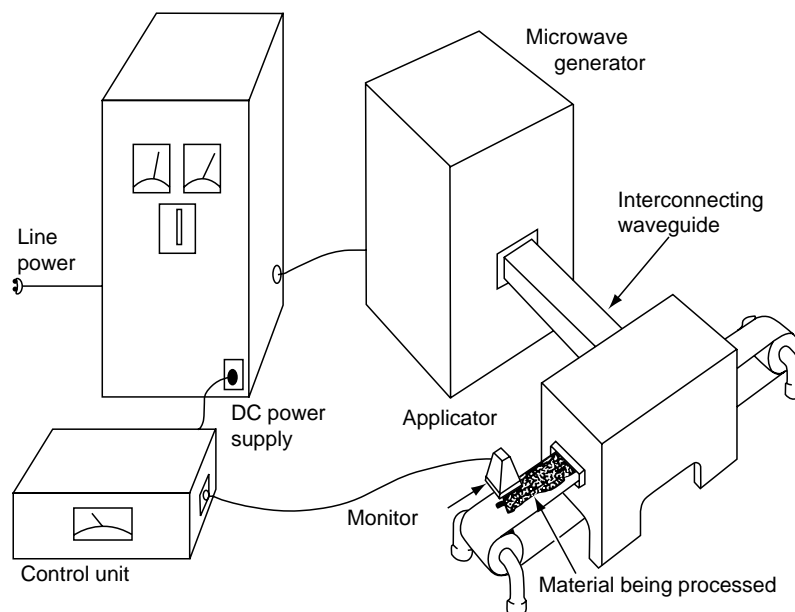


FIGURE 12.7 A typical microwave heating system utilizing a conveyORIZED cavity applicator. A feedback system monitors the heated material and automatically adjusts the output power of the magnetron to control the final moisture.

loops that monitor some function of the load, such as moisture, and automatically control the output power to give better and faster control of the moisture content.

12.4.3.2 Leakage and Safety Control Systems

As mentioned earlier, the amount of radiation leaking from a microwave or dielectric heating system must be controlled, both to contain RF interference within acceptable limits and for personnel safety. Numerous devices, often called “chokes” or “attenuating tunnels,” are used for this purpose at conveyor openings, around doors and windows, at seams, and the like. Good engineering design should make it possible to keep leakage radiation well below the limits and guidelines set by the various controlling governmental organizations.

12.4.3.3 Protective Devices

Several protective devices are used in microwave systems to prevent high levels of reflected microwave energy from damaging the magnetron or klystron. The simplest of these are thermal switches that sense overheating of the tube and shut off the power. These may not be sufficient to protect the tube, however. Another method is the use of directional power sensors that discriminate between forward and reflected power and can shut off the systems when the latter becomes excessive. By far, the most sophisticated system is the ferrite circulator or isolator, which by

influencing the magnetic field passes microwave energy only in the forward direction, causing the reflected power to be shunted off into a dummy load. This system is highly efficient and especially recommended for high-power applications.

12.5 INDUSTRIAL APPLICATIONS OF MICROWAVE AND DIELECTRIC DRYING

It has been estimated that in Western Europe and the United Kingdom, RF equipment with 30 MW is manufactured annually compared with about 2 MW of microwave industrial equipment [9]. Roughly, half of the RF is used for plastic welding, with the rest given over to diverse systems. The industrial applications of microwave and dielectric heating are many and varied. In some cases, the application is unique to one form of energy or the other; in other cases, either form may be used. Although it is not possible to give any hard and fast rules for selecting one over the other, there are some guidelines that may be followed [8].

12.5.1 GUIDELINES FOR THE SELECTION OF MICROWAVES OR DIELECTRICS

12.5.1.1 Size of Load

If the load is very large or very wide, dielectric heating may be preferred. The depth of penetration is also

directly proportional to wavelength, and in dielectric heating it is measured in meters; in microwaves it is measured in centimeters. On the other hand, if the piece is small, microwave heating is preferred.

12.5.1.2 Watt Density

If the watt density requirement is very high, microwaves may be preferred to avoid arcing and burning of the material. For example, a bulky product consisting of loosely packed particles with a loss factor less than 0.05 favors microwaves.

12.5.1.3 Power

If the power requirement is high, over 50 kW, economics favor the dielectric system.

12.5.1.4 Geometry

If the product has an irregular shape with no rectangular cross section, the multimode microwave cavity will provide more uniform heating.

12.5.1.5 System Compatibility

If the system requires the use of pneumatic or hydraulic presses, metal conveyors, metal dies, or tenter frames, dielectric heating may be the only choice.

12.5.1.6 Self-Regulation

If the load fluctuates rapidly or goes through drastic changes in dielectric constant or dissipation factor during its heat cycle, partially tuned electrodes and instantly variable power from dielectric heating are advantageous.

12.5.1.7 Self-Limiting

If the load has a low-frequency loss tangent greater than 1.0, it is more resistive than capacitive, and in many cases this element disappears as the load dries or cures. RF will heat the resistive elements, and it will be more self-limiting in these areas and will not overheat or overdry. However, if the load has a low-frequency dissipation factor of less than 0.5, then it is mostly capacitive, and reduction of the dielectric constant as it dries or cures will be the predominant change and will permit better self-limiting at microwave frequencies. For example, dielectric heating has very little leveling effect on the moisture content of paper below about 5% because it is self-limiting in that region, but microwave energy could be used to dry the paper to near zero moisture content.

12.5.2 DIELECTRIC DRYING SYSTEMS

There are numerous systems in the lumber, furniture, textile, paper, food, tire, and ceramic industries, to name but a few. A brief description of these follows. In only a few cases, such as the postdrying of crackers, cookies, and biscuits or the drying of foundry sand cores are there many systems utilizing the same basic equipment. In most cases, the systems are customized or "one of a kind" so the number of actual applications is much larger. The same holds true for microwave drying.

12.5.2.1 Lumber

Dielectric heating is used for both drying and gluing lumber. It is used in the manufacture of plywood for drying of the veneer in order to remove pockets of moisture and provide moisture leveling. Otherwise, during hot pressing, steam pockets would form and delaminate the board. It is also used to cure the glue in plywood, medium-density fiberboard, and particle board. In all these cases, the dielectric system can utilize the plates of the presses as electrodes to give fast, efficient heating. These systems range in output power from 250 to over 1500 kW.

The rapid drying of lumber is also made use of in the furniture industry; in which precut furniture parts may be dried in minutes rather than days or weeks and the shrinkage is well controlled. Dielectrics are also used to dry the glue. Golf clubs are also dried in this manner [10,11].

12.5.2.2 Textiles

A large number of dielectric drying systems are used in the textile industry for drying of textile packages, hanks, skeins, tops, and loose stock. Speed, prevention of surface overdry, and leveling effects are all benefits of this technique, which results in superior product quality. Another benefit is more even distribution of dyes due to the diffusion of water vapor rather than liquid water during drying. Textile systems are in the 50 to 100 kW range primarily, although some as large as 250 kW and as small as 3 kW have been built [12]. A recent RF application is for the drying of loose fibers, especially for high-grade animal hair such as cashmere in which a loss of quality is unacceptable. The lower the temperature of the fiber mass the better and, by drawing air heated by the waste heat from the RF tube through the bed, a low fiber temperature can be maintained [9].

12.5.2.3 Paper

Dielectric heating is used to dry printing inks, adhesives, and coating materials on paper, as well as to dry the paper itself. This has been combined with hot air, infrared, and other heating media to achieve optimum results. Some examples of commercial installations include the following [9]:

1. Business forms, in which the RF units are used in conjunction with both presses and collators, with energy savings up to 75%. Up to 14-part forms can be dried as quickly as 180 m/min.
2. Direct mail line speeds up to 10,000 sheets/h or 30 m/min for web-feed systems.
3. Envelopes, for which the selectivity of RF energy heats and dries only the adhesive lines, leaving the bulk of the paper cool and flat.
4. Book binding, in which case the RF dries the adhesive, usually low-cost polyvinyl acetate.
5. Varnishes and water-based coatings for book covers, record sleeves, confectionary boxes, publicity brochures, and more.
6. Film laminates, in which nontoxic water-based adhesives are used in conjunction with polyester and polypropylene films.

For papermaking, machines are in use with 50 to 500 kW of RF at 27.12 or 13.56 MHz. The RF is able to overcome the most common problems encountered at the dry end of the process: low efficiency due to moisture distribution through the paper thickness, uneven moisture distribution across the width of the web, temporary unevenness or streaks of moisture due to some failure of the equipment, and cyclic variations in the machine direction.

12.5.2.4 Automobile Tires

An unusual application is the drying of the latex coating on fiberglass for fiberglass cord automobile tires. The coating is necessary to prevent abrasion of the fiberglass. Air drying of the coating must be slow to prevent surface skinning, which results in rupture of the coating when the internal steam pressure becomes sufficiently large. The dielectric drying of the coating results in uniform moisture loss in as little as 2.5 s, thereby increasing the solids level of the coating from 25 to 98%. Superior bonding of the latex to the fiberglass results from this process as well. PPG industries utilize such equipment with a capacity of several megawatts [13].

12.5.2.5 Food

Dielectric drying has several uses in the food industry, including drying of breakfast cereal; postbake drying

of crackers, cookies, and biscuits; and postbake drying of dog biscuits. There are now several hundred such postbaking systems in operation in North America and Europe. These usually consist of a short postbaking conveyor, 3 to 4 m in length, immediately following the fuel-fired oven. Thus, the product may exit the baking oven at a much higher moisture content, which is then rapidly removed by the RF dryer. Oven speeds may thus be increased by 30 to 50% or more, yielding higher quality product through the avoidance of case hardening [9].

12.5.2.6 Ceramics

Many drying systems of hundreds of kilowatts in power have been installed in the last 10 years for drying of ceramic monoliths, which must be done to permit firing of the ceramic. Because ceramics are good insulators, normal drying times are 24 h or more at high oven temperatures. The dielectric drying system accomplishes this in 20 to 22 min, which results in substantial energy savings as well as high product quality due to the uniformity of drying.

12.5.3 MICROWAVE DRYING SYSTEMS

Numerous industries use microwaves for drying, many of which are the same as those that use dielectric drying. However, there are several unique drying systems.

12.5.3.1 Food

Microwaves are used to dry pasta products, and there are over a dozen operational industrial systems. The systems utilize microwaves and hot air of controlled humidity to dry pasta and macaroni products in less than 1 h instead of the conventional 8-h drying time. These systems handle approximately 3000 lb of product per hour with 60 kW of microwave energy at 915 MHz. These systems offer substantial savings in energy, operation, and maintenance. They also provide bacteriologically more acceptable product, with reductions in microbial contamination and insect infestation [14].

Other food industry drying applications include drying of onions, seaweed, and potato chips [15]. The drying of onions is particularly interesting in providing substantial benefits in terms of moisture leveling, a 30% reduction in energy costs in the final drying, and a reduction in bacterial count of 90%. Here, hot air reduces the moisture level from 80 to 10% and the microwaves from 10 to 5% [16]. This is an ideal example of combining the two forms of energy in the most economical form.

12.5.3.2 Lumber

An unusual microwave application was in the drying of lumber for the manufacture of baseball bats from wood from the tanoak tree. This wood normally takes up to 2 years to dry. A microwave system at 2450 MHz heated the wood for 4 h, after which it was allowed to dry for another 2 weeks at ambient conditions. This was done on precut billets, which were later shaped into the bats, but is no longer used [17].

12.5.3.3 Laboratory Analysis

Several microwave systems have been developed for analytical laboratory drying to determine solids and moisture content. These systems have great advantages of speed with good precision and accuracy. It is often possible to do a complete moisture determination in 2 or 3 min that might otherwise take several hours.

12.5.3.4 Microwave Freeze Drying and Vacuum Drying

There has been great interest for many years in the possibility of utilizing microwaves for freeze drying. A problem in freeze drying is that, as the moisture front recedes, the product becomes harder to dry because of the reduced thermal conductivity of the material. Microwave radiation could be ideal to provide the required heat to the receding moisture. Unfortunately, at the high vacuum pressures involved in freeze drying there is increased opportunity for ionization of the gases, causing plasma discharge that can burn the product. There is indication that these and other problems have been overcome, and commercial microwave freeze drying systems are now feasible and may, in fact, already be operational for coffee [18].

The microwave vacuum dryer offers an interesting alternative to freeze drying, and several systems are in commercial production, manufacturing fruit juice concentrates, tea powder, and enzymes. Pilot-plant tests have also been successfully performed for drying such vegetables as mushrooms, onions, and asparagus. Still another pilot system is used for the drying of soya beans. The operational cost of microwave vacuum drying is said to be midway between spray and freeze drying [19].

In these systems, the material, often in paste form, is spread on a conveyor belt and passed through the specially built tunnel at a vacuum of 1 to 20 torr. This causes formation of a foam that, when dried, has excellent rehydration properties. An advantage of this method is that it allows materials of much higher solids

content than in spray and freeze drying [19], which reduces the cost since less energy must be expended.

There has also been great interest in a new system aimed at drying grain with a combination of microwaves and vacuum. By pulling a vacuum of about 20 torr, moisture in the grain can be evaporated at approximately 125°F rather than the 200°F air temperature currently used [20].

A recent overall review of microwave applications in the food industry covers these and other systems and discusses the industry's problems in adopting microwave technology [21].

12.5.3.5 Pharmaceuticals

The pharmaceutical industry has become very interested in microwave vacuum drying, particularly for the manufacture of tablet granulations [22]. These are blends that are then formed into tablets. During the course of manufacture, they may be mixed with water, ethanol, or acetone and must, subsequently, be dried. These microwave systems are gaining use and may combine mixing, granulating, lubricating, and dry sizing in a single step. Systems as large as 1200 L, employing 36 kW of microwave power, are in use. They demonstrate advantages in operator safety, cleaning, pollution control, and energy savings at costs often comparable to conventional systems.

12.5.3.6 Industrial Coating

There are a number of microwave drying systems for drying coatings on plastics and paper. (Included in these is the drying of silver halide on photographic film.) These combine high-speed drying with moisture leveling effects for high efficiency.

12.5.3.7 Ceramics

The ceramic industry has, for many years, examined the use of microwaves for drying purposes. Today, several uses are operating successfully. One such system, MCB Ceramics in Toronto, Canada, uses microwaves at two stages to replace a slow, hand-operated batch system with a continuous process. A 27-kW microwave oven is used to speed up the initial drying in the mold to 20 min from its previous 1 h, during which the microwaves are applied for only 2 min. The final drying used to take 24 h, but now is done with microwaves in only 8 min, after which the piece is glazed and fired. The process is used to produce small bathroom accessories such as towel bar holders and soap dishes [23].

In another process, ceramic filters, which are used to clean the slag in foundries before pouring the liquid

metals, are uniquely produced with microwaves. These filters, which may be as large as 12-in.² and 2-in. thick, are made by coating, both internally and externally, an expanded polyurethane or rubber foam with a ceramic slurry. This is then dried in about 25 min for even heating, and then the filter is placed into a kiln that burns away the foam leaving the porous ceramic structure (Krieger, 1994, private communication).

12.5.3.8 Casting Molds

The use of microwaves in the foundry industry for drying and polymerizing the sand molds used for casting is very important. This allows the complete recovery of the sand and provides a great increase in speed in making the mold, which otherwise must be slowly dried and cured with hot air. An example of this application is the manufacture of the internal castings for automobile engines. Many RF systems are also used for this process.

A new method of casting, the lost foam process, makes excellent use of the unique heating mechanisms of microwaves. Intricate castings, such as automobile engines and marine parts, are made of polystyrene foam and coated with a ceramic that must then be dried. The foam is an excellent insulator so hot air drying may take many hours, whereas microwave alone or in combination with hot air shortens the time dramatically. The system for automobile engines for the Ford Motor Company utilizes foam clusters hanging from a monorail traveling through a microwave oven 45 ft × 25 ft × 7 ft and utilizes 48 kW of microwave power. In this case, the clusters are first dried with warm air for about 70 min and then finish dried with microwaves for 20 min, a much shorter and efficient process than the 4 h required if only hot air is used.

The demands on this process are very stringent because the ceramic coating must be 100% dry, have a smooth, evenly coated exterior and interior, have no brittleness nor be overdried and browned, and have no flaking or separation of the coating. The equipment must also deal with varying process rates and so the number of units in the oven may vary from full capacity to none and with a large opening at the entrance and exit with no significant leakage so as to maintain personnel safety. The system has been operating flawlessly since 1985.

12.5.4 CRITERIA FOR SUCCESSFUL MICROWAVE AND DIELECTRIC DRYING SYSTEMS

There are several criteria for successful microwave and dielectric drying systems. Cost is reduced. This is often a major factor. Cost savings may be realized

through energy savings, increased throughput, labor reduction, reduction in heat load in the plant, speedup of the process, operational efficiencies, and reduced maintenance costs.

Quality is improved. Two examples are the drying of the latex-coated fiberglass cord and the drying of onions. In the first, there is prevention of rupture of the coating and, in the second, a reduction of bacterial contamination.

Yield is higher. The avoidance of high surface temperatures prevents overheating of the material dried and may lead to a lower level of rejects. The near instantaneous control of temperature in these systems allows better control of drying to closer tolerances. Moisture leveling effects also avoid over- or underdrying of products.

Product cannot be produced by any other way. Again, the careful control of temperature combined with the unique manner in which microwave and dielectric energy couple into materials allows the drying of extremely thermolabile materials with no damage to the product. Occasionally, a unique beneficial effect may be obtained, such as the slight puffing of the pasta noodles when they are microwave dried. This allows them to be cooked more quickly.

It is usual that two or more of these attributes may be combined. However, the bottom line is economics, and if a process does not produce a sufficient return on investment, it will not meet with success.

12.6 ECONOMICS OF MICROWAVE AND DIELECTRIC DRYING SYSTEMS

In analyzing the economics of a microwave or dielectric drying system, the costs can be divided into capital and operating costs, and this latter may be further broken down to tube replacement, general maintenance, energy, and floor space costs. We could also add costs for cooling water for the tubes or for materials specific to a system; however, we will concentrate on only the following: capital equipment cost, tube replacement cost, and energy costs.

12.6.1 CAPITAL EQUIPMENT COSTS

Capital equipment costs refer to the cost of the equipment and is sometimes a bit difficult to define. For example, a microwave drying system may increase the throughput of a product dramatically and thereby necessitate the purchase of additional packaging equipment, conveyors, feed systems, and the like. In that case, the overall capital outlay would be much higher than for the microwave dryer alone.

Another problem in trying to compare the costs of microwave and dielectric dryers is that the latter

often include hydraulic or pneumatic presses, materials-handling devices, and the like, built into the system, whereas this is unusual for a microwave system [8]. Also, suppliers express their costs in many different ways.

In general, though, we can limit ourselves to looking only at the basic drying system, consisting of the generator, tube, applicator, control system, and conveyor, if there is one. In that case, some rough figures are available. Dielectric systems are usually the less costly of the two, especially at high power output, and recent estimates given by Wilson [24] of 3500 to 10,000 per kilowatt, with the higher cost associated with lower power equipment. Microwave systems vary between 7000 and 10,000 per kilowatt, again with the lower cost associated with higher power equipment. Higher costs are also related to higher degrees of sophistication and automation of these systems. Therefore, for drying systems, by estimating that 1 kW of microwave or dielectric energy will remove 3.0 lb of water per hour, we begin to get a feeling for the economic feasibility of applying these energy systems to a specific drying process. If it appears that over 100 kW is required to remove the water, then the capital cost may be prohibitive, particularly for the microwave system. However, it is imperative to remember that it is unlikely that all the water will be removed by microwaves, but they should be employed discretely along with a conventional form of heat energy. In that case, they may only be required to remove the last few percentages of moisture from a material, at a much reduced capital cost.

As an example, consider the drying of bread crumbs from 27 to 5% moisture, at a rate of 1000 lb of wet bread crumbs per hour. In such a system, it would be necessary to evaporate 231.6 lb of water per hour (1000 lb of bread crumbs contain 270 lb of water, but when dried to 5% contain only 38.4 lb of water). This would require about 77 kW to dry ($231.6/3.0$), plus an additional 20 kW to heat to the drying temperature, for 97 kW, neglecting all heat losses. If we assume this system to have a coupling efficiency of 75% (that is, the efficiency of coupling microwaves into the product), then a system of 130 kW is required, which would cost approximately one million dollars. On the other hand, if a conventional hot air dryer is used to reduce the moisture from 27 to 12% and the microwave dryer to finish drying it to 5%, then the amount of water to be evaporated by the microwave system is only 61.6 lb, which requires 20.4 kW ($61.1/3.0$). Since the product is already hot, we need to only increase the output to 27 kW to account for the 75% coupling efficiency ($20.4/0.75$). Such a system would cost of the order of 200,000 plus the cost of the less expensive conventional dryer.

Note, however, that there are times when such a microwave or dielectric dryer can be less costly than the conventional, especially when drastic improvements are made in the drying rate such that 25% or less time is required to dry a product.

12.6.2 TUBE REPLACEMENT COSTS

The cost of tubes varies greatly, depending upon the output power. The least expensive tubes by far are the microwave oven tubes, and the output power of which is of the order of 1000 W and may be purchased, in quantity, for under \$15 each. However, higher power tubes are far more expensive. The approximate replacement costs and the approximate tube lives are shown in Table 12.7.

12.6.3 ENERGY COST

There was a time when electrically powered systems were considered too costly to operate for high-power applications. However, today, with the rising costs of oil and gas, this is no longer necessarily true, especially when far greater heating and drying efficiencies are possible with these systems. It is not practical to try to compare gas, oil, and electricity prices at this point as these are so variable. However, we can look at the conversion efficiencies of the various systems. In general, the conversion efficiency of electricity to microwave energy is estimated to be of the order of 45 to 50%, which includes losses in conversion from ac to dc (about 4%), from dc to microwaves (about 40%), and waveguide and applicator losses (about 10%). For dielectrics, there is an approximately 60% overall efficiency, based on a combined filament, control, and dc supply efficiency of 92%, times a tube efficiency of 73%, times a circuit efficiency of 90% [8]. Thus the energy cost per hour may be calculated as

$$\begin{aligned} \text{Energy cost/hour} \\ = (\text{utility rate/kWh})(\text{kW of system})(\text{efficiency}) \end{aligned} \quad (12.9)$$

12.6.4 OTHER COSTS

Other costs may be considered, such as the cost for floor space for the system and the cost of maintenance. These may vary for individual situations and cannot be considered here. We may, however, consider the cooling water cost. Many of the tubes are water cooled, and the cost for the cooling water should be considered. The water requirements are directly proportional to the output power of the tube; for example, a 6-kW magnetron at 2450 MHz

TABLE 12.7
Tubes: of Approximate Replacement Costs

Size (kW)	Type	Frequency (MHz)	Life (h) ^a	Actual Cost (\$)	(\$ Per Hour of Operation Per kW)
<i>Microwave tubes</i>					
2.5	Magnetron	2450	4,000	1,500	0.15
6.0	Magnetron	2450	6,000	3,300	0.09
50	Klystron	2450	25,000	69,000	0.06
30	Magnetron	915	8–10,000	5,300	0.02
50	Magnetron	915	6–8,000	6,000	0.02
<i>Dielectric tubes</i>					
5	Up to 100	5–10,000	1,600	0.04	
10	Up to 100	5–10,000	2,200	0.02	
50	Up to 100	5–10,000	3,600	0.01	
100	Up to 30	5–10,000	8,400	0.01	
200	Up to 30	5–10,000	16,800	0.01	

^aApproximate.

requires 1.5 gal/min; a 30-kW magnetron at 915 MHz requires 5 gal/min.

An article by Jones and Metaxas [25] describes a case study on providing additional drying capacity in a papermaking operation. This paper compares the present conventional drying system to microwave, RF, and infrared systems and provides the analysis that led to the final decision. A detailed cost-benefit analysis is provided and is a good summary of the procedures that should be followed in such circumstances.

Another good summary of the economic factors to be considered in choosing a microwave or dielectric drying system is given in an early paper by Jolly [26]. Here a detailed comparison is given for the economic considerations that enter into the choice of a new microwave system over the standard conventional procedure.

12.7 CONCLUSION

The application of microwave and dielectric heating to industrial drying systems is of increasing interest, particularly because of the increased energy and operational efficiencies they afford. Their unique heating means provide benefits not obtainable from other, more conventional methods. However, their high capital costs generally require that they be used judiciously in conjunction with more conventional heat forms. Their unique properties also require the systems be designed by engineers thoroughly familiar with the art and science of microwave and dielectric heating applications.

Considerable research and development has been devoted to the understanding of the fundamentals of

dielectric drying and to the development of industrial applications in the past decade. For more recent, authoritative, in-depth reviews of this subject the interested reader is referred to the papers by Schmidt et al. [27] and Jones [28]. Turner and Rudolph [29] have presented a model for combined microwave and convective drying. Cohen et al. [30] have demonstrated the benefits of microwave-assisted freeze drying of peas. A number of other applications can be found in the proceedings of the biennial International Drying Symposium (IDS) series. Further, a special issue of *Drying Technology—An International Journal* dealt with all aspects of dielectric drying [31]. Two recent articles discuss moisture transport and the drying by means of microwaves [32,33].

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13 Solar Drying

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13.1 INTRODUCTION

Open-air sun drying has been used since time immemorial to dry plants, seeds, fruits, meat, fish, wood, and other agricultural or forest products as a means of preservation. However, for large-scale production the limitations of open-air drying are well known. Among these are high labor costs, large area requirement, lack of ability to control the drying process, possible degradation due to biochemical or microbiological reactions, insect infestation, and so on. In order to benefit from the free and renewable energy source provided by the sun several attempts have been made in recent years to develop solar drying mainly for preserving agricultural and forest products.

13.2 ASPECTS AND LIMITATIONS OF SOLAR DRYING

13.2.1 GENERAL CONSIDERATIONS

Among the advantages of solar drying one may cite a free, nonpolluting, renewable, abundant energy source that cannot be monopolized [1]. At the same time, in using solar radiation for planned drying, several difficulties must be overcome. There is the very basic problem of the periodic character of solar radiation, a problem that gave rise to the idea of storing part of the energy gained during radiation periods. This difficulty can be eliminated, aside from employing heat

storage devices, only with the use of an auxiliary energy source. Even the radiation periods may produce certain difficulties. First, the intensity of incident radiation is a function of time. This is a circumstance that demands adequate control strategy and the means necessary for the control. Another problem is caused by the low energy density of solar radiation, which requires the use of large energy-collecting surfaces (collectors).

Thus, the nature of solar radiation has innate problems that require means (heat stores, auxiliary energy source, control system, and large-surface solar collectors) for their solution, and so the investment costs are considerable. Obviously, a prerequisite to utilizing solar energy is economics and the need to achieve an acceptable rate of return.

An examination of the technoeconomics of solar drying has led to the knowledge of the main factors, their roles, and influencing mechanisms. The first obvious discovery was that solar energy can be economically used for drying only if the purpose can be coordinated with the specific characteristics of solar radiation. Thus, geographic circumstances deciding the number of sunny days yearly and the incident radiation intensity give different energy gain in various areas of the Earth. The relatively small energy flux density of solar radiation implies that it is particularly suited to drying processes with small energy demands.

Seasonal changes of solar radiation suggest the use of solar drying in the maximum radiation intensity season: e.g., part of the agricultural products should

be dried during this period [202]. Matching of the drying process and the specific characteristics of solar radiation is also important in governing the investment costs. Because of the small flux density of solar radiation, a high-temperature drying medium can only be produced with concentrating collectors. Such collectors are generally very expensive. Cheaper, flat-plate collectors, on the other hand, can be applied only for producing a moderate-temperature medium (usually under 60°C), and their efficiency improves with a decrease in operation temperature. So dryers with flat-plate collectors can be used for products requiring low-temperature drying.

One way to reduce the costs of solar collectors is to strive for cheap and simple construction. This can mean a decrease in operation life and efficiency so that the task must be handled as an optimization problem. Another possibility is multipurpose construction, for instance, building the collector into the roof structure as an integrated part.

At this point technical development can proceed in two directions: simple, low-power, short-life, and comparatively low-efficiency dryers in one direction, and high-efficiency, high-power, long-life, expensive dryers in the other direction. The latter are characterized not only by an integrated structure but also by integration in an energy system involving processes other than drying. The aim is twofold: coupling a solar energy dryer to a farm's energy system gives the possibility of using solar collectors practically throughout the whole year, for example to produce hot water when the dryer is not in use; also, the hot water tank of the farm can be used as a heat store of the solar system.

13.2.2 ROLE AND IMPORTANCE OF SOLAR DRYING IN THE DEVELOPING COUNTRIES

Due to the lack of adequate preservation methods, direct open-air drying is still a widely used means of food preservation in most parts of the developing world [147,152]. This traditional practice has its inherent disadvantages:

- Damage to the crop by rodents, birds, and animals
- Degradation through exposure to direct irradiation of the sun and to rain, storm, and dew
- Contamination by dirt, dust, wind-blown debris, and environmental pollution
- Splitting of the grain bleaching and loss of germination capability due to overdrying
- Insect infestation
- Growth of microorganisms
- Additional losses during storage due to insufficient or nonuniform drying

Postharvest losses can be estimated at more than 30% [148,149] and it could be reduced to a great extent by adequate drying of crops [150].

There are two possible ways for the proper preservation by drying: by using fossil fuels and by using solar energy. Not considering the disadvantageous environmental pollution effects caused by the CO₂, SO₂, and NO_x emission, the use of fossil fuel fired or electrically powered dryers is limited and inappropriate for most of the farmers in developing countries. The main reasons are as follows:

- Expensive investment and high energy costs
- Lack of skilled personnel for operation and maintenance
- Conventional sources of energy are either unavailable or unreliable [151]

During the last decades, several developing countries have started to change their energy policies toward further reduction of petroleum import and to alter their energy use toward the utilization of renewable energies [152].

With very few exceptions, the developing countries are situated in climatic zones of the world where the insolation is considerably higher than the world average of 3.82 kWh/m² day [153]. In Table 13.1 daily average horizontal insolation data and sunshine hours are given for some developing countries [152,154].

TABLE 13.1
Total Horizontal Solar Insolation and Sunshine Hours for Some Developing Countries

Country	Average Insolation (kWh/m ² day)	Sunshine Hours (h/d)
Cameroon	3.8–5.5	4.5–8.0
Egypt (Cairo)	6	9.6
Guatemala	5–5.3	—
India	5.8	8–10
Indonesia	4.24	—
Kenya	5.25–5.6	6–7
Malaysia	4.41	—
Mali	4.34	8.4
Mauritius	4.5	7
Mexico (Jalapa, Veracruz)	4.65	—
Nicaragua	5.43	—
Nigeria	3.8–7.15	5–7
Papua New Guinea	4.6–9.6	4.5–8
Philippines (Metro Manila)	4.55	—
Sierra Leone	3.4–5.3	3–7.5
Thailand	4.25–5.66	—
Togo	4.4	5.5–7.2

An alternative to traditional drying techniques and a contribution toward the solution of the open-air drying problems is the use of solar drying. The main reasons are as follows [216]:

1. Solar drying provides the desired reduction of losses together with improved quality of the dried products.
2. The time of drying can be significantly reduced.
3. The harvesting period can be shortened, which enables the soil to be prepared for the cultivation of another crop.
4. The drying season can be lengthened by successive harvests and by using solar dryers in which various types of products can be preserved.
5. Farmers may have a greater income by the production of marketable crops.
6. The additional costs involved in installing solar dryers can be returned by the increased profits.

Accordingly, the availability of solar energy and the operational marketing and economy reasons offer a good opportunity for using solar drying all over the world. A great number of successful practical applications have already been reported [152,155–174,206,211,215,217].

13.3 CONSTRUCTION PRINCIPLES OF SOLAR DRYERS

13.3.1 MAIN PARTS OF SOLAR DRYERS

Solar dryers have the following main parts:

1. Drying space, where the material to be dried is placed and where the drying takes place
2. Collector to convert solar radiation into heat
3. Auxiliary energy source (optional)
4. Heat transfer equipment for transferring heat to the drying air or to the material
5. Means for keeping the drying air in flow
6. Heat storage unit (optional)
7. Measuring and control equipment (optional)
8. Ducts, pipes, and other appliances

13.3.2 CLASSIFICATION OF SOLAR DRYERS

The structure of solar dryers is adjusted to the quantity, character, and designation of the material to be dried as well as to the energy sources used. Accordingly, a great variety of solar dryers have been developed and are in use. The following classification suggests three main groups for solar dryers on the basis of the energy sources used [175]:

1. *Solar natural dryers* using ambient energy sources only
2. *Semiartificial solar dryers* with a fan driven by an electric motor for keeping a continuous air flow through the drying space
3. *Solar-assisted artificial dryers* able to operate by using a conventional (auxiliary) energy source if needed

13.3.3 SOLAR NATURAL DRYERS

In the main group of solar natural dryers two subgroups are included: the subgroup of the passive, natural convection solar dryers (cabinet, tent type, greenhouse type, chimney-type dryers) and the subgroup of active, partly forced convection solar dryers having a fan driven by electric energy converted by photovoltaic solar cells [226] or driven by a small wind turbine.

13.3.3.1 Cabinet Dryers

The simplest solar dryers are the cabinet dryers (Figure 13.1). Their main characteristic is that the heat needed for drying gets into the material through direct radiation and through a south-oriented, transparent (glass or foil) wall 1. Other walls of the dryer are opaque and well insulated. The drying material 2 is spread in a thin layer on a tray 3. The bottom plate of the tray is perforated. Air flows through the holes by natural convection through the material and finally leaves through the upper part of the cabinet [2,3]. The design of the dryer is simple, and its cost is low. It is suitable for drying small quantities (10–20 kg) of granular materials (e.g., for individual farmers). The products dried in cabinet dryers are mainly agricultural products—vegetables, fruits, spices, and herbs.

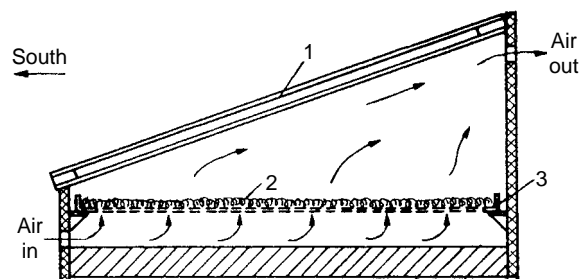


FIGURE 13.1 Structure of a cabinet dryer. (From Special issue, *Sunworld*, 4, 179, 1980; Garg, H.P., in *Proceedings of the Third International Drying Symposium* (J.C. Ashworth, Ed.), Drying Research Limited, Wolverhampton, England, 1982, p. 353.)

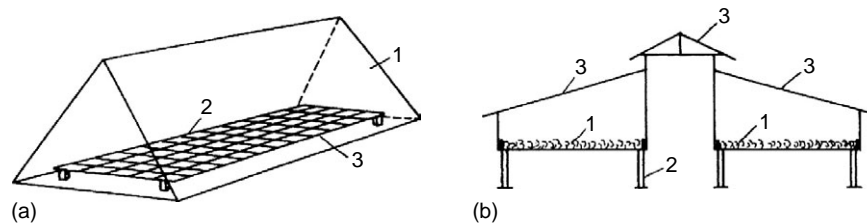


FIGURE 13.2 Cabinet dryer variations: (a) tent-type dryer; (b) its terrace-type solution. (From Special issue, *Sunworld*, 4, 179, 1980.)

Drying of the material can be made more even by periodic turning over of the material. It is employed chiefly in tropical countries, but during the warm months it can be used in the temperate climates as well. The usual size of the drying area is 1–2 m² (see also Figure 13.47).

A variation of the cabinet dryer is the tent dryer, which consists of a triangular framework covered with a thin sheet (Figure 13.2a). The south-oriented front wall 1 is transparent; the back wall is covered by a black sheet. The material is spread on a tall tray made of netting or wire mesh.

Another type of tent dryer has its roof covered by polyethylene sheet. The drying material is spread over a concrete floor (e.g., coffee beans) [2]. Another type of the tent dryer is the terrace dryer. Its cross-section is sketched in Figure 13.2b [2]. The drying shelves 1 stand on posts 2; the roof and the front wall are covered with a polyethylene foil 3. Certain types of terrace dryers are made with roofs that open so that under favorable weather conditions the drying material is exposed to direct radiation. Constructing tent and terrace dryers is cheap and simple. They are widely used for drying coffee. In Colombia, about 70% of the coffee beans are dried in such dryers [2].

13.3.3.2 Natural Convection, Static Bed, or Shelf-Type Dryers

The capacity per unit area of cabinet dryers is limited by two conditions: need for direct radiation on the drying material and small airflow rate. To dry larger quantities of material, the basic area of the dryer has to be increased. To avoid this problem it is preferable to place the material in several independent layers; the necessary heat transfer is thus accomplished by convection. The increase in the mass flow rate of air can be achieved by increasing the effects that produce natural convection. These effects must also be increased if the air is to be circulated through a material laid in several layers one over the other, or through a thick layer, as in the case of the static bed type. To keep up the natural pressure difference without using a ventilator (for instance, in a field), the “chimney effect” must be exploited. For this purpose the vertical flow of hot air in the dryer must be increased.

In Figure 13.3 a scheme of the so-called shelf dryer can be seen [4]. The material to be dried is placed on perforated shelves 1 built one above the other. The front wall of the case faces south, its top and sides 2 are covered by transparent walls (glass or sheet),

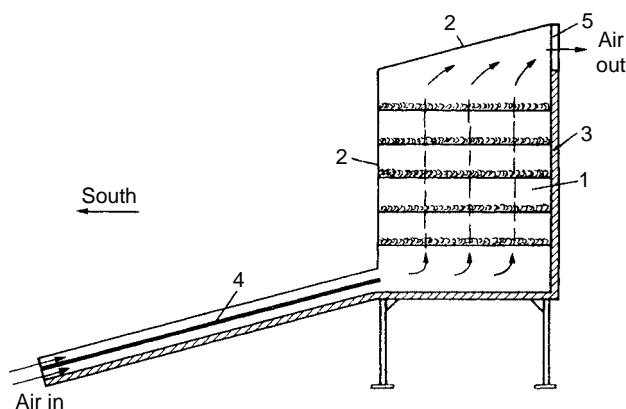


FIGURE 13.3 Shelf-type dryer with separate collector. (From Wibulswas, P. and Niyomkarn, C., *Reg. Workshop on Solar Drying*, CNED-UNESCO, Manila, 1980, p. 1.)

and the back wall 3 is heat insulated. The back wall and the floor are covered with a coating of black paint. The ambient air is warmed in a flat-plate collector 4 joined to the bottom of the case, and it flows up to the space under the lowest shelf. Moist air exits to the open through the upper opening of the case 5. In the scheme shown in Figure 13.3 the chimney effect is ensured by the increased height (approximately 1 m). Test measurements have shown that over a period of 5 y the best results were obtained with the aid of a glass-covered absorber plate placed in the middle of the air opening. Costs of utilized energy (Thailand climate, January–April) amount to about U.S. \$.03/kWh. The experiments indicated that separation of the collector is only justified with a high-efficiency collector. The dryer is suitable for drying fruit and vegetable goods. (For a theoretical analysis of shelf-type solar dryers, see Ref. [5].)

For large amounts of material to be dried the airflow rate through the dryer should be increased. In case of static bed-type solar dryers an appropriately high chimney has to be connected to the dryer housing [204]. Figure 13.4 gives the cross-section of a chimney-type dryer designed and built for drying 1000-kg rice [2]. Rice is placed in a static bed I in a 0.1-m thick layer. The collector 2 consists of a plastic covering and roasted rice shell, the latter playing the role of absorber. The front surface 3 over the layer of rice is also transparent. The wall of the chimney 5 is made of black plastic foil. The framework of the dryer is wood and wire; manufacture is inexpensive and simple. The air needed for drying amounts to $5.7 \text{ m}^3/\text{min}$ per m^3 rice. The chimney is 5 m high. Drying is not uniform, so the rice in the static bed must be turned over at intervals. The duration of drying is 3–4 d in the case of $15 \text{ MJ}/\text{m}^2$, day mean global sun radiation, and 23 m^2 collector surface. With the application of a larger (36 m^2) collector surface, drying time can be reduced to 1–2 d in good weather. As a rule of thumb, the solar collector sur-

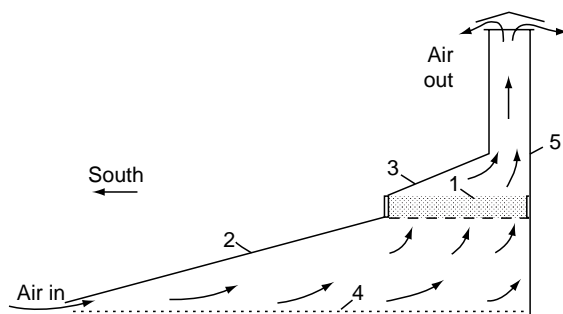


FIGURE 13.4 Static bed-type solar dryer with chimney. (From Special issue, *Sunworld*, 4, 179, 1980.)

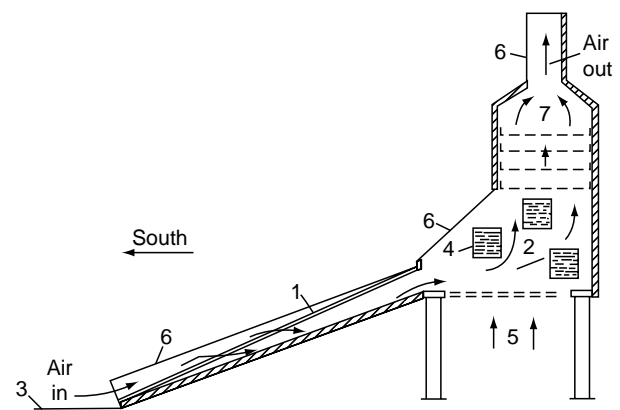


FIGURE 13.5 Tray- and chimney-type solar dryer with heat storage. (From Puiggali, J.R. and Lara, M.A., in *Proceedings of the Third International Drying Symposium* (J.C. Ashworth, Ed.), Drying Research Limited, Wolverhampton, England, 1982, p. 390.)

face must be approximately three times the surface of the bed.

Preliminary steps [6] have been made for the development of heat-storing chimney-type dryers with the purpose of extending the drying process over radiation-free periods. A schematic of the dryer is shown in Figure 13.5. Air gets through the collector 1 to the heat-storing space 2. The collector is foil covered; its angle of inclination can be adjusted to a small degree. A reflection panel 3 is placed near the air entry, which serves for warming the entering air to a small degree. Water-filled vessels 4 serve to store heat. The walls of the heat-storing space are insulated by reflecting panels that can be turned down for the night. During night operation the outside air can be let into the heat-storing bodies through openings 5 made in the bottom of the heat-storing space. The front and sidewalls of the heat-storing space are covered with transparent foils like the southern sidewall of the chimney 6. Its back wall and bottom plate, as well as the drying space, are well insulated. In the drying space 7, the drying material is spread on trays with perforated bottoms. Test measurements indicate about 10% drying efficiency as related to the input solar energy (see also Figure 13.48).

13.3.4 SEMIARTIFICIAL SOLAR DRYERS

The greatest advantage of chimney-equipped natural convection dryers is that no auxiliary energy source is needed and thus they can be operated far from populated areas. The disadvantage is that the height of inexpensive-finish chimneys (without special stiffeners and foundation) is limited mainly because of the increased wind loading. Limitation of the chimney

height means a limitation on the hydrostatic pressure difference and also that of maintainable airflow rate. Another disadvantage of natural convection dryers is that the air entering the collector flows through the drying space and then into the atmosphere. During drying the temperature of the material approaches the dry bulb temperature and the enthalpy increase of the air taken up from the collector is used for drying in decreasing quantity while an ever-increasing part leaks into the atmosphere through the chimney as exit heat loss.

In view of this, solar dryer variations have been developed in which a small power ventilator is fitted for maintaining airflow whereas recirculation is controlled by simple flaps built at suitable spots to improve thermal efficiency. The construction of such dryers is relatively simple and inexpensive. In a high-performance tent-type dryer variation designed for drying 3–4 tons of peanuts [7], the material to be dried is placed and dried in a drying drum located in a closed chamber with perforated walls, which plays the role of a solar collector as well. A ventilator delivers outside air into the chamber and as the air is warmed it comes into the drum through its perforated mantle and from there into the open air; a part of the exit air may be recirculated if desired. Far from the grid (e.g., in rural areas), the electricity for driving the fan can be produced by photovoltaic modules mounted on the covering surface of the collector [208,227].

An effective solution of semiartificial solar dryers is the directly irradiated, foil-covered solar tunnel dryer with integrated collector section [183]. Scheme and operation of that dryer is given in Figure 13.46 (see Section 13.7.3).

13.3.4.1 Room Dryers

Figure 13.6 shows the schematic structure of a solar timber dryer [8,9]. A stack with air clearances 1 is made of 30–65-mm wide coniferous and oak timber in the inner space of a building. The northern wall of the building 2 is well insulated; its roof 3, southern wall 4, and sidewalls are made of special two-layer transparent synthetic plates. The dryer is built on a concrete base 5. Solar radiation coming in through the transparent walls warms the black-painted aluminum absorber 6. The airflow delivered by an axial ventilator 7 flows along the two sides of absorber 6; one part of it enters the stack at the backside, the other from the bottom. The adjustable angle of inclination of the upper (top, roof) part of the absorber 6 makes it possible to control the quantity of air directly led into and circulated in the stack. The proportion of fresh and recirculated air can be changed by simple flap valves 8. The flow volume of the

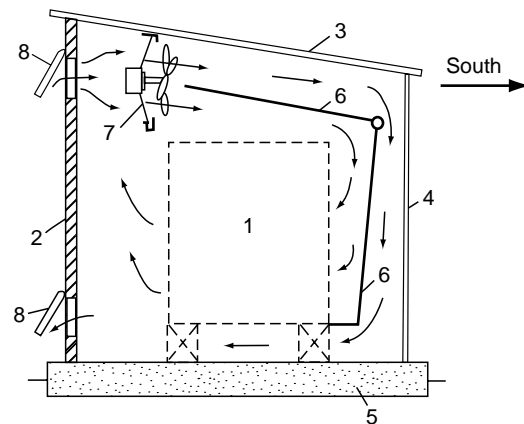


FIGURE 13.6 Forced convection solar dryer for timber. (From *Résumé de l'étude en cours on CTB sur l'utilisation de l'énergie solaire*, Centre Technique du Bois, Paris, 1978; Yang, K.C., *Forest Product J.*, 30, 37, 1980.)

ventilator is 2.5 m³/s, with 180 Pa. Depending on the width of the building, one or several ventilators may be used. (For example, two ventilators are needed for a 5.64 m wide building. The height of the southern wall is 2.50 m; the width of the building is 3.05 m. Stack volume is 5–9 m³. For tests made with a dryer of similar construction, see Refs. [10,11]; with rock heat storage placed on the floor, see Ref. [12].)

The application of forced airflow is necessary for drying products in static beds, which form a comparatively large flow resistance in the bed. Such products include grains and hay. One solution is to build a solar room dryer [13] as shown in Figure 13.7a. The grains to be dried are placed as a bed 1 on a perforated flooring. Collectors 2 are located on the southern wall and the roof of the building. The air warmed in the collectors is kept in circulation through channel 3 by fan 4 and ventilated through the bed across the lower distributing space 5. Wet air exits to the open air from the roof space through sidewall openings. The collector area was chosen to be 4 m² for 1 m³ wheat of 20–24% initial moisture content (double covering, 200 W/m² long-term mean collector power). Measurements showed that a maximum of 55°C entering air temperature could be reached without influencing germination ability. In the lower layers there is the possibility of overdrying, which may be avoided economically by employing separate heat storage.

The role of the dryer housing can be played by a grain bin [8,14]. In this case the collectors are integrated with the wall of the cylindrical- or square-shaped bin. Other variations use a separate plane collector system [15].

Figure 13.7b shows the schematic construction of a solar rough fodder dryer [16]. The material is placed in a static bed 1. The collector system 2 is placed, as in

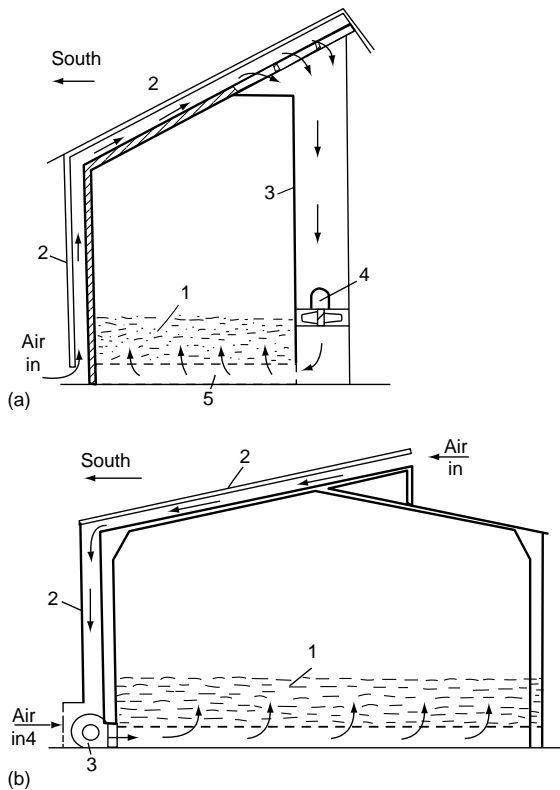


FIGURE 13.7 Solar room dryers: (a) room solar grain dryer. (From Wieneke, W., *Agricultural Mechanization in Asia, Autumn*, 11, 1980.) (b) Solar fodder dryer. (From Darnedde, W. and Peters, H., *Landtechnik*, 29, 1978.)

Figure 13.7a, on the southern wall and the roof, but with airflow in the opposite direction. The collecting channel is placed at the bottom, joined to the housing of a fan 3. The fan is able to draw in outside air directly 4. For bad weather there is the possibility of using an auxiliary energy source (e.g., gas-heated heat exchanger) on the suction side of the ventilator. (For further applications of collectors integrated into the roof, see Refs. [17,18].)

13.3.4.2 Solar Dryers with Physical Heat Storage

The application of heat storage in solar drying system is justified by three circumstances:

1. Drying period can be extended by the stored energy
2. The surplus energy appearing at the radiation peaks can be stored to avoid local overdrying
3. The temperature of the drying air can be controlled to avoid damage to material

In any case, when dimensioning the collector surface of the dryer with heat storage, attention must be paid to the fact that the energy getting into the storage unit forms a part of the energy gained by the collector system. Also, use of heat storage will necessarily involve a decrease in the temperature level of the energy obtained. In the case of directly radiated heat storage (formed, e.g., as the absorber of the collector) this effect is less important.

One pays for the advantage of using heat storage with higher investment and operating costs. Careful technoeconomic evaluation must be made before using solar energy storage in solar drying.

Natural or artificial materials may be employed for heat storage. Natural materials (water, pebble bed, and rock bed) are usually cheaper than synthetic materials (e.g., latent heat-storing salt solutions and adsorbents). Detailed discussion of heat stores is beyond the scope of this chapter.

Sensible heat storage of high capacity calls for water as the working medium (indirect heating system). Accordingly, the collectors are more expensive, and the application of a water–air heat exchanger also involves further cost (see Section 13.5.2).

In Figure 13.8 the construction of a solar dryer with water storage is shown [19]. The dryer is an indirect system. Pump 2 circulates the working medium of the collectors 1 along a pipe 3 and warms the fluid in a storage tank 4. The dryer uses outside air

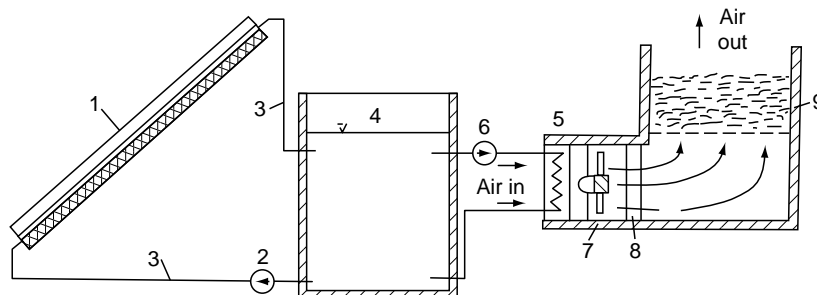


FIGURE 13.8 Solar dryer equipped with water-type heat storage. (From Auer, W.W., in *Drying '80* (A.S. Mujumdar, Ed.), Hemisphere, New York, 1980, p. 292.)

drawn by ventilator 7 and heat exchanger 5. The primary medium of the heat exchanger is the fluid from tank 4 circulated by pump 6. Air can be warmed to the necessary temperature by the heater 8. The material to be dried is placed in a static bed 9. Measurements have proved that 50–60% of the energy needed for drying can be gained from solar energy.

Figure 13.9 shows the arrangement of a solar dryer with rock-bed storage [20]. The dryer is a direct system; the collectors 1 are located on the ground and have an area of 193 m^2 . The air warmed in the collectors is forwarded into drying space 3 by a fan 2. The dryer has room for a maximum of 6.5 m^3 timber 4. In the upper roof space two ventilators are placed that are used for continuous circulation of the air. Vents 6 are placed in the sidewalls of the roof space for allowing inflow and outflow of air. The rock bed 7 is about 22 tons of 19-mm crushed basalt. The dryer operates as follows.

1. During the warming period, a fan 2 revolves, first slowly, then at full rotation; a damper 8 opens gradually; damper 9 is in the position marked in the figure by the dotted line. Air flows from the dryer space back into the collector.
2. During drying and charging of heat storage, damper 9 is in the position shown in the figure by a solid line, air flows from the dryer space into the rock bed and back to the collector.
3. During operation when there is no solar radiation, damper 9 is in the medium position, air flows from the dryer into the rock bed, is warmed, and flows again into the dryer. The operation of dampers 6 is controlled by the wet bulb temperature measured in the dryer.

The economic design of the rock-bed storage device is of great importance [213].

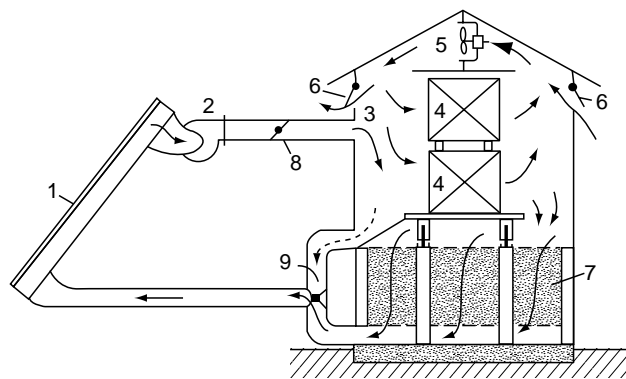


FIGURE 13.9 Solar timber dryer equipped with rock-bed heat storage. (From W. Read, R., Choda, A., and Cooper, P.I., *Solar Energy*, 15, 309, 1974.)

13.3.5 SOLAR-ASSISTED ARTIFICIAL DRYERS

13.3.5.1 Solar-Assisted Dryer for Seeds

In Figure 13.10 the scheme of a solar-assisted seed dryer is presented. Figure 13.10a shows the cross-section of the dryer with layer-type arrangement and Figure 13.10b a containerized construction for sensitive materials [185,197,199]. The drying space is divided into two cells 1, 2 for the better direction of the drying process (see also Section 13.7.3). Each cell has an individual fan 5 of two RPM stages. Fans are arranged in separate spaces 3. Among the two-fan spaces the space 4 of the auxiliary heater 6 operating with natural gas is situated (Figure 13.10c).

As solar energy converters, uncovered flat-plate air collectors 7 are used integrated into the roof structure of the building. Air ducts of the collector field are connected to a collecting–distributing air channel 10. By moving sliding plates 11 the collecting channel can be opened and connected to the fan spaces to divide the total preheated airflow, in a proper ratio, between the cells. For seed grains of small dimensions a layer-type static bed is preferred (Figure 13.10a). For seed grains of larger dimensions (e.g., beans) use of containers with perforated bottoms is recommended to avoid the possible damages during transportation and feeding in and out. Moist air leaves the drying cells through the openings 20.

The main technical data of the dryer are as follows:

Number of cells: 2

Effective surface area of the bed for one cell: 56 m^2

Dry mass of seed for one cell and seed grains of meadow grass: 5,600 kg

Mass flow rate of air of one fan: at RPM 1,090 per min, $41,000 \text{ m}^3/\text{h}$; at RPM 475 per min, $12,500 \text{ m}^3/\text{h}$

Surface of the collector field: 191 m^2

Average effectiveness of the collector: 0.3

Average temperature increase of the air preheated by solar energy (July, Hungary): at RPM 1090 per min, 2.9°C ; at RPM 475 per min, 9.86°C

Output of the auxiliary air heater: 93 kW
(medium-scale crop-dryer with unglazed collector is presented in Ref. [218]).

13.3.5.2 Solar-Assisted Dryer with Gravel-Bed Heat Storage

The construction of a high-performance raisin dryer is shown in Figure 13.11 [19] with rock-bed heat storage. The collector system 1 consists of 42.7-m long units with a surface area of 1812 m^2 , located on

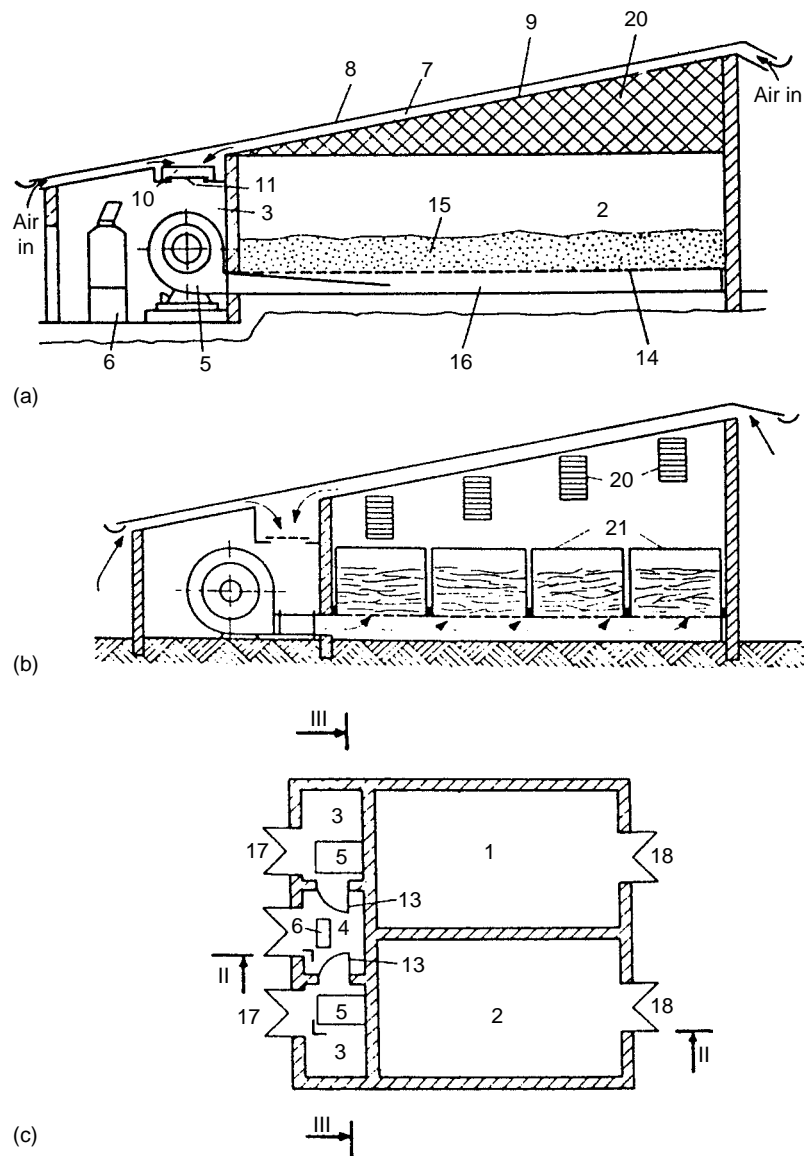


FIGURE 13.10 Solar-assisted dryer for seeds: (a) cross-section of the dryer in layer-type arrangement; (b) containerized construction; (c) ground plan of the dryer.

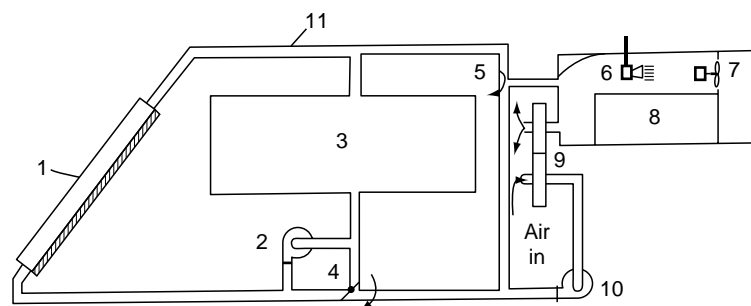


FIGURE 13.11 Solar raisin dryer with gravel-bed heat storage. (From Auer, W.W., in *Drying '80* (A.S. Mujumdar, Ed.), Hemisphere, New York, 1980, p. 292.)

the ground. Fresh air is drawn into the system through a heat recovery wheel 9 by ventilator 10, and with the damper 4 in the horizontal position it is sent to the collector. Air coming from the collectors through the collecting duct 11 arrives in space 6, where a gas burner heats it as needed. Ventilator 7 sends the warm air into a drying tunnel 8 and from there through the heat recovery wheel into the open air. When switching on ventilator 2 a part of the air flows through the rock pile storage 3. If the collector system is out of operation, air can be circulated into the drying space through the heat storage with damper 4 in the perpendicular position. Further, if damper 5 closes the upper duct, the dryer can operate with auxiliary energy as the only energy source. The solar energy system covers 69% of the energy needed for drying in a yearly 214-d sunny season in California. (For a similar solution for crop dehydration, see Ref. [21].)

13.3.5.3 Solar-Assisted Dryer Combined with Heat Pump and Heat Storage

Heat pumps are coolers (refrigerators) that raise the energy gained by cooling from a low-temperature energy carrier with the aid of further external (driving) energy to a higher temperature level and transfers it from there to an energy-carrying medium [22,23]. The term heat pump refers to the fact that both the cooling and the heating performance of the refrigerator are utilized.

Figure 13.12 shows the schematic arrangement of a solar dryer equipped with absorption heat pump and heat storage [19]. A part of the enthalpy of entering outside air 1 is used—interposing pump system 2—for evaporating sprayed water in an evaporator 3. The water vapor goes over to the brine sprayed into tank 4. Pump 5 feeds the brine through a regenerator heat exchanger 6 into a high-pressure boiler 7. Water in the boiler is distilled with the help of solar energy obtained in a collector 10 and stored in a water tank 11, and by using auxiliary energy A to the extent necessary the strong solution is led back into tank 4 through regenerator 6. The high-pressure water vapor condenses in condenser 8 and with the help of the pump heat exchanger system 9 warms the air of reduced moisture content, which is supplied to the dryer. The condensed high-pressure water flowing through an expansion valve E cools and arrives in evaporator 3. This system was originally designed for drying peanuts. A “hybrid” solar dryer with heat pump and photovoltaic modules has been constructed for drying vice [219].

13.3.5.4 Solar-Assisted Dryer Integrated into a Complex Energy System

One economic factor in the use of solar dryers is the amount of solar energy over the year and the yearly drying period. In any case, even in the drying season there are unavoidable breaks, and during the rest of the year the collector system cannot be used for

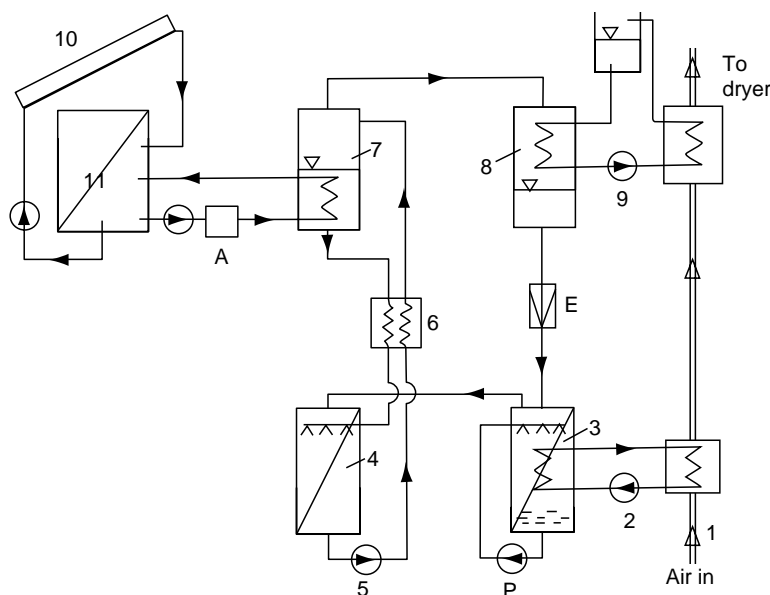


FIGURE 13.12 Solar dryer for peanuts equipped with absorption heat pump and heat storage. (From Auer, W.W., in *Drying '80* (A.S. Mujumdar, Ed.), Hemisphere, New York, 1980, p. 292.)

utilizing solar energy. For year-round utilization it is desirable to look for other possibilities for using solar heat. Such a possibility is given, for instance, by satisfying the hot water needs of a stock-breeding farm. With the integration of the solar system into the hot water system of the farm, investment costs can be saved. The storage tank of the hot water system can be used as water storage for the solar system. In addition, the complex system solution may permit increasing the efficiency of the solar system and thus the amount of solar energy obtainable. The efficiency of flat-plate collectors (for more details see Section 13.5.2) improves with decrease in operational temperature. So it is more advantageous if the collectors are used for warming cold water from wells or the water supply system than if the working medium returning from the fluid–air heat exchanger of the dryer is led into the collector at a temperature higher than the ambient.

Figure 13.13 shows the scheme of a solar alfalfa dryer joined to the hot water system of a stock-breeding farm [24,25,199]. The fluid medium collector system 5 built on top of the dryer building is connected to a closed circuit. The system can have different operating modes. When valves 2 and 3 are closed, the collector system works on the fluid–air heat exchanger 6 and serves dryer 7. With valves 1 and 3 closed, the water heat storage 8 is warmed. In the transition position of valves 1 and 2 (valve 3 is closed), the two modes can partially operate simultaneously. If valves 1 and 4 are closed, the drying air is warmed in heat exchanger 6 by using the hot water reserves of heat storage 8.

The air leaving the dryer has almost the same enthalpy it had on entering the dryer. A considerable part of the enthalpy used on drying can be regained by condensing the absorbed water vapor. For this purpose a heat pump may be inserted in the energy system. Figure 13.14 illustrates the scheme of a system complete with a heat pump [24]. Part of the moist air leaving the dryer flows through the evaporating heat exchanger 9 of the heat pump, and a proportional part of its moisture content is condensed. The heat input to the working medium of the heat pump (complemented by the input energy of compressor 10 and with the aid of the condenser heat exchanger 11) can be taken into the hot water system. Depending on the ambient state, the air leaving heat exchanger 9 can be returned to heat exchanger 6 of the dryer. (Other labels in the figure are the same as those in Figure 13.13.) In the case of a dryer connected to the energy system of a cattle-raising farm, a heat pump can be also used for cooling milk and producing hot water at the same time.

Inasmuch as the stock-breeding farm possesses a biogas-producing system, the hot water produced can be utilized for heating the gas-producing containers in place of biogas, which can be utilized in other ways. In this case, naturally, biogas can be used as an auxiliary energy source for the dryer during periods of bad weather.

When solar dryers are integrated into the complex energy system of a farm, adsorbent beds can also be utilized as auxiliary units. Adsorbent materials have to be regenerated for exploiting their dynamic adsorption capacities. The regeneration temperature is

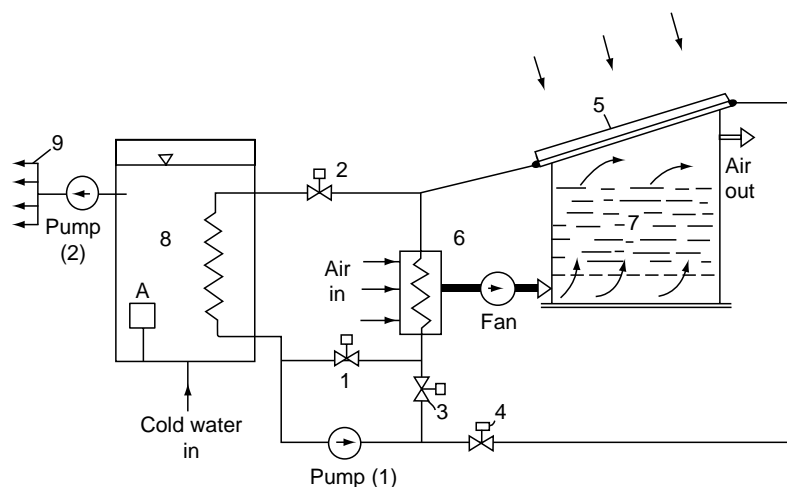


FIGURE 13.13 Solar hay dryer connected to technological hot water system of stock-breeding farm. (From Imre, L., Kiss, L.I., and Molnár, K., in *Proceedings of the Third International Drying Symposium* (J.C. Ashworth, Ed.), Drying Research Limited, Wolverhampton, England, 1982, p. 370; Imre, L., Farkas, I., Kiss, L.I., and Molnár, K., in *Third International Conference on Numerical Methods in Thermal Problems*, Seattle, 1983.)

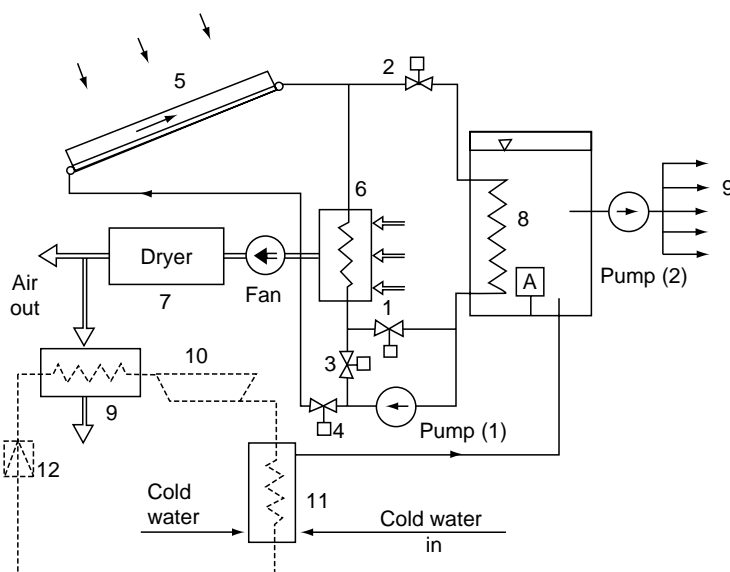


FIGURE 13.14 Complex solar dryer system combined with heat pump. (From Imre, L., Kiss, L.I., and Molnár, K., in *Proceedings of the Third International Drying Symposium* (J.C. Ashworth, Ed.), Drying Research Limited, Wolverhampton, England, 1982, p. 370.)

typically over 150°C, depending on the adsorbent [26]. A condition of economical application is that a considerable part of the energy used for regeneration should be used for producing hot water, for example. During breaks in solar radiation, drying can be continued at the expense of the adsorption capacity of the adsorbents. Therefore, the advantage of applying adsorbents is that a considerable part of the energy used for regenerating the adsorbent can be utilized more cheaply for other purposes than using the same energy for heating the drying air.

13.3.5.5 Solar-Assisted Adsorption Dryer

Figure 13.15 shows the scheme of a complex solar system complemented with adsorbent units [24]. Air warmed in the air-type collector system is delivered to the dryer by fan 1. The ducts of the adsorbent units are joined to the duct section before the fan at points 2 and 3. Unit A2 in the state drawn in the figure (thick lines) is under regeneration; active unit A1 is undergoing adsorption. At joint 3 dry air flows to the airflow of fan 1. Fan 2 draws air through filter F; then the airflow is divided into two parts. In the open position of 8 and closed position of 9 a part of the airflow goes through A1 to the suction side of fan 1; the other part is warmed in heat exchanger H1 and serves the purpose of regenerating unit A2 and then moving into heat exchanger H2. The primary medium of heat exchanger H1 is air heated by energy source E (e.g., biogas), and driven by fan 3. The medium

leaving heat exchangers H1 and A2 flows into the second part of heat exchanger H2. Heat exchanger H2 is connected to the water system of tank T of the hot water system. Water is circulated by pump 1. Pump 2 supplies hot water for 15 consumer lines. (A is the auxiliary energy source of the water tank.)

The energy conditions of solar dryers integrated into a complex energy system are favorable. At the same time, the number of necessary auxiliary equipment is greater, investment costs are higher, and control of the system is more complicated. The economics of such a system depend on local conditions.

Solar dryers can be used with an adsorption bed for energy storage [12,26–29]. The storage of energy accordingly occurs, partly in the form of physical heat and partly in the form of moisture adsorption capacity. The total energy storage capacity of the adsorbent heat store per unit mass is about ten times greater than that of a physical heat store. An optimally designed timber dryer fitted with adsorbent energy storage has been compared with rock-bed storage and proven competitive [28,29].

13.4 ECONOMICS OF SOLAR DRYERS

13.4.1 MAIN ECONOMIC FACTORS

The economics of solar drying depend on the costs involved and benefits gained. Interpretation of the benefits gained by solar dryers is less unambiguous than that of other solar systems [196]. The main

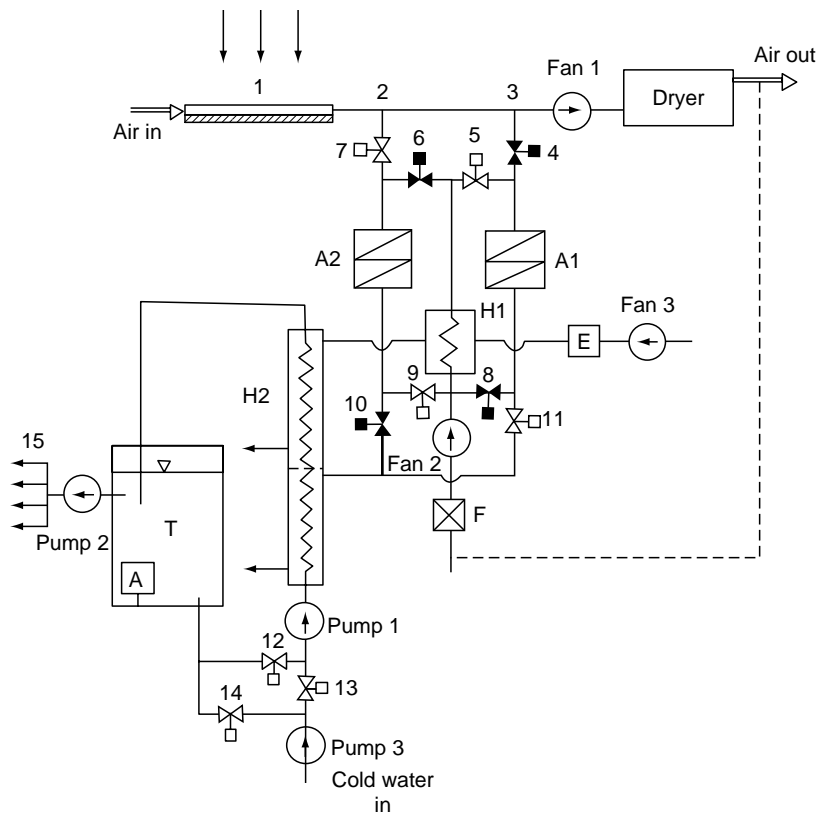


FIGURE 13.15 Arrangement of solar-assisted dryer combined with adsorbent units. (From Imre, L., Kiss, L.I., and Molnár, K., in *Proceedings of the Third International Drying Symposium* (J.C. Ashworth, Ed.), Drying Research Limited, Wolverhampton, England, 1982, p. 370.)

reason can be found in the great variety of solar dryers. For evaluation of the gain a correct basis should be interpreted by finding the extra income (i.e., the savings) of the solar dryers as compared to a basic solution [175,176].

Natural solar dryers should be compared to open-air drying. Savings are realized by reducing the losses of the open-air drying and no energy savings can be considered. Semiartificial solar dryers should be compared to an artificial dryer with the same performance. Their advantages are in reducing the first costs by an unsophisticated construction and by the energy substituted by solar energy. With solar-assisted artificial dryers, savings should be interpreted by the substituted energy. As costs, the investment of the solar energy converting system should be taken into account only.

The savings depend on the lifetime of the dryer and, for the last two main groups of solar dryers, on the cost of the substituted fuel or energy carrier. The lifetime of the dryer can be estimated in advance; it is in any case related to the maintenance costs as well. An error made in the estimation of lifetime may cause significant uncertainty in the economic evaluation.

The price of the dried products and the substituted energy is not stable. These prices may change during the lifetime of any dryer. For changes in the prices, predictions can be used; however, these must be taken as estimates, again causing some further uncertainty in economy calculations.

The sum of overall installation costs is composed of investment costs, interest and maintenance, service, tax, and insurance charges. Inflation modifies the total cost of installation. In view of the uncertainties mentioned it is expedient to make some estimations for the calculations.

13.4.2 DYNAMIC METHOD OF ECONOMIC EVALUATION

Economic evaluation of solar dryers usually aims at determining the payback time. The dynamic method of calculations takes the influence of inflation into consideration. The following considerations summarize this method, according to Böer [30].

Payback occurs when the accumulated savings S equals the sum of investment capital I plus yearly interest and the accumulated costs E :

$$S = I + E \quad (13.1)$$

The annual accumulated savings can be calculated from the net income D by reducing the mass and quality losses and thus by increasing the marketing price of the product in case of natural solar dryers. For semiartificial and solar-assisted artificial dryers savings can be calculated from the price of the substituted conventional energy D . Considering the annual interest rate r and the yearly inflation rate e for the prices of energy, for n years the annual accumulated savings can be calculated as follows:

$$S = \frac{(1+r)^n - (1+e)^n}{r-e} D \quad (13.2)$$

The sum of first investment cost C with interest will be, during n years,

$$I = C(1+r)^n \quad (13.3)$$

Accumulated yearly costs, taking the annual fixed charge rate mC and inflation rate i for equipment into consideration, will be

$$E = \frac{mC(1+r)^n - mC(1+i)^n}{r-i} \quad (13.4)$$

Knowing C and D , diagrams can be made for the determination of payback time n , referring to values

of r , m , i , and e . From these diagrams the requirements for the expected payback time can be easily seen. Figure 13.16 shows the payback time as a function of D/C for various values of r , m , i , and e , following Böer [30]. One can see from the diagram which D/C values can bring about the desired payback time when the various other parameters are at given values. With parameters differing from the above, the calculation must be made separately following Equation 13.1 through Equation 13.4. A comparison of curves 1 and 2 indicates the influence of interest rate r in cases with no inflation; curves 3 and 6, when compared, lead to the effect of energy prices. As can be seen in all the cases examined, the D/C ratio needed for a 10-year payback time falls in the 0.12–0.23 range. Since payback time is a function of D/C , it is obvious that cheaper (smaller C) and less efficient (smaller D) installations are justified insofar as realization of less expense does not mean a significant decrease in the durability (lifetime) of the system.

Payback calculations refer to the whole solar energy drying system [31,32]. With appropriate division of the costs, there is of course nothing in the way of making the calculation only for the collector system [33,34]. Construction of the collectors can be planned on the basis of the economic optimum [35,36].

When the application of a solar dryer results in improved quality of the dried product, the value of D is savings S and has to be increased in relation to the value of the quality enhancement.

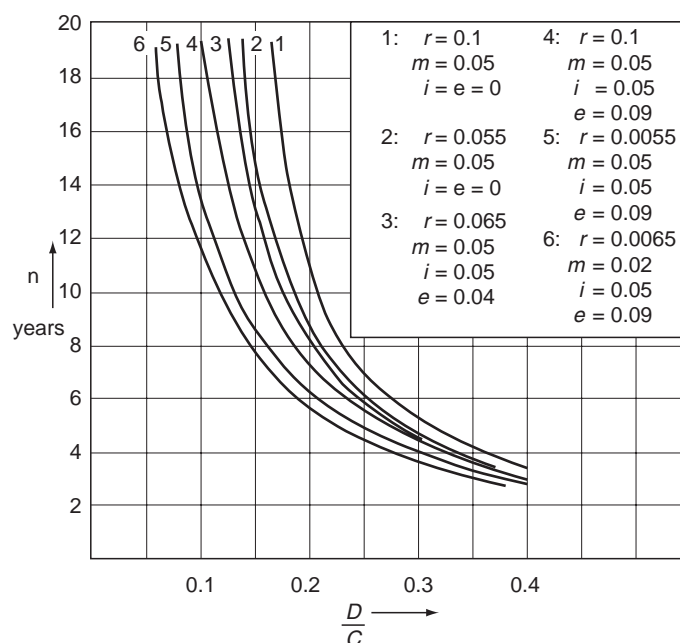


FIGURE 13.16 Payback time in years as a function of D/C , with different parameters r , m , i , and e . (From Böer, K.W., *Solar Energy*, 20, 225, 1978.)

13.5 KEY ELEMENTS OF SOLAR DRYERS

13.5.1 SOLAR COLLECTORS

13.5.1.1 Construction of Solar Collectors

The solar collector plays the part of primary energy source for a solar dryer. Essentially it has functions of energy conversion and energy transfer. As energy converter the collector converts the direct and diffuse radiation coming from the sun into heat. This energy transformation takes place in the so-called absorber of the collector (Figure 13.17). The absorber is made of a material of high absorption coefficient for the radiation of the sun or has a coating with such a character. The radiation absorbed causes the inner energy of the absorber to grow and its temperature to rise.

The energy-transferring function is to transfer the radiation energy transformed into heat in the absorber to the working medium of the collector. The working medium is, with direct system solar dryers, the drying air itself; with indirect systems this is an appropriately chosen liquid (distilled water or, in winter operation, a fluid with low freezing point, oil, and nonaqueous liquids).

Heat transfer between absorber and the medium flowing through the collector occurs by convection. Only part of the heat coming from the incident radiation gets into the working medium.

The part of the radiant energy irradiated that causes an increase of enthalpy of the working medium flowing through the collector is considered utilized heat; the rest is heat loss. For attaining a realistic rate of heat loss most collectors are covered with transparent materials to solar radiation (glass, plastic foil, and others). If the absorption of the covering material and its reflection is high for the absorber's own long-wave radiation, it will reduce the radiation loss of the absorber. For reducing the radiation loss, a

coating that selectively reflects long-wave radiation can also be applied to the covering. Since the temperature of the covering is considerably lower than that of the absorber, the coating will also reduce the convective heat loss from the structure to the ambient air. The number of coverings is usually not more than two. For the reduction of further heat loss it is desirable to insulate the nontransparent parts of the collectors. Efficient means for reducing convective heat loss are collectors with the space between covering and absorber evacuated (vacuum collectors); this makes the collectors expensive.

The final form of the collector is a problem of reaching a technoeconomic optimum (see for instance, the D/C ratio in Section 13.4). In general, beyond a certain limit the reduction of heat loss is no longer economical.

The simplest types of solar dryers (e.g., cabinet dryers, tent dryers, and certain chimney shelf dryers) (see Figure 13.1, Figure 13.2, and Figure 13.4), do not employ a separate absorber; the role of the absorber is played by the irradiated material itself. The majority of high-performance solar dryers are equipped with flat-plate collectors [205].

Figure 13.18 presents flat-plate collectors without covering using air as the working medium. Air flows in the channel between the absorber and the heat insulation. The absorber is a commercially available, rolled metal sheet, usually with a surface coating. Absorbers made of zinc or steel galvanized with zinc can be applied without a coating. The application of collectors without a covering is justified for low-performance dryers.

Figure 13.19 shows some variations of air-type collectors with one covering. These structures can also be made with two coverings. Flow under the absorber reduces the convective heat loss of the air from the covering. There are designs in which air

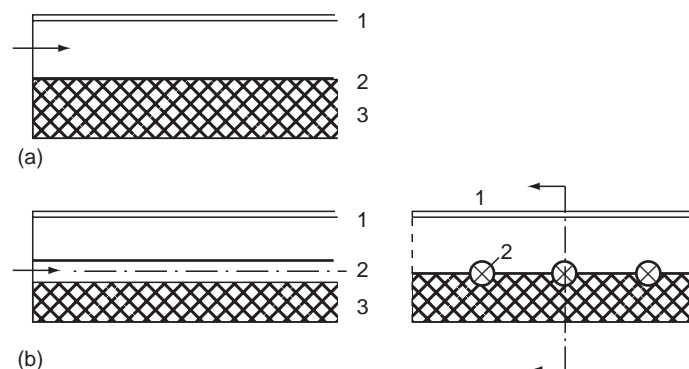


FIGURE 13.17 Setup of flat-plate collectors: (a) air; (b) liquid as working medium (1, covering; 2, absorber; 3, heat insulation).

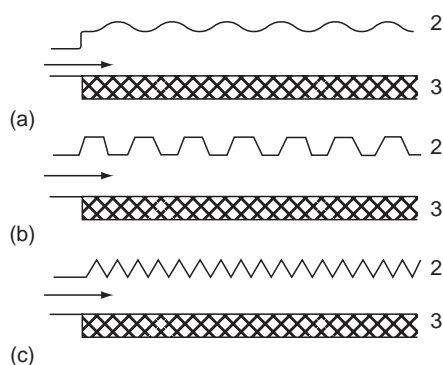


FIGURE 13.18 Scheme of flat-plate collectors without covering, with air as working medium: (a) corrugated plate; (b) trapezoid plate; (c) triangle waved plate (as absorber).

flows on both sides of the absorber. A corrugated or finned absorber surface improves heat transfer between air and absorber. With the latter, flow direction is usually parallel to the fins.

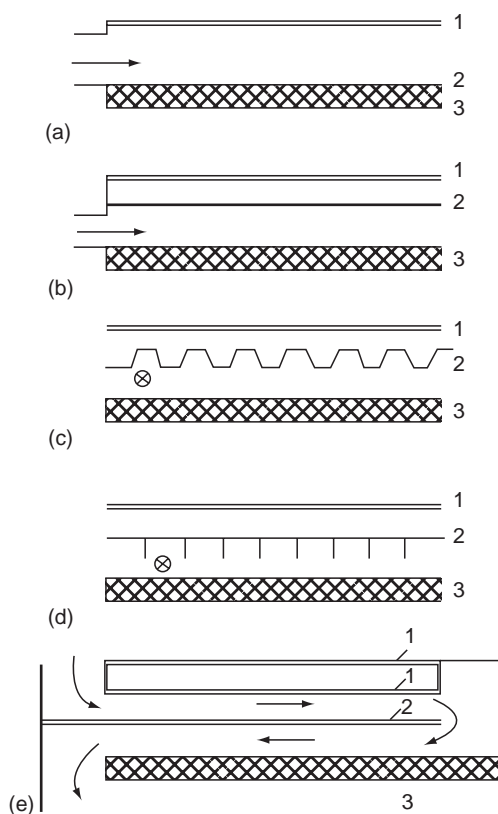


FIGURE 13.19 Scheme of air-type flat-plate collectors with covering: (a) plane absorber, flow over absorber; (b) plane absorber, flow under absorber; (c) absorber with corrugated surface; (d) finned absorber (in c and d, flow is under the absorber, perpendicular to the plane of the paper); (e) plane absorber, flow over and under absorber (two way). (From Vijeyesundera, N.E., Ah, L.L., and Tijoe, L.E., *Solar Energy*, 28, 363, 1982.)

Improvement of heat transfer between air and absorber is aimed at collector designs with absorbers of divided surface (Figure 13.20). The air is forced to flow through the gaps. There are a large number of matrix-type and porous absorbers. The matrix-type absorber, for instance, can be made of wire bundles and of slit and expanded aluminum foil (Figure 13.20c).

An air collector with integrated latent heat storage is shown schematically in Figure 13.21a. The casing of the heat storage material acts as the absorber [37]. Here advantage over air-warmed storage is that heat store warms through direct irradiation; thus heat-storing materials having a phase-change temperature [38,39] higher than the temperature of the air can be used. An integrated rock absorption and storage air collector system has also been developed [40].

Another variation of combined collectors is the hybrid (two-working media) collector (Figure 13.21b). Hybrid collectors are liquid-type collectors in which air flows over an absorber that is common for air and liquid. The application of hybrid collectors is reasonable if the dryer is connected with sensible heat water storage [25]. During collection, hybrid collectors warm air and water simultaneously, the latter serving to charge the heat storage tank. During the radiation-free period of operation (e.g., night), the hybrid collector works as a water–air heat exchanger and, using hot water from the heat storage tank, preheats the drying air. Both integrated collector types can operate at night as they are usually made with a

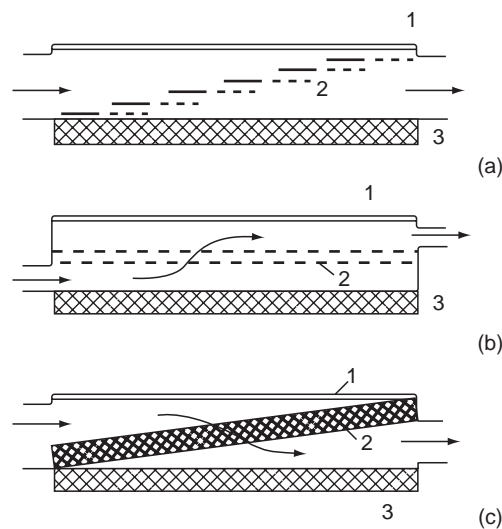


FIGURE 13.20 Air-type flat-plate collector constructed with divided surface absorber: (a) stepwise divided absorber made of overlapped plates; (b) perforated double-plane absorber; (c) matrix-type absorber.

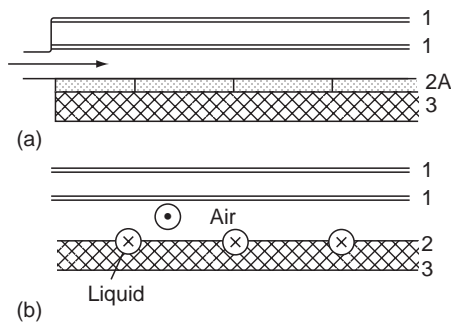


FIGURE 13.21 Integrated collector types: (a) latent heat storage filling, as absorber (2A); (b) hybrid (air-liquid) collector.

double (glass-glass or glass-foil) covering to reduce radiation loss at night.

A very simple design of air collectors is the polyethylene tube collector [41]. A black absorber tube is placed inside a clear tube connected to the fan. The tube assumes its cylindrical form on overpressure. It can be laid on the ground in appropriate length, chiefly for agricultural drying purposes.

Air-type collectors have the advantage of causing no serious consequences if leaking occurs [42]. Difficulties may arise in uniform distribution of air: for advantageous even distribution, considerable fan performance is needed. This has some unfavorable influence on operating costs.

Collectors with a liquid working medium are used for indirect-type solar dryers, usually with water heat storage. Their application for high-performance installations is justified because no large and costly distributing and collecting air channel system is needed as for the air-type collectors. On the other hand, a water-air heat exchanger must be employed.

A drawback of the liquid-type collectors is the danger of leakage and freezing. The former can be averted by appropriate junctions that permit dilatation, the latter by using antifreeze liquids as working media, for example, by integration into the hot water system of the farm for year-round performance.

Owing to the widespread application of solar hot water systems, a great number of variations of fluid collectors have been developed and commercialized circulation. It is mostly the 1–2 m² surface-mounted units that are available commercially; these, joined in an appropriate number, form a full collector system. The advantage of commercially available collector surfaces is the quick and simple replacement of the elements and guaranteed thermal efficiency. A disadvantage is the usually high investment cost and the long payback time. For indirect solar dryers, lower cost is involved with panel-type collectors, if the costs

of collectors need to be reduced. Panel collectors can be used for both air and fluid working media. The absorber and heat insulation form a single continuous surface, and it is only the glass covering that is made of smaller framed parts. If placed on the roof of an agricultural building, the collector integrated in the roof structure can play the role of roofing as well. The absorber surface of panel-type collectors can be mounted on 5–10-m long elements corresponding to the full width; thus the number of joints is considerably less than those of surfaces made of small collector units. In this way not only the construction cost is lower, but the probability of leakage is also reduced.

The absorber of the liquid-type collector most frequently used for solar dryers is a surface formed of ribbed pipes. Its material is metal or synthetic (plastic). Plastic is used for producing carpet-like collectors containing appropriate fluid ducts. The collector carpet usually brought into circulation can be spread over a very large, contiguous surface and can be used with panel collectors, too, with or without covering. Its advantage is simple mounting; its disadvantage is its wear. The ultraviolet (UV)-stabilized construction has a comparatively long lifespan. Plastic is also used for making absorbers of pipelines laid side by side [43]. Their disadvantage—apart from that already mentioned—is that a great number of connections and very careful mounting are required. Another disadvantage of plastics is their sensitivity to high temperature. Even on using a cover a temperature rise in the absorber of over 100°C may occur if there is no heat removal. The temperature of collector types without a covering does not rise to a dangerous level.

Most liquid collectors used with high-performance solar dryers are made with a finned metal tube absorber. The absorbers applied to solar hot water producing systems are often made of sheet halves with stamped passages bonded by seam welding or by rolling them together (Figure 13.22a).

Flat-plate collectors with finned tube absorber (shown in Figure 13.22b) can be built of extruded elements. This is proposed for integrated or panel collectors. The absorber elements perpendicular to the plane of the paper can be ordered from the manufacturer by length of the panel. The structure must be designed so that dilatational movement of the elements is possible. Collectors built of absorbers from pipes soldered or welded to a sheet are shown in Figure 13.22c through Figure 13.22f. The type in Figure 13.22f ensures great strength (rigidity) even with long panel collectors.

The materials commonly used for finned tube absorbers are copper, aluminum, or steel [44]. Copper is rather expensive for dryer collectors. Aluminum gives

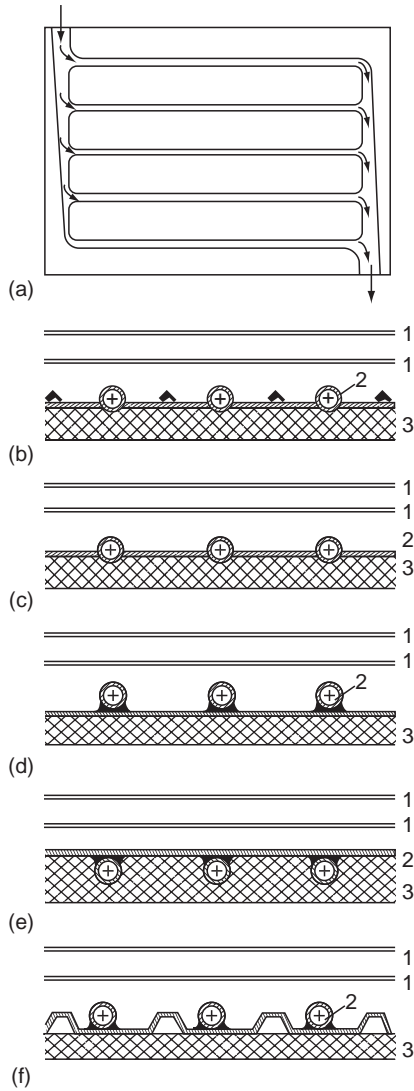


FIGURE 13.22 Some designs of liquid-type collectors: (a) absorber plate made of stamped sheets; (b) collector with extended finned tube absorber elements; (c) through (f) different tube-sheet, flat-plate collectors.

a long operation life with nonaqueous working media. The corrosion of steel can be reduced with the application of inhibitors.

13.5.1.2 Efficiency of Flat-Plate Collectors

The surface required for the collectors of solar dryers can be determined from the energy demand of the dryer. In most solar dryers, drying takes place in stages and only a small part of a dryer is used for drying of continuous material flow.

In the case of drying in stages, the energy demand is not constant: it is greater at the beginning of the drying process and decreases as drying pro-

ceeds. For dimensioning the collectors, the starting point must be the drying requirements and thus the drying characteristics of the material. Drying requirements specify the planned drying time and the permissible material temperature, among others. The drying characteristics of the material serve to determine, with the help of simulation or laboratory drying experiments (or both), the necessary inlet characteristics of drying air (temperature and relative humidity) and the necessary air mass flow rate \dot{m}_a for the drying period with the highest energy demand.

The air is usually taken from the surroundings. In the case of a direct system, the air collector, and with an indirect system the liquid-type collector and liquid-air heat exchanger, serve to heat the air. If heat storage is also employed, the temperature of the medium leaving the collector has to be set to a value that can ensure the prescribed temperature of the drying air even when air heated by the heat storage is used.

In the case of drying in continuous material flow or of preliminary drying, the drying energy demand for a given material is nearly constant over time. The standard energy demand of the dryer ϕ_d is covered by the enthalpy increase of the drying air. When using a direct system, the temperature of the air entering the dryer $T_{d,in}$ is equal to the temperature of the air leaving the collector ($T_{c,out} = T_{d,in}$). The necessary enthalpy increase of the air in the collector is

$$\phi = \dot{m}_a c_{p,a} (T_{c,out} - T_{c,in}) \quad (13.5a)$$

With air-type collectors, if recirculation from the dryer is not employed the temperature of air entering the collector is equal to the ambient temperature ($T_{c,in} = T_o$).

With liquid-type collectors, Equation 13.5a is valid for the air flowing through the fluid-air heat exchanger ($T_c = T_H$). The necessary temperature T'_f of the fluid entering the heat exchanger from the collector ($T'_f = T_{c,out}$) can be determined by the efficiency of the heat exchanger H . For $\dot{m}_a c_{p,a} > \dot{m}_f c_{p,f}$ the energy balance for adiabatic heat exchanger gives

$$\phi_d = \dot{m}_f c_{p,f} H (T'_f - T_o) \quad (13.5b)$$

If the heat loss of the flow duct system ϕ_l is not negligible [45], the energy demand for the collector is $\phi_u = \phi_d + \phi_l$.

Therefore from Equation 13.5b

$$T'_f = T_{c,out} = \frac{\phi_u}{\dot{m}_f c_{p,f} H} + T_o \quad (13.5c)$$

The necessary enthalpy increase for the fluid flowing through the liquid-type collector is

$$\phi_u = \dot{m}_f c_{p,f} (T_{c,out} - T_{c,in}) \quad (13.6)$$

When using air collectors, the full airflow demand by the dryer is generally led through the collector (the value of $\dot{m}_{a,c}$ will be equal to the energy demand of the dryer, \dot{m}_a). The air heated in the collector can be mixed with the air sucked in directly from outside in a mixing space formed on the suction side of the fan.

The \dot{m}_f mass flow rate in the collectors of indirect system dryers can be chosen, within certain limits. However, \dot{m}_f is interdependent with the thermal efficiency η of the collector. After clearing up the necessary energy flow rate to be utilized in the collector, the required collector surface must be determined. On the basis of utilized ϕ_u heat flow rate and the energy flux of the incident radiation (energy flow rate per surface unit: irradiance), the efficiency of the collector can be expressed as

$$\eta = \frac{\phi_u}{A_c I} \quad (13.7)$$

where A_c is the necessary collector surface. Equation 13.7 can be interpreted only as a transient value owing to the time dependence of the irradiance.

For a definite period, the so-called long-term efficiency of the collector can be expressed with the time integral of utilized and input energy flow rates

$$\eta = \frac{\int_{\tau_o}^{\tau} \phi_u d\tau}{A_c \int_{\tau_o}^{\tau} I d\tau} \quad (13.8)$$

The duration for averaging can be chosen in accordance with the operating time of the collector and the purpose of calculation (daily, monthly, or yearly long-term efficiency).

The efficiency of the collector can be determined by calculation and measurements. For design purposes, different calculation methods can be used [34,35,37,46–52]. The efficiency data for commercially available collectors are determined by standard measurements [53,54]. (For “second law efficiency,” see Ref. [55].)

13.5.1.3 Simplified Calculation of Collector Efficiency

In Equation 13.7 of the instantaneous efficiency, utilized heat flow rate ϕ_u is the difference between the heat flow rate absorbed by the absorber ϕ_a and the heat flow rate lost ϕ_l to the ambient air

$$\phi_u = \phi_a - \phi_l \quad (13.9)$$

where

$$\phi_a = t\alpha I A_c \quad (13.10)$$

is the heat flow rate absorbed by the absorber from the irradiation getting through the covering, and

$$\phi_l = A_c U_l (T_a - T_o) \quad (13.11)$$

is the heat flow rate transferred to the ambient air from an absorber at T_a temperature. In Equation 13.11, U_l is the overall heat transfer coefficient of the collector to the ambient air. Substituting into Equation 13.7, the instantaneous efficiency of the collector is [46–50]

$$\eta = t\alpha - U_l \frac{T_a - T_o}{I} \quad (13.12)$$

If t , α , and U_l are taken as constant values, instantaneous efficiency in the function

$$f = \frac{T_a - T_o}{I}$$

(efficiency function, an independent variable) can be plotted as shown in Figure 13.23. At a given operating point, the utilized energy flow rate from the collector is $\phi_u = \eta A_c I$.

These considerations can be appropriately applied according to Equation 13.8 for expressing the long-term efficiency by substituting time averages $(I)_{av}$, $(T_a)_{av}$, and $(T_o)_{av}$:

$$\eta = t\alpha - U_l \frac{(T_a)_{av} - (T_o)_{av}}{(I)_{av}} \quad (13.13)$$

From Equation 13.12 and Equation 13.13 the threshold value of incident radiation flux can be determined

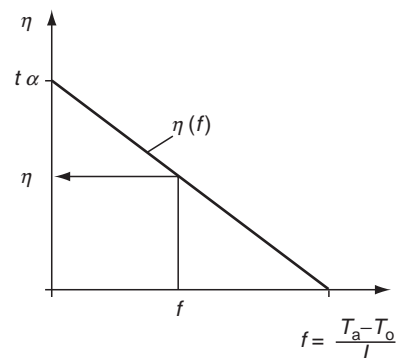


FIGURE 13.23 Instantaneous efficiency diagram of a flat-plate collector.

with which the absorbed energy flow rate and the loss heat flow rate are equal and thus the efficiency is zero:

$$I_{th} = \frac{U_l[(T_a)_{av} - (T_o)_{av}]}{t\alpha} \quad (13.14)$$

From I_{th} , using the appropriate meteorologic data, the possible operation time of the collector can be stated.

The instantaneous efficiency of the collector can also be expressed from the known inlet temperature T_{in} of the working medium with the aid of the heat removal factor F_R [56]:

$$\eta = F_R \left[t\alpha - \frac{U_l(T_{in} - T_o)}{I} \right] \quad (13.15)$$

The coefficient F_R takes into consideration the relative decrease of efficiency caused by the increase of T_a absorber mean temperature compared with T_{in} inlet temperature of the working medium.

Further, instantaneous efficiency can be expressed directly as a ratio of useful heat flow rate coming into the working medium and the incident heat flow rate on the absorber:

$$\eta = \frac{\dot{m}c_p(T_{c,out} - T_{c,in})}{A_c I} \quad (13.16)$$

In practice, $\eta(T_{in} - T_o)$ or $\eta(T_{c,out} - T_{c,in})$ diagrams are often used in place of the $\eta(f)$ efficiency diagram. For representation of the thermal behavior of collectors, besides those above, other practical diagrams, such as $\eta(\dot{m}c_p)$ and $\eta(T_{c,in})$, function curves, can also be used. In these cases other factors in the η equation appear as the parameters of the efficiency curves.

The simplified calculation method has several weak points. One is that the value of T_a must be known to perform the calculation. The temperature of the absorber changes in the flow direction of the working medium, and T_a can be interpreted only as a mean temperature and can be determined only with knowledge of the absorber temperature distribution.

The greatest error appears in the application of the overall heat transfer coefficient U_l and its use as a constant value. U_l models the overall effect of complex and nonlinear heat transfer processes. Its value for a given collector depends on the local values of T_a , on the sky temperature T_s in view of radiation, on the mass flow rate of the working medium, and on the weather (e.g., wind) conditions. In the value of U_l , the temperature dependence of the heat transfer from the covering is strong. One can interpret the value of U_l as the sum of three coefficients: heat transfer from top covering (U_t), from the bottom plate (U_b), and from the edges (U_e):

$$U_l = U_t + U_b + U_e \quad (13.17)$$

For a simple determination of the top loss coefficient U_t , a set of diagrams is presented by Duffie and Beckman [56].

13.5.1.4 Simulation of Flat-Plate Collectors

For computer simulation of the performance of flat-plate solar collectors, several methods and computer programs have been used. Finite-difference [57], network [26,37,54,57], stochastic [58], dynamic [59], and simplified models [60] and methods (see Ref. [99]) have been elaborated.

In the following a short description of a simulation method based on a heat flow network model is given. This method was applied for optimum design and control of roof panel-type collectors of solar dryers [26,37,212].

The essential point of a heat flow network model is the division of the structure of the collector into discrete parts with temperatures that can approximately be characterized by a single value. In the network model the discrete parts are represented by nodes. The heat capacities and heat sources and the so-called temperature sources modeling the boundary condition reference temperatures are connected to the nodes. In this way the ambient air is also represented in the network by a node. The nodes are connected to a network by heat transfer resistances characterizing the thermal interactions among the discrete parts.

Identification of network elements proceeds with constant values at the beginning of the calculation. The identification of temperature-dependent and time-changing network elements takes place in subroutines in the computer program, over the course of calculation, with time increments.

The system of equations for the network can be written from the node and branch equations [37]. For its solution, finite difference or finite time-element schemes can be applied. An advantage of the network model is its flexibility and multipurpose applicability for collectors of different construction. Refinement of discretization is simple to accomplish.

The construction of a network model is shown for a hybrid collector in Figure 13.24a (double covering, finned tube absorber) [57]. The main model conditions are as follows:

- M1: In the collector plane, the temperature distribution is uniform in the direction perpendicular to the flow of the medium.
- M2: For collectors with pipe ducts, the temperature nonuniformity of the absorber perpendicular to the flow direction of the medium is

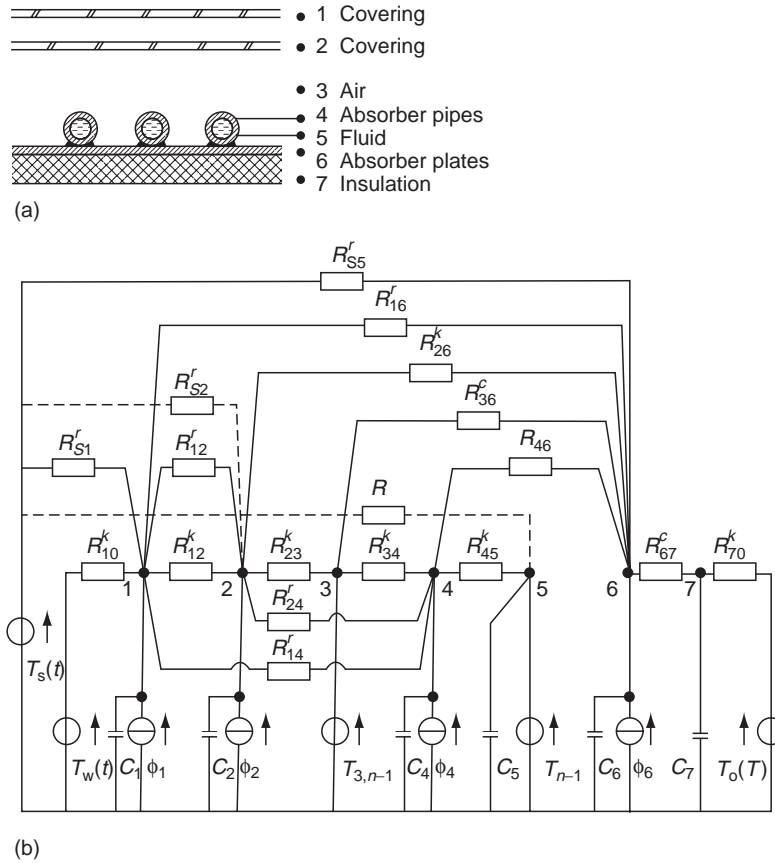


FIGURE 13.24 Heat flow network model of a hybrid collector: (a) setup of the collector; (b) the scheme of the HFN model. (From Imre, L. and Kiss, L.I., in *Numerical Methods in Heat Transfer*, Vol. 2 (R.W. Lewis, K. Morgan, and B.A. Schrefler, Eds.), Wiley, Chichester, England, 1983, chap. 15.)

- taken into account by an average temperature using the fin efficiency [43,56].
- M3: The frame of the collector is assumed well insulated.
- M4: The flow distribution between the collector tubes is uniform (for the effect of nonuniform distribution, see Ref. [51]).
- M5: The spectral variations of absorption and transmission relative to heat radiation are taken into account by average values weighted with the solar spectrum and with the distribution of the low-temperature Planck radiator.

The collector of length L is divided in the flow direction of the working medium in z number of discrete sections of length ΔL ($L = z \Delta L$). The length of the sections is not necessarily equal:

$$L = \sum_{n=1}^z \Delta L_n$$

Further, if it is assumed that the thermal relation between each discrete section is established only by

the flowing medium, the full network model of the collector can be separated into individual network models of discrete sections z . Therefore the solution for the whole collector will be produced through a sequence of the solutions of the individual (in this case, seven-node) part networks proceeding by time increments along z sections.

In Figure 13.24b, the connection of the heat flow network of the n th discrete part is sketched. The numbering of the nodes in Figure 13.24b corresponds to the numbers of the discrete parts in Figure 13.24a.

In the network model, heat flow sources ϕ_1 through ϕ_6 represent the heat flow rates absorbed from incident radiation; C_1 through C_7 are the heat capacities of the discrete parts; $T_s(\tau)$ and $T_o(\tau)$ are the temperature sources giving the temperature of the sky and the ambient air, respectively; and $R_{i,j}^r$, $R_{i,j}^k$, and $R_{i,j}^c$ are radiation, conduction, and convection heat transfer resistances between the discrete parts, respectively. Network element $T_{i,n-1}$ is the temperature source representing the inlet temperature of the working medium flowing into the n th section from

the $n - 1$ discrete section. From the heat balance equation,

$$T_{i,n-1} = T_{i,n-2} + \frac{\phi_{n-1}}{\dot{m}c_p} \quad (13.18)$$

In the case of liquid-type collector simulation, $T_{3,n-1}$ is omitted (see Figure 13.24b). To use the network model for an air-type collector, branches 3–4 and 4–5 of the network are disconnected.

For liquid–air hybrid collectors with counterflow movement of the media, the heat flow network of the collector cannot be separated into z number of part networks with m nodes. In this case the part networks of the sections must be connected into a single zm node network with the aid of the relationship $R_{i(n,n-1)} = -C^{-1}$ formal resistances inserted between the nodes of the working media. (In writing the nodal equations, the formal resistances will be taken into account only in the direction of the medium flow [61].) Taking the lengthwise heat conduction in the absorber into account will lead to a network of the same type. In the following, the correlations for calculating the network elements are described.

The heat flux incident on a given element can be written as

$$\phi_i(\tau) = \alpha_i I(\tau) \prod_{m=1}^{i-1} t_m \quad (13.19)$$

The incident solar flux density I and its directional distribution vary in time. In $I(\tau)$ time function the geographic position, the relative position of the collector to the sun, is present as a regular, periodic variation. The degree of cloudiness, the humidity content of the air, and the degree of pollution of the atmosphere cause a stochastic change in the $I(\tau)$ value, which is superimposed on the regular variation. For the calculation, the solar radiation and meteorologic data relating to the given geographic position are necessary. Such data are usually available [43,56,62–78]. Direct and diffuse components of solar radiation for simulation are characterized by an intensity–time function. There is a significant body of knowledge for determining the effect of cloudiness, the relationships of direct and diffuse radiation intensities [43,56,72,79–84]. The time dependence of absorbance and transmittance of the materials due to direction can be taken into account with an average value [37] (see also Ref. [104]). (For the determination of typical weather for use in solar energy simulations, see also Ref. [100].)

Assuming that the diffuse radiation is isotropic, the density of energy flow rate from incident solar

radiation for a collector in a given location can be calculated from the following equation [37,43,56,84–89]:

$$I(\tau) = I_{H-T}(\tau)(\beta(\tau)\chi(\tau) - \chi(\tau) + 1) \quad (13.20)$$

where $I_{H-T}(\tau)$ is the time-dependent total flux density incident on a horizontal surface.

$$\chi = \frac{I_d}{I_t} \quad (13.21)$$

is the instantaneous ratio of direct and total flux density, a function of time, weather, and the location.

$$\beta(\tau) = \frac{\cos(\phi - \psi) \cos \delta \cos \omega + \sin(\phi - \psi) \sin \delta}{\cos \phi \cos \delta \cos \omega + \sin \phi \sin \delta} \quad (13.22)$$

is a factor characterizing the relative position of the collector and the sun, where ϕ is the latitude, ψ is the inclination angle of the collector, δ is the angle of declination, and ω is the hour angle.

The daily changes of ambient air temperature T_o are known from meteorologic data. The standard sky temperature T_s in clear weather is influenced by air temperature and humidity. A good approximation for determining T_s is given by Swinbank's formula [90]:

$$T_s = 0.0552(T_o)^{1.5} \quad (13.23a)$$

Bliss's correlation [91] also takes air humidity content into account

$$T_s = T_o \left(0.8 + \frac{T_{dp} - 273}{250} \right)^{0.25} \quad (13.23b)$$

where T_{dp} is the local dew-point temperature of the air (temperatures in Kelvin). (For examining night conditions, see Refs. [92–94].)

Heat transfer resistances can be interpreted by the branch equation of the heat flow network:

$$\phi_{i,j} = \frac{\Delta T_{i,j}}{R_{i,j}} = K_{i,j} \Delta T_{i,j} \quad (13.24)$$

where $K_{i,j}$ is the conductivity of the branch. Radiation conductivities for surface units are

$$K_{s,i}^{(r)} = \varepsilon_i \sigma (T_i + T_s)(T_i^2 + T_s^2) \quad (13.25)$$

$$K_{i,j}^{(r)} = \varepsilon_{i,j} \sigma (T_i + T_j)(T_i^2 + T_j^2) \quad (13.26)$$

$$\varepsilon_{i,j}^{-1} = \varepsilon_i^{-1} + \varepsilon_j^{-1} - 1 \quad (13.27)$$

where ε_i is the emissivity of the glass, ε_j is the emissivity of the absorber, and σ is the Stefan–Boltzmann constant.

Heat conductivity in solid elements of δ thickness for surface unit is

$$K_{i,j}^{(c)} = k\delta^{-1} \quad (13.28)$$

Conductivities (convection) for the outer surface units are

$$K_{i,o}^{(k)} = h \quad (13.29)$$

where h is the heat transfer coefficient that depends on wind velocity. Approximate correlations for h as a function of wind velocity w are

$$\begin{aligned} h &= 5.7 + 3.8w \quad \text{for } w < 5 \text{ m/s} \\ h &= 7.6w \quad \text{for } w \geq 5 \text{ m/s} \end{aligned}$$

(See Ref. [95] for further details on wind effects.)

Heat transfer coefficients for forced and natural convection in air gaps can be calculated from correlations [43,56]. Substituting, on the basis of the branch Equation 13.23a and Equation 13.23b, the ϕ_{ij} values into the nodal equations, we get the following system of equations for the network:

$$\underline{C} \dot{\underline{T}}(\tau) + \underline{K}(\underline{T}, \tau) \underline{T} = \underline{\phi}(I, T_s, T_o, \tau) \quad (13.30)$$

where \underline{C} is the diagonal matrix of the nodal heat capacities, $\dot{\underline{T}}$ is the column vector of the time derivatives of the nodal temperatures, \underline{T} is the column vector of the nodal temperatures, and $\underline{\phi}$ is the source vector.

The solution of the system (Equation 13.30) can be obtained numerically by time discretization. Because of the nonlinearities, the integrated mean value of the conduction matrix $\underline{\hat{K}} = \underline{\hat{K}}(T(\tau))$ for a given period of time can be created and a numerical scheme of one or two time levels can be applied [96]. Favorable results have been obtained by the linear Galerkin scheme [97]:

$$\begin{aligned} \left[\frac{1}{3} \underline{\hat{K}} + \frac{1}{2\Delta\tau} \underline{C} \right] \underline{T}^{(k)} &= \left[-\frac{1}{6} \underline{\hat{K}} + \frac{1}{2\Delta\tau} \underline{C} \right] \underline{T}^{(k-1)} \\ &+ \frac{1}{6} \left[\underline{\phi}^{(k)} + 2\underline{\phi}^{(k-1)} \right] \end{aligned} \quad (13.31)$$

In this equation, k is the number of discrete time sections in the step-by-step solution ($\tau = k\Delta\tau$). Meteorologic data are at one's disposal hourly, so in

the calculation a time step of 1 h can be applied. For stability it is advantageous if the system of equations is well conditioned. Accordingly, heat capacities of insignificant influence are best neglected. As a result of calculation, the time dependence of the medium outlet temperature, as well as the time variation of the approximate temperature distribution of the collector, is obtained. The results can be used for instantaneous and steady-state efficiency diagrams, for collector design and optimization, and for the solution of process control problems [33,34,43,50,56,98]. It is of course desirable to verify the calculation results with experimental measurements where possible.

13.5.1.5 Thermal Performance of Flat-Plate Collectors

Figure 13.25 shows the characteristics of a single-covering, air-type collector (black absorber) obtained by calculations [37]. In the figure, ΔT is the rise in temperature of air in the collector ($\Delta T = T_{c,out} - T_{c,in}$), η is the instantaneous efficiency, and \dot{m} is the mass flow rate of air. The parameter in the figure is the surface area per unit collector length (series connection).

In Figure 13.26, the instantaneous efficiency and the outlet temperature of a liquid-type collector (single-covering, steel finned tubes, black absorber) are illustrated as a function of liquid heat capacity flow

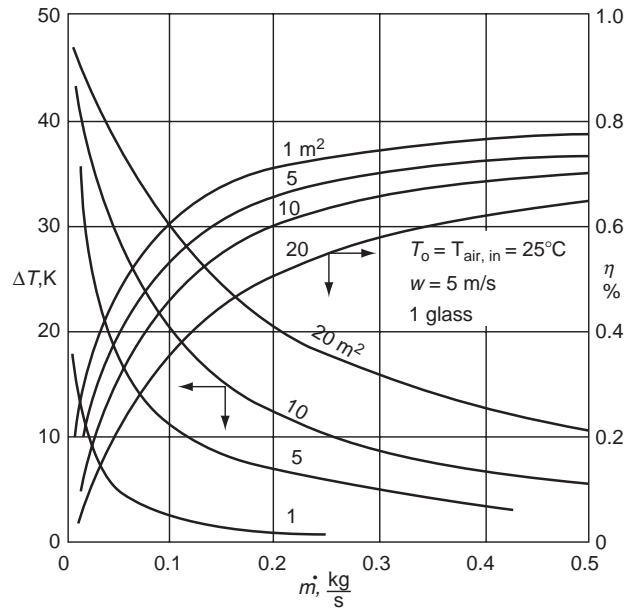


FIGURE 13.25 Main characteristics of an air-type collector. (From Imre, L. and Kiss, L.I., in *Numerical Methods in Heat Transfer*, Vol. 2 (R.W. Lewis, K. Morgan, and B.A. Schrefler, Eds.), Wiley, Chichester, England, 1983, chap. 15.)

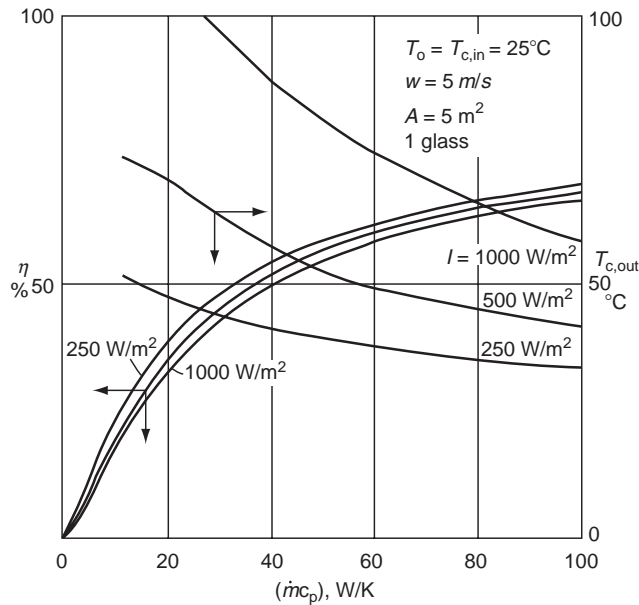


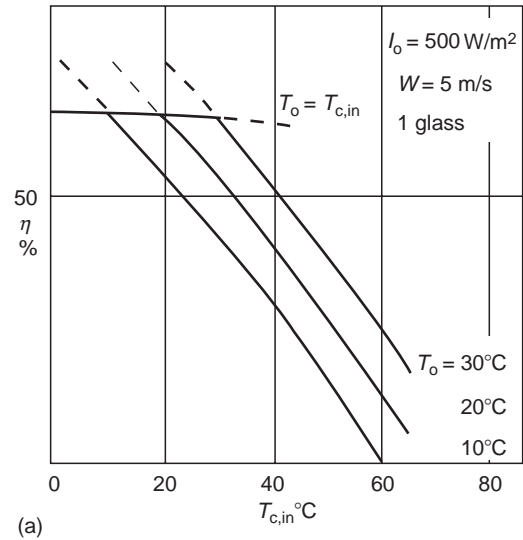
FIGURE 13.26 Main characteristics of a liquid-type collector. (From Imre, L. and Kiss, L.I., in *Numerical Methods in Heat Transfer*, Vol. 2 (R.W. Lewis, K. Morgan, and B.A. Schrefler, Eds.), Wiley, Chichester, England, 1983, chap. 15.)

rate [37]. The parameter is the irradiation. As the variations of I are accompanied by nonlinear heat transfer resistance variations, the curves for different I values deviate. The entry temperature of the medium is equal to the outside temperature ($T_{c,in} = T_o$).

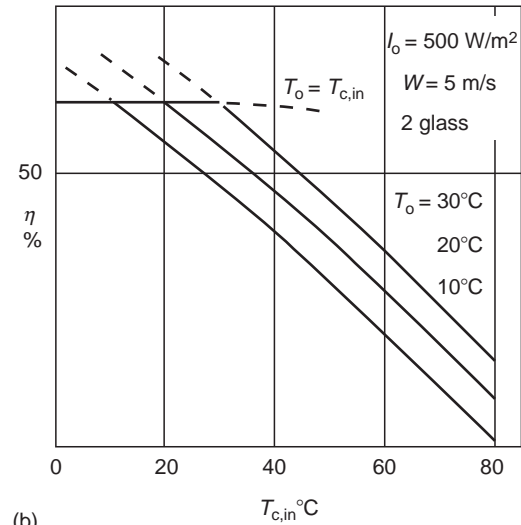
Figure 13.27 illustrates the instantaneous efficiency of a liquid-type collector as a function of the inlet temperature with single (Figure 13.27a) and double (Figure 13.27b) covering [37]. The parameter is the outside temperature. With an increase in the inlet temperature, the mean temperature $(T_a)_{av}$ and the heat loss of the absorber increase while the efficiency deteriorates. With double covering, the efficiency with a higher inlet temperature is greater than that with a single covering.

Figure 13.28 shows the instantaneous efficiency diagrams for different collectors. It can be seen from this figure that the instantaneous efficiency of collectors now in general use can be as high as 50–60%. The value of daily long-term efficiency amounts to approximately 25–30%.

On the basis of the long-term collector efficiency, the collector surface necessary for the operation of the dryer can be determined. Because the instantaneous efficiency of the collector is over one part of operation time greater than the long-term efficiency, the energy utilized by the collector over this period is greater than the necessary value. The surplus energy can be stored for the period when there is no solar energy available. In cheap, simple construction dryers without



(a)



(b)

FIGURE 13.27 Instantaneous efficiency of a liquid-type collector as a function of inlet temperature; parameter is the ambient temperature: (a) single covering; (b) double covering. (From Imre, L. and Kiss, L.I., in *Numerical Methods in Heat Transfer*, Vol. 2 (R.W. Lewis, K. Morgan, and B.A. Schrefler, Eds.), Wiley, Chichester, England, 1983, chap. 15.)

heat storage, this surplus energy can be used for some temporary enhancement of the drying process if the material to be dried can withstand it.

13.5.2 HEAT STORAGE FOR SOLAR DRYERS

From a thermal viewpoint heat storage of solar dryers can be classified into two main groups:

1. Directly irradiated heat storage
2. Heat storage charged by the working medium of the collector

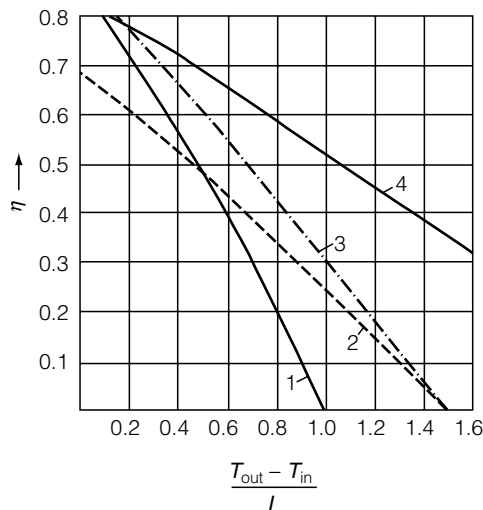


FIGURE 13.28 Efficiency diagrams of collectors of different construction: (1) black absorber, one covering; (2) black absorber, two coverings; (3) selective absorber, one covering; (4) vacuum collector.

The storage temperature of directly irradiated heat storage is not limited by the collector outlet temperature of the medium. However, in heat stores warmed by the working medium the maximum temperature cannot exceed the collector outlet temperature.

The aim of heat storage is to store surplus energy appearing in strong radiation periods; however, the aim may also be to store enough energy for full-scale drying operation at night as well. When determining the necessary surface area of the collector, the

amount of energy to be stored must also be taken into account.

13.5.2.1 Directly Irradiated Heat Storage

Directly irradiated heat storage of solar dryers can first be used in direct air systems. The collector sketched in Figure 13.21 is built with phase-change storage. The phase-change material is placed in a plastic honeycomb matrix casing. Thermal expansion is made possible by additives. The outer surface of the plastic cells containing the phase-change material take over the function of the absorber. The thickness of the cells is limited by phenomena occurring in the course of phase change [38,39,101,102]. Therefore the mass to be placed on 1 m^2 is also limited, and so is the overall amount of storage heat.

The primary purpose of applying directly irradiated latent heat storage is to attain an equalizing effect during cloudy periods over the day as well as to lengthen the daily drying time. To show the thermal behavior of a collector integrated with latent heat storage, Figure 13.29 is presented. The temperature–time function was determined by calculation using the simulation model described above and checked by measurements. The heat capacities of other elements of a heat storage collector are negligible when compared with that of the heat storage. The effect of latent heat was built in the volumetric heat capacity of the absorber and simulated [38,103,104] (data: material, $\text{CaCl}_2 \times 6\text{H}_2\text{O}$; phase-change temperature, 29°C ; latent heat, $\lambda = 209 \text{ MJ/m}^3$; size of cells, $9 \text{ mm} \times 10 \text{ mm}$; mass/ m^2 , 6.3 kg).

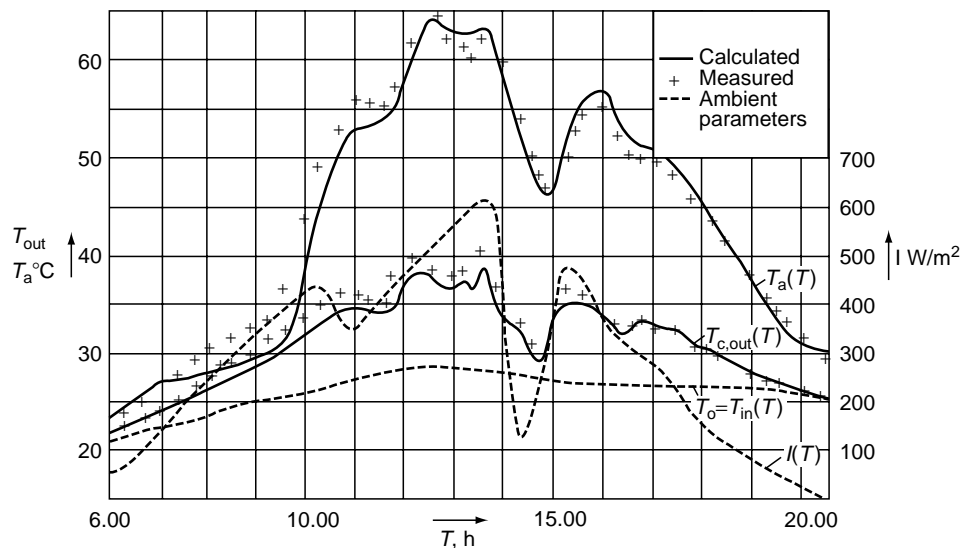


FIGURE 13.29 Behavior of air-type collector integrated with phase-changing filling (functions of T_{out} and T_a determined by simulation and measurements).

The absorber may be a solid (e.g., concrete) wall with a black coating. In the case of solar dryers, this solution is justified if the wall is the dryer housing wall. Its operating principle is the same as that of the Trombe wall used in passive solar heating [56]. (For collector-integrated solar water heaters, see Refs. [105,108].)

13.5.2.2 Heat Storage Charged by the Medium of the Collector

In heat storage warmed by the working medium, the maximum temperature cannot exceed the maximum temperature of the medium leaving the collector. As storage material, various solids, water, various phase-change materials, as well as chemical storage materials may be used.

The temperature of the medium entering the storage is equal to the temperature of the medium leaving the collector if the heat loss of the duct is negligible. The medium in liquid systems after leaving heat storage flows back into the collector. With air systems sometimes the air leaving the heat storage unit is exhausted into the atmosphere.

Water-type heat storage may be a direct or indirect system. In direct systems the working medium flows in a closed circuit (Figure 13.30a). In indirect systems, the working medium can be liquid (anti-freeze liquids in some cases). The closed collector circuit and the closed storage circuit are connected through a heat exchanger (Figure 13.30b). In water heat storage, two types of storage systems are employed: stratified and well-mixed storage.

In stratified water storage the warm water from the collector enters near the top of the tank; the fluid led back to the collector is drawn from the bottom of the tank by a pump. Thus in the upper layers of the tank there is always warm water, and the lower layers contain cold water. The advantage of this method is that the collector receives cold water as long as cold layer exists near the bottom of the tank [56,106,107]; accordingly, the collector works with approximately constant efficiency. The thickness of the transient temperature zone is determined by the time boundaries of the temperature changes of the water coming from the collector. In operation periods of reduced radiation, the temperature of the water from the collector is lower than that of the temperature in the top layer. This water descends and causes mixing in the tank.

When using stored hot water (Figure 13.31), the water is led into the air–water heat exchanger from the top of the tank. The returning water enters at the bottom of the tank. The mass of water in the tank therefore makes on charging a slow downward motion and on discharge a slow upward motion. The rate of this motion depends on the mass flow rate of

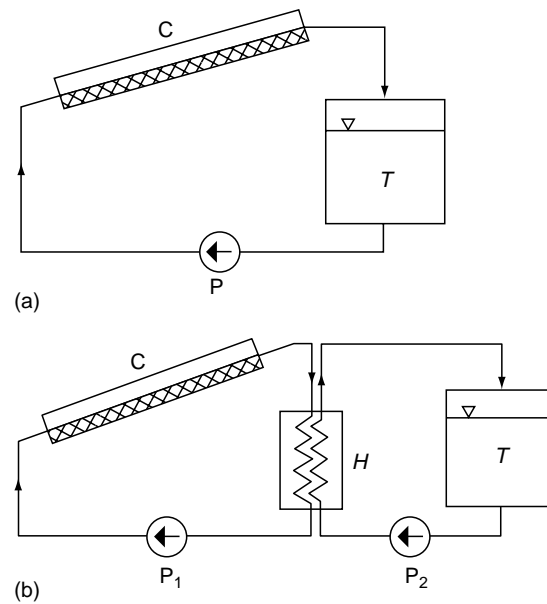


FIGURE 13.30 Water-type heat storage: (a) direct; (b) indirect (C, collector; T, tank; P, pump; H, heat exchanger).

the water (and on other effects, see below). The arrangement shown in Figure 13.31 allows regulation of the water flow rate, the change of operational mode, and a simultaneous drying–charging operation.

The amount of heat to be stored Q can be determined from the heat demand of the dryer:

$$Q = \int_0^{\tau} \phi(\tau) d\tau \quad (13.32)$$

Storage temperature is limited by the fluid temperature that can be attained in continuous operation of the collector. In a direct storage system (Figure 13.30a) disregarding heat loss from the pipes, the temperature of the medium coming from the collector can be calculated by using Equation 13.16:

$$(T_{c,out})_{av} = (T_{c,in})_{av} + \frac{A_c I \eta}{\dot{m} c_p} \quad (13.33)$$

The heat to be stored in τ charging time, neglecting the heat loss, is

$$Q = M_T c_T [(T_{c,out})_{av} - (T_{c,in})_{av}] = \dot{m} c_p [(T_{c,out})_{av} - (T_{c,in})_{av}] \tau \quad (13.34)$$

Hence,

$$M_T = \frac{\dot{m} c_p}{c_T} \tau \quad (13.35)$$

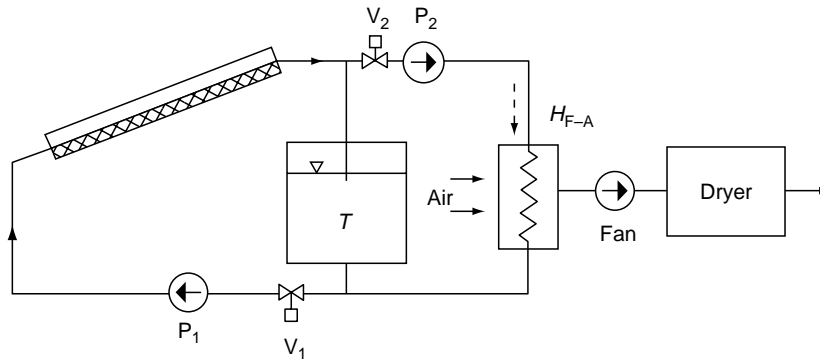


FIGURE 13.31 Connection of water-type heat storage to the dryer (H_{F-A} , fluid-air heat exchanger; V_1 and V_2 , valves).

This calculation is approximate because it refers to continuous operation assessed from time averages and does not consider existence of stratification and the consequences of heat loss. A more exact calculation can be made from the discretized model of the storage tank with due regard to heat loss [56,106,107].

In indirect storage systems (Figure 13.30b), the temperature of hot water entering the store T_T can be calculated from the effectiveness H of the heat exchanger ($\dot{m}_c c_{pc} \geq \dot{m}_T C_{PT}$):

$$T_{T,in} = T_{T,out} + H(T_{c,out} - T_{T,out}) \quad (13.36)$$

The relationship corresponding to Equation 13.34 can be written with the respective values of $T_{T,in}$ and $T_{T,out}$.

In the case of stratified heat storage the mass flow rate of the collector medium is held during charging at a relatively low value for the purpose of getting as high a value of $T_{c,out}$ exit temperature as possible. A disadvantageous consequence (see Figure 13.25) is only moderate efficiency.

With a well-mixed heat storage system, the temperature of the water in storage is practically uniform. Mixing of the water can be produced by several (eventually simultaneous) effects. These are a large mass flow rate and inlet velocity, horizontal location of the tank, in an indirect system the heat exchanger located at the bottom of the tank, baffle plates in the tank, and the use of a circulating pump for mixing.

In the course of charging the perfectly well-mixed heat storage, the temperature of the mass of the water in storage and the temperature of the water returning to the collector rises (see also Section 13.6.3). As a consequence, the collector works at an ever-increasing temperature level with an ever-decreasing efficiency. However, its long-term efficiency is not necessarily lower than that of stratified storage collectors, although these receive, during a major part of charging, an

entering fluid of low temperature. That is, the mass flow rate of the working medium can be considerably greater in collectors of well-mixed heat storages as there is no interest in a significant increase of fluid temperature in the collector. With a higher mass flow rate, however, the efficiency of the collector will increase (see Figure 13.26). Taking the usual daily 6–8 h charging time, stratified heat storage and well-mixed heat storage are about equal from a thermal point of view.

In the operation of a solar dryer, technological interest is attached to a sufficiently high temperature of the fluid leaving the collector for preheating the drying air. At the same time, during certain periods of operation, simultaneous actions must be carried out for charging the store and for drying (see Figure 13.31). Consequently, the mass flow rate of working medium of the collector must be limited to reach the necessary exit temperature of the collector. Therefore, in the operation of solar dryers, the stratified heat store is preferred.

In solar dryers with air-type collectors rock-bed storage (crushed stone or pebbles of 2–4 cm size) is used most commonly. When selecting the material for the rock or pebble bed, among other considerations, there is the question of the pressure drop across the bed. The flow resistance of a pebble is usually smaller than that of crushed rock. Uniform pebble or rock size must be chosen to obtain uniform air distribution in the bed. The necessary mass of rock-bed or pebble-bed heat storage is typically about threefold that of water-type heat storage.

Because of the point contact between the particles in the rock bed, the heat conduction is negligible. Therefore, rock beds work practically as stratified heat stores no matter the arrangement for air inlet and exhaust.

During charging of the rock-bed heat store, the cooling of the air at the entry spot takes place within a layer of a certain thickness. If the entry temperature of the air was constant, this layer of changing

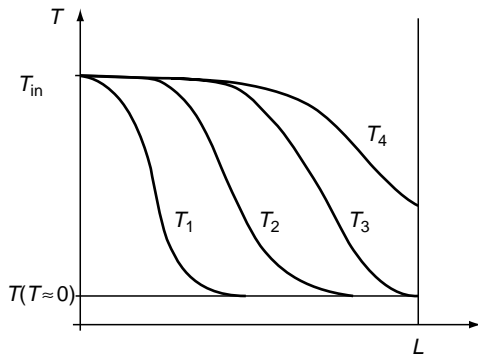


FIGURE 13.32 Passing of temperature wave in a gravel-bed heat storage with depth L ($T_{in} = \text{constant}$).

temperature would push down deeper and deeper (Figure 13.32). However, because during the day periods of varying radiation intensity occur, it is possible that during certain periods of the day the temperature of entering air is lower than that of the heat-storing material at the place of entry. In this case the air in the entry layer continues to get warmer and delivers the earlier stored heat into deeper layers of the storage, to a lower level of temperature, and enlarges the width of the layer of changing temperature. The principle of approximate calculation of rock- or pebble-bed heat storage is similar to that of a stratified water-type heat storage. Figure 13.11 present the arrangement of solar dryers equipped with pebble-bed heat storage.

Between water and solid-bed heat storages there is a so-called transitional option heat storage with the aid of fluid-filled cans. An example of this is shown in Figure 13.5. In heat storage devices warmed by a working fluid, phase-change materials can also be used [38,39,101,102]. Phase-change energy storage has two main advantages. One advantage is that the energy stored per unit volume of the store is significantly greater than that for pebble-bed heat storage, for example. One of the cheapest materials with good working characteristics [109] is $\text{Na}_2\text{SO}_4 \cdot 10\text{H}_2\text{O}$; with this the proportion of stored energy is about fourfold (latent heat $\lambda = 251 \text{ kJ/kg}$; phase-changing temperature, 32°C). Also, the fixed phase-change temperature reduces temperature variations.

The application of phase-changing materials (PCM) also involves certain problems. Eutectic salts are susceptible to phase separation [110,111]. To avoid this [38,112] it is desirable to make the PCM containers thin (Figure 13.33). Another phase-change material is paraffin wax ($\lambda = 209 \text{ kJ/kg}$). Its disadvantage is inflammability.

A general problem with PCM stores is the fabrication of an inexpensive casing resistant to corrosion.

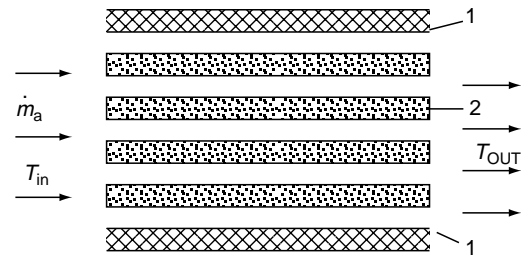


FIGURE 13.33 Example for placing phase-changing fillings in the heat storage: (1, insulated wall; 2, containers; $\dot{m}_a = \text{constant}$).

For some PCMs, degradation has been experienced over the course of repeated cycling. The choice between phase change and sensible heat storage must be made on economic grounds.

Heat storage by chemical reactions has also been studied [56,112,113]. These methods have not progressed beyond the experimental stage and do not appear applicable for solar drying applications.

Energy storage in adsorbent beds can be applied to solar dryers because air of reduced humidity content leaks from the adsorbent. Silica gel, activated aluminum oxide gel, zeolites, and various molecular sieves can be used as adsorbents [27,29,114]. Detailed analyses and optimization procedures are known for energy storage in timber dryers [29] and garlic dryers [12]. The applications of adsorbent energy storage are also found in complex systems [25]. Although the advantages are undisputed, the costs of adsorbents and the auxiliary equipment are considerable.

13.6 SIMULATION OF SOLAR DRYERS

13.6.1 PURPOSE OF SIMULATION

Simulation is an important tool for design and operation control [200,201,220]. For the designer of a drying system, simulation makes it possible to find the optimum design and operating parameters. For the designer of the control system, simulation provides a means to devise control strategies and to analyze the effects of disturbances.

13.6.2 METHODS OF SIMULATION

Various simulation models for solar processes have been reported. They differ mainly in the assumptions made and strategies employed to solve the model equations. A majority of the models refer to solar heating [115–117]. One widely used simulation program is the TRNSYS [56,118]. The f -chart method for solar heating of buildings is also well known [56,119,120]. Other relevant references are design of

closed-loop solar systems [121], control and optimization [117], and process simulation based on a stochastic model [58] and for a resistance network [56,122].

Selcuk et al. [5,123] describe the simulation of a shelf-type dryer. Close [124] worked out the simulation of an air-type solar drying system equipped with gravel-bed heat storage using a 10-node discretized model for the heat storage and forward finite difference technique for solving the equation system. Imre et al. [25,125] presented a simulation of solar dryer for alfalfa.

Daguenet [187,188] elaborated a simplified methodology for the calculation of solar-assisted convective dryers considering the meteorological conditions. Norton and Hobson [189] suggested a finite difference numerical analysis for chimney-type solar dryers for drying crops. Tiguert and Puiggali [190,191] published a performance model for a solar natural chimney dryer and are given in Figure 13.5. The one-dimensional, concentrated parameter model consists of four elementary models (i.e., for the collector, for the heat storage in daytime and in nighttime, and for the material situated in thin layers in the drying space).

Steinfeld and Segal [192] proposed a simulation model for a solar thin-layer drying process including the technique of estimating the solar radiation, the procedure of obtaining the thermal performance of a solar air-heater, and an analysis of the drying process based on the Lewis analogy and the equilibrium moisture content (EMC) concept. Hasnaoui et al. [186] elaborated a one-dimensional model for a solar dryer of static thick bed.

Patil and Ward [194] prepared a simulation model for a solar-assisted bin dryer with and without mixing in atmospheric air. The model consists of the part model of the solar collector and the part model of the thick through-flow layer divided into thin layers.

Weitz et al. [195] proposed a simulation model for a multisshelf type semiartificial solar dryer for drying processes in thin layer based on Luikov's theory by considering the shrinkage of the material. Mahapatra et al. [206] elaborated simulation model for directly irradiated solar dryer with integrated collector.

13.6.3 SIMULATION MODEL OF SOLAR DRYERS

Solar dryers are thermohydraulic systems composed of various units such as collector, heat storage, heat exchanger, pump, ventilator, tubes, valves, closing and controlling devices, and the dryer. A simulation model of solar dryers is therefore made up of three main subsystems:

1. Model of the flow subsystem
2. Model of the thermal subsystem
3. Model of the drying volume

13.6.3.1 Model of the Flow Subsystem

The units of the solar dryer are joined into a system by flowing media; therefore a model of the system is best built on the flow model of these media [126]. The number of flow models is necessarily equal to the number of working media in the solar dryer.

The flow model serves to determine the mass flow rates of the flowing media. In the following section the mass flow network (MFN) modeling method is presented. The MFN model of solar dryers essentially divides the flow system into discrete parts in which the effects causing pressure changes are modeled by network elements; the network elements are joined according to the flow path of the medium.

Pressure changes can be produced by external as well as internal effects. An example for an external effect is due to the pump or the fan. Internal effects can be the consequences of cross-sectional changes, heat transfer, acceleration or deceleration, and flow resistances. Some of the most important MFN elements are given in Table 13.2 [26].

In the MFN model the flow resistances and pressure sources are dependent on heat flow or temperature. Discretization has to be made so that over an individual discrete section the temperature-dependent characteristics can be expressed by lumped values with sufficient accuracy. Pipes, which can be taken as isothermal and have a constant cross-section can be modeled as a single flow resistance.

Flow resistances are characterized for the j th section on the basis of the pressure drop Δp_j and the mass flow rate \dot{m}_j (see Table 13.2):

$$R_j(\dot{m}_j, T_m) = \Delta p_j \dot{m}_j^{-1} \quad (13.37)$$

For the simulation it is necessary to know the state functions $\Delta p(\dot{m}, T, n)$ of the pumps and fans used.

Figure 13.34a shows the arrangement of an indirect system solar alfalfa dryer with water-type heat storage serving simultaneously as the hot water supply [25]. (For a detailed description, see Figure 13.13.)

The system has two working media. The primary flow circuit is closed; its medium is water. The secondary flow circuit is open; the working medium is air. Thermal interaction of the working media of the two circuits is carried out by a heat exchanger.

Figure 13.34b and Figure 13.34c outline the reduced MFN model of the primary and secondary flow circuits, respectively. In the reduced model of the primary flow circuit (Figure 13.34b), R_c and R_H are the resultant flow resistances of the collector and the heat exchanger, respectively, including the resistances of the pertinent pipe sections. (In the detailed model there are several partial resistances corresponding to the discretization.)

TABLE 13.2
Identification of the Elements of the Mass Flow Network Model

Element	Relationship	Note
Flow resistances	$R = \Delta p \dot{m}^{-1}$	
Tube friction resistances		From Poiseuille equation ($\dot{m} = \rho A w$)
Laminar flow	$R = \frac{c \eta L}{\rho d_h^3 A}$	
Turbulent flow	$R = \frac{2fL}{\rho d_h A^2} \dot{m}$	From Fanning equation
Shape resistance	$R = \frac{s}{2\rho A^2}$	From $\Delta p = s \frac{\rho}{2} w^2$
Jump-like change of cross-section	$R = \xi_B \frac{\dot{m}}{2\rho} \left(\frac{1}{A_1^2} - \frac{1}{A_2^2} \right)$	ξ_B : Borda–Carnot coefficient
Pressure sources		
Centrifugal pump and fan	$\Delta p = A + B\dot{m} + C\dot{m}^n$	Equation of the characteristic
Change of cross-section ($A_1 \geq A_2$)	$\Delta p = \frac{\dot{m}}{2\rho} \left(\frac{1}{A_2^2} - \frac{1}{A_1^2} \right)$	From Bernoulli equation
Hydrostatic pressure source	$\Delta p = \rho g (z_1 - z_2)$	z_1, z_2 level heights
Closed thermosyphon loop	$\Delta p = g \oint \rho(T(z)) dz$	$T(z)$: temperature as a function of the level height z
Heat input pressure source between points 1 and 2 of a pipe ($A = \text{constant}$)	$\Delta p = \frac{R}{(c_p)_{av}} \left[\rho_{av} \frac{\phi_h}{\dot{m}} - \frac{\dot{m}^2}{2A^2} \left(\frac{1}{\rho_1} - \frac{1}{\rho_2} \right) \right]$	From enthalpy balance equation
Acceleration pressure source	$\Delta p = \rho a L$	$\rho = \text{constant}$
Mass flow source (independent)	$\dot{m} p^{-1} = \text{constant}$	Volumetric pump

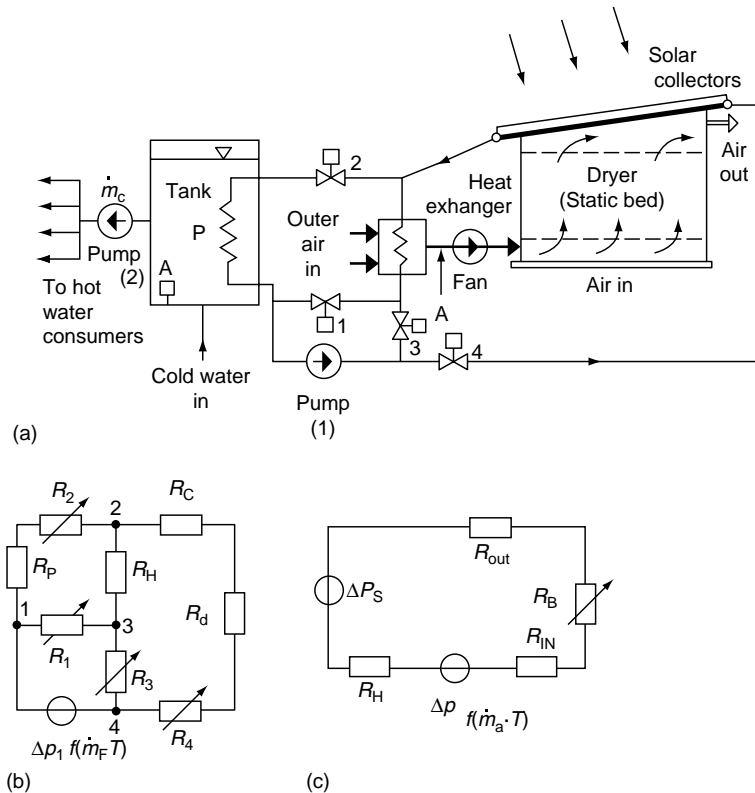


FIGURE 13.34 Solar-assisted indirect drying system for alfalfa: (a) scheme of system; (b) reduced mass flow network model of the fluid; (c) reduced mass flow network model of the airflow system.

In Figure 13.34b, R_d is the flow resistance of the forward pipe section, R_1 through R_4 are the variable shape resistances of the control valves, and Δp is the pressure source due to the pump.

In the model of the secondary flow circuit, Δp_s is fixed by the atmospheric pressure, Δp_f is the pressure source of the fan, R_H is due to the heat exchanger, and R_{in} and R_{out} , respectively, are the inlet and outlet flow resistances of the dryer itself. R_B refers to the varying flow resistance of the bed with thickness changing over time.

The system of equations of the MFN model consists of the nodal equations (Equation 13.38), the loop equations (Equation 13.39), and the branch equations (Equation 13.40). If the network contains m number of independent nodal points, i number of branches for each node, l number of independent loops, and j number of branches for each loop, the equation system can be written in the form

$$\left(\sum_i \dot{m}_i \right)_m = 0 \quad (13.38)$$

$$\sum_j [\Delta p_j + \Delta p(\dot{m}, T, n)]_l = 0 \quad (13.39)$$

$$\Delta p_j = R_j(\dot{m}_j, T) \dot{m} \quad (13.40)$$

The temperature-dependent elements of the MFN can be identified from the thermal subsystem model.

13.6.3.2 Thermal Subsystem Model of the Primary Circuit

The thermal subsystem model of the primary circuit consists of the models M of the collector M_C , the heat exchanger M_H , and the heat storage M_T , in accordance with the mode of operation. The modes of operation are as follows (Figure 13.35):

Mode I: Valves 2 and 3 are closed, and 1 and 4 are open; the air passing through the heat exchanger can be preheated by the water coming from the collector (see also Figure 13.34a).

Mode II: Valves 1 and 3 are closed, and 2 and 4 are open; the water in the tank can be heated by the spiral pipe P.

Mode I + II: By partial closing of valves 1 and 2 modes I and II can be maintained simultaneously.

Mode III: Valves 1 and 4 are closed, and 2 and 3 are open; collectors are not operated. The water of the tank is led to the heat exchanger of the dryer. (A is the auxiliary heater; see Figure 13.34a.)

13.6.3.2.1 Thermal Model for the Collector

The model of the collector can be built according to different model concepts. A detailed description of heat flow network model is given in Section 13.5.1. The equations of this model are given in Equation 13.30.

13.6.3.2.2 Thermal Model for the Heat Exchanger

The thermal model M_H of the water–air heat exchanger of the system outlined in Figure 13.34a serves to determine the outlet temperatures T_F'' and T_A'' of the working media, when the inlet temperatures and the thermal capacity flows $\dot{C} = \dot{m}c_p$ in modes I, I + II, and III are known. Because eventual transients are slow, the effect of the thermal capacities is disregarded.

Effectiveness H of the heat exchanger as a function of the ratio \dot{C}_A/\dot{C}_F must be known for the calculation. For an adiabatic heat exchanger,

$$T_A'' = T_A' + H(T_F' - T_A') \quad (13.41)$$

$$T_F'' = T_F' - H \frac{\dot{C}_A}{\dot{C}_F} (T_F' - T_A') \quad (13.42)$$

13.6.3.2.3 Model for Heat Storage

For the dryer in Figure 13.34a, the tank of the warm water system of the farm is used as heat storage. The emergency energy source A is built into the tank. The heat storage is of the well-mixed type. The model for heat storage serves for determining the outlet temperature T_{F2} of the collector working medium and the temperature T_T of the water in store. The equations of the storage with $C_T = M_T c_p$ at any time, for the case of $\dot{C}_c = \dot{m}_c d_p$ hot water consumption, are

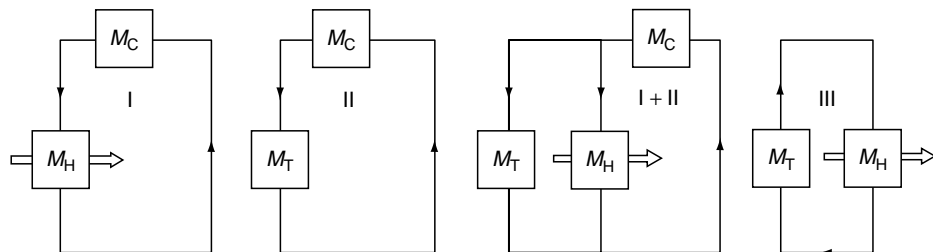


FIGURE 13.35 Thermal subsystem models according to the operating modes of the solar dryer.

$$C_T \frac{dT_T}{d\tau} + \dot{C}_c(T_T - T_{w,in}) = \dot{C}_F(T_{F1} - T_{F2}) + \phi_A + \phi_T \quad (13.43)$$

$$T_{F2} = T_T + (T_{F1} - T_T) \exp\left(-\frac{hA}{\dot{C}_F}\right) \quad (13.44)$$

where $T_{w,in}$ is the temperature of entering water, ϕ_A is the heat flow rate of the auxiliary energy source, ϕ_T is the rate of heat loss from the storage tank, T_{F1} is the temperature of water from the collector, and A is the heat transfer surface of coil heater.

13.6.3.2.4 Model for the Control System

The control strategy of dryer operation must be known for the simulation of the system, and the operating model for the control system has to be built into the simulation model. The model contains the condition system of the operation mode change, with the values of resistances R_1, \dots, R_4 (see Figure 13.34b) ordered to the corresponding points of time or to the limit values of the state characteristics of the working media of the material to be dried and of the atmospheric states. An advantageous solution is optimum control by sampling, using microprocessors (for optimum temperature control, see, e.g., Ref. [127]).

13.6.3.2.5 Coupling Equations

Insofar as heat losses from the pipes between the units are disregarded, the temperature of the medium leaving the preceding unit is equal to the inlet temperature of the adjoining unit:

$$T_{c,out} = T'_F \quad T_{c,out} = T_{F1} \quad T'_F \equiv T_{F1} \quad (13.45a)$$

$$T_{c,in} = T''_F \quad T_{c,in} = T_{F2} \quad T''_F \equiv T_{F2} \quad (13.45b)$$

If the heat losses from the connecting pipes are taken into consideration [45], further component models must be developed for determination of the losses. In this case the coupling equations are interpreted for the relation of loss part models and the models of the units. If the loss heat flow rate of the collector-storage pipe is ϕ_{C-T} , that of the storage-collector duct is ϕ_{T-C} and those of the collector-heat exchanger-collector pipes accordingly are ϕ_{C-H} and ϕ_{H-C} , respectively, and the coupling equations between the units can be written for the different modes of operation on the basis of the enthalpy balance equations

$$T'_F = T_{c,out} - \frac{\phi_{C-H}}{\dot{C}_F} \quad (13.46a)$$

$$T_{F1} = T_{c,out} - \frac{\phi_{C-T}}{\dot{C}_F} \quad (13.46b)$$

$$T_{c,in} = T''_F - \frac{\phi_{H-C}}{\dot{C}_F} \quad (13.46c)$$

$$T_{c,in} = T_{F2} - \frac{\phi_{T-C}}{\dot{C}_F} \quad (13.46d)$$

In the combined mode I + II of operation, the inlet temperatures can be determined from the mixing enthalpy balance equation.

13.6.3.3 Thermal Subsystem Model of the Secondary Circuit

The airflow circuit of the system in Figure 13.34a is open; thus the operation of the dryer does not react to the operation of the primary circuit. In the case of solar dryers, with partial recirculation of the drying air [25], the condition of the air entering the dryer depends on the operation of the dryer itself. Accordingly, with recirculating systems' dryer operation reacts on the primary circuit and the calculation of the primary and secondary circuit must be coupled.

13.6.3.3.1 Model for the Dryer

In this system the thermal model of the secondary circuit contains the heat exchanger part model M_H and the dryer part model M_D . Using the heat exchanger Equation 13.41, the inlet air temperature of the dryer can be determined; the solution of the MFN model equation system in Figure 13.34b gives the mass flow rate of the drying air \dot{m}_a . Thickness and initial state of the material in the dryer can be considered as given. For the simulation of static bed dryers, different methods are used [128–130].

In the drying space of the system in Figure 13.34a, the drying of alfalfa takes place in a static bed so that a new wet layer is laid on that already dried until the maximum layer thickness (approximately 6 m) is reached. Thus the thickness of the bed grows; consequently the value of the air mass flow rate belongs only to the thickness of the given layer of material (to R_B ; Figure 13.34c). Owing to the net weight of the layers laid one above the other, the porosity of the bed is not constant, either. The bed porosity, the heap density of the alfalfa, and the specific phase contact surface also depend on the thickness of the bed.

The model assumptions regarding alfalfa drying are as follows:

1. The drying of leaves and stems are modeled separately (two-component model: $k = 1, 2$).
2. The internal moisture-conduction resistance in the leaves and in the crushed stems is disregarded.

3. The alfalfa bed is divided into discrete layers along the height z , and within these layers the temperature T_k and material moisture content X_k are characterized by lumped values.
4. In the discrete layers the airflow rate is divided proportionally to the drying surfaces of the components.
5. The mixed mean state of the air leaving the components is regarded as standard for the state of the air entering successive layers (X_{av} , T_{av} , enthalpy).

The equation system describing the dryer model for a given elementary layer of the bed [25,125,131] is as follows: moisture mass balance for the material,

$$\frac{\partial X_k}{\partial \tau} = -\beta_k a_k (X_k - X_{e,k}) \quad (13.47)$$

enthalpy balance for the material,

$$\frac{\partial T_k}{\partial \tau} = \frac{h_k a_k}{c_k \rho_k} (T_a - T_k) - \frac{\beta_k a_k r}{c_k} (X_k - X_{e,k}) \quad (13.48)$$

moisture mass balance for the air,

$$\frac{\partial x_k}{\partial z} = A \frac{\sigma_k a_k}{\dot{m}_{a,k}} (x_{e,k} - x_k) - A \frac{\rho_a \xi_k}{\dot{m}_{a,k}} \frac{\partial x_k}{\partial \tau} \quad (13.49)$$

enthalpy balance for the air,

$$\begin{aligned} \frac{\partial T_{a,k}}{\partial z} = & \frac{a_k A}{\dot{m}_{a,k} c_{pa}} [\sigma_k (x_{e,k} - x_k) c_{pw} + h_k] (T_k - T_a) \\ & - \frac{A \rho_a \xi_k}{\dot{m}_{a,k}} \frac{\partial T_{a,k}}{\partial \tau} + \frac{(1 - \xi_k) b_k A \rho_k}{\dot{m}_{a,k} c_{p,a}} \end{aligned} \quad (13.50)$$

equation of desorption isotherms,

$$f(X_{e,k}, x_{e,k}, T_a, p_w, p_b) = 0 \quad (13.51)$$

The values of drying and material characteristics (σ_k , β_k , a_k , b_k , ξ_k , c_k , ρ_k , and sorption isotherms) in the equation system of the dryer model must be determined by experiment [132].

13.6.4 STRATEGY OF SOLUTION

Numerical solution of the system of equations above can be obtained by discretization in time [25]. The main steps of the solution process are as follows:

1. Data input, calculation of constant network elements and characteristics
2. Selection of time step
3. Solution of the working model of the process control system on the basis of the initial state; determination of mode of operation
4. On the basis of the initial state $T^{(k)}$, calculation of the dependent network elements and characteristics
5. Generation of the MFN models corresponding to the mode of operation
6. Solution of the MFN models for the subsequent period; determination of $\dot{m}_F^{(k+1)}$ and $\dot{m}_a^{(k+1)}$, for example by the Newton–Raphson method
7. Solution of thermal component models on the basis of $\dot{m}_F^{(k+1)}$, $\dot{m}_a^{(k+1)}$, and the initial state $T^{(k)}$ (for the collector, the finite time-element scheme see Ref. [31]; for the storage, the finite difference scheme can be applied); determination of the temperature of air entering the dryer
8. Solution of the equations for the dryer by applying an implicit finite difference scheme; determination of material temperature and moisture distribution in the bed

13.6.5 RESULTS OF SIMULATION

As an example, the results of simulation of the dryer in Figure 13.34a are presented [25]: the thickness of the fresh alfalfa layer in the bed, $z = 0.3$ m; the discretized layer thickness, $\Delta z = 0.075$ m; time step for collector $\Delta \tau_c^* = 0.5$ h, for the dryer, $\Delta \tau_d = 0.1$ h; air mass flow rate density, $\dot{m}_a = 0.171$ kg/s·m². After leaf wilting, \dot{m}_c (dry basis) for leaf is $x_1(0) = 1.47$ kg/kg; for stem, $x_2(0) = 1.62$ kg/kg. Air inlet temperature $T_{a,in}$ as a function of time and the drying curves are given in Figure 13.36. As can be seen from the

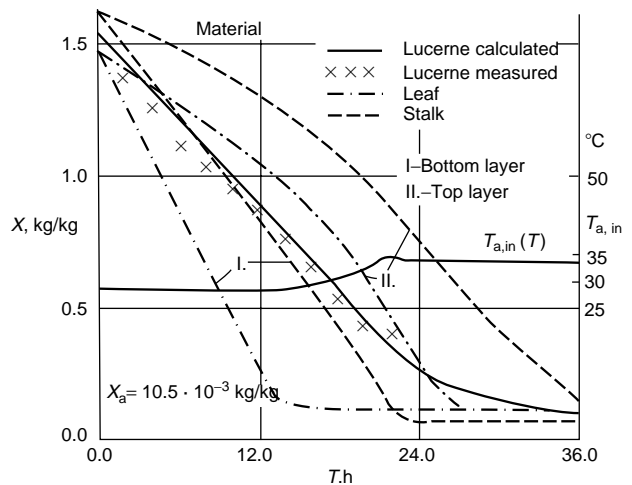


FIGURE 13.36 Drying curves of alfalfa.

figure, the drying rates for the leaf and stem are substantially different. In the top layer of the bed, the stem reaches $x_2 = 0.14$ kg/kg moisture content in about 42 h. Under the conditions given, drying can be carried out in 2-d cycles.

13.7 DIRECTION AND CONTROL OF SOLAR DRYERS

13.7.1 AIMS OF THE DIRECTION AND CONTROL

The direction and control of solar dryers aim to ensure the economical operation of the dryer in every stage of the drying process considering the actual state of the material under drying as well as the actual meteorological conditions [177].

13.7.1.1 Economy Aspects

The economic analysis of solar drying is presented in [Section 13.4](#). In this section economy aspects are interpreted in connection with the direction and control only. To fulfill the requirements of a good economy the savings should be increased and the costs should be reduced.

Utilizing solar energy for drying is not simply a method for saving conventional energy carriers but a technology for producing dried materials of high quality. This aspect is especially important when drying materials sensitive for quality deterioration. Products serving human and animal foods are generally very sensitive and their main characteristics (i.e., color, smell, taste, shape, nutritive, and other internal substances) are highly dependent on the thermal and sorption history of the material from the harvesting to the preserved state and on the time interval of drying. In Table 13.3 drying data are given for some agricultural products. Cereal grains and grain legumes need to be dried from an initial moisture content of about 30% (wet basis) at harvest to a level of 12%. Leafy green vegetables and fruits have an initial moisture content of about 60–80% to be reduced to the range of 10–25% for safe storage. Safe drying air temperatures are of 35–60°C and for some products these temperatures are higher at the end phase of the drying.

The quality of the dried products has an effect on the economy by influencing the marketing capability and income of the products because a higher price can be achieved by better quality. To ensure the required preconditions for drying of such sensitive materials a technological direction and process control is needed. The loss in quality of the dried product should be considered as a saving of negative value [176].

TABLE 13.3
Drying Data for Some Agricultural Products

Product	Moisture Percent (wb)		Drying Air Temperature (°C)
	Initial	Final	
Bananas	80	15	70
Barley	18–20	11–13	40–82
Beets	75–85	10–14	—
Cardamom	80	10	45–50
Cassava	62	17	70
Chilies	90	20	35–40
Coffee seeds	65	11	45–50
Copra	75	5	35–40
Corn	28–32	10–13	43–82
Cotton	25–35	5–7	—
French beans	70	5	75
Garlic	80	4	55
Grapes	74–78	18	50–60
Green forages	80–90	10–14	—
Hay	30–60	12–16	35–45
Longan	75	20	—
Medicinal plants	85	11	35–50
Oats	20–25	12–13	43–82
Onions	80–85	8	50
Peanuts	45–50	13	35
Pepper	80	10	55
Potato	75–85	10–14	70
Pyrethrum	70	10–13	—
Rice	25	12	43
Rye	16–20	11–13	—
Sorghum	30–35	10–13	43–82
Soybeans	20–25	11	61–67
Spinach leaves	80	10	—
Sweet potato	75	7	75
Tea	75	5	50 ^a
Virginia tobacco	85	12	35–70
Wheat	18–20	11–14	43–82

^aAt the end of drying, for 2–3 h 100°C.

Source: From Mahapatra, A.K. and Imre, L., *Int. J. Ambient Energy*, 10(3), 163, 1989.

One of the main components of savings for semi-artificial and solar-assisted dryers is the price of the conventional energy carriers substituted by solar energy. It should be emphasized that the solar energy utilized by the solar dryer is not equal to the energy collected by the solar collector and transferred into the drying air but the energy effectively used in the drying process. The energy effectiveness of solar dryers depends also on the exit energy losses, which can be reduced by applying a proper direction and control strategy. In the case when the energy effectiveness of a given solar dryer is lower than that of a

conventional one a negative value in savings should also be considered.

The possible savings and the investment and maintenance costs of solar dryers are interdependent. Well-directed solar dryers need higher costs. Inexpensive, simple, and unsophisticated solar dryers, generally, have no appropriate devices for direction and control. These types of solar dryers are of great importance first of all for country use, for substituting open-air drying of nonsensitive materials and eliminating the well-known disadvantages of natural drying.

Solar dryers of high performance or for drying of quality-sensitive materials should be well directed. The higher investment and maintenance costs can be balanced by the better quality of the dried product and by the longer annual operation time. The annual operation time can be extended when drying materials of long growing time and several harvests in a year (e.g., meadow grass, alfalfa) or in cases of solar dryers applicable for drying of various materials having differing ripening times, one after the other (e.g., herbs, medicinal plants, spices and aromatic plants, seeds). Another way for year-round utilization is the multipurpose application of the solar energy converter of the dryer when, in the idle periods of drying, the solar energy collected is utilized for other technological purposes (e.g., for satisfying technological hot water demands of an agricultural farm). These complex systems should have appropriate direction and control devices to realize an economical operation strategy.

13.7.1.2 Strategy of Direction

The strategy of the direction should be elaborated by taking into consideration the drying characteristics of the material and the regulation possibilities of the solar dryer to be used. In the knowledge of the drying characteristics of the materials to be dried (i.e., sorption isotherms, drying curves) the appropriate schedule of the drying operation should be elaborated and, using it as a basis, the possible methods for interventions should be determined. Even in case of the most simple solar dryers some methods in the direction are recommended.

13.7.2 DIRECTION AND CONTROL ACTIONS

13.7.2.1 Direction of Drying Operation

Direction includes actions that are required for the realization of an appropriate drying process in the dryer (i.e., to follow with attention the actual state of the material under drying) and determine and execute the necessary interactions by applying a direction strategy.

The first phase of the drying to be directed is the feeding of fresh material into the dryer. It should be

emphasized that the good quality of the fresh material is a precondition of the good quality of the dried product. This action should be in harmony with the ripening state of the material and the point of views of the drying should be asserted in the harvesting technology.

The main direction actions of the drying process can be summarized as follows:

1. Feeding fresh material into the dryer
2. Turning or tedding the layer of the material under drying occasionally in the case of unsophisticated (e.g., tent type) dryers
3. Regulating the airflow rate
4. Regulating the recirculation of the air
5. Regulating the intermittent drying process (determining the beginning and the interval of the break)
6. Separating the solar dryer from the atmosphere in the night and in rainy weather when no auxiliary energy source exists
7. Regulating the operation of the auxiliary energy source
8. Distribution of solar energy collected inside the drying space and between the cells in the case of multicell solar dryers
9. Determining the mode of operation in the case of complex and multipurpose solar dryers and ensuring the optimal distribution of solar energy collected between the dryer, the storage, and the other heat consumers
10. Regulating the operation of storage in the case of solar dryers with heat storage
11. Determining the appropriate inlet temperature of the drying air in the different stages of the drying process

13.7.2.2 Control of Drying Operation

Control actions of the drying operation are concerned with holding the given values of some operational parameters determined by the direction strategy. The main control actions are as follows:

1. Temperature control of working mediums
2. Relative humidity control
3. Mass flow rate control of flowing mediums
4. Switch in and out devices (e.g., fans, humidifiers, valves or dampers, auxiliary heaters) when the limit values of some parameters occur
5. Control of charging the thermal storage of the system
6. Control of the rate of drying
7. Control of the recirculation
8. Control of the intermittent drying process

13.7.3 PRINCIPLES OF THE DIRECTION

13.7.3.1 Direction Strategy of Static Bed Solar Dryers

For the drying process with forced convection in a static bed, the static bed is generally arranged in a drying chamber and the heat is transferred from the drying medium to the material by convection. In through-flow dryers air is led below the layer and flows upward through the bed.

The state of the material to be dried in a static bed can be approximately characterized by the change of state of the drying air flowing through the bed. The change of state of the air can be followed in the enthalpy (h) and absolute moisture content (x) chart of Mollier.

In Figure 13.37 the h - x diagram for an open-cycle drying process is presented. The actual state of the atmospheric air is represented by 0. Supposing that the mass flow rate of the air is \dot{m} , heat flux into the air is ϕ_h , the temperature increase of the air can be calculated:

$$\Delta t = t_1 - t_0 = \frac{\phi_h}{c_p \dot{m}} \quad (13.52)$$

where c_p is the specific heat capacity of the air. The temperature of the air entering the dryer is t_1 . Supposing further that the drying by convection is nearly adiabatic, the change of state of the air in the bed is approximately of $h_1 = \text{const.}$ The air will approximate the equilibrium relative humidity U_e of the material and the absolute moisture content of the air will increase by Δx (in Figure 13.37, point 2). The mass

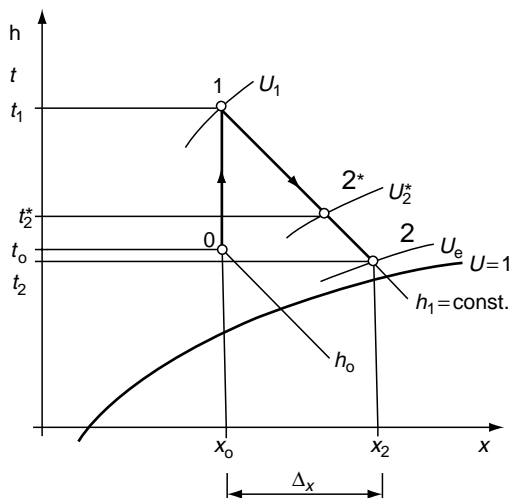


FIGURE 13.37 Change of state of the air in an open-cycle drying process in h - x chart.

flow rate of evaporation N from the material can be expressed indirectly:

$$N = \dot{m} \Delta x \quad (13.53)$$

and the energy flux practically consumed for drying is

$$\phi_D = \dot{m} \Delta x r = N r \quad (13.54)$$

where r is the total heat of evaporation of the water from the material under drying. The ϕ_D value can also be expressed by the temperature difference of the air:

$$\phi_D = \dot{m} c_p (t_1 - t_2) \quad (13.55)$$

When the material to be dried has a quasiconstant rate period between the initial integral moisture content W_0 and the first critical moisture content $W_{cr,1}$ the drying rate curve $N(W)$ has the shape given in Figure 13.38. In the figure the temperature curve $t(W)$ is also given. In the constant rate period the temperature of the material approximates the wet bulb temperature t_{wb} and remains practically constant ($t_m \cong t_{wb} = \text{const.}$). In the falling rate period the temperature of the material will approximate the dry bulb temperature t_1 .

The moisture distribution in the bed of a thickness Δz can qualitatively be characterized by the curves in Figure 13.39 given for the different stages of drying. For time $\tau = \tau_1$ from the beginning, the upper layer of the bed above point 1 is almost of the initial moisture content. Drying of the surface layer starts at time $\tau = \tau_2$ and will be continued with approximately constant rate until the moisture content of the surface layer will be reduced to the first critical ($W_{cr,1}$) value $\tau = \tau_3$. In time interval $\tau < \tau_3$, ϕ_D remains

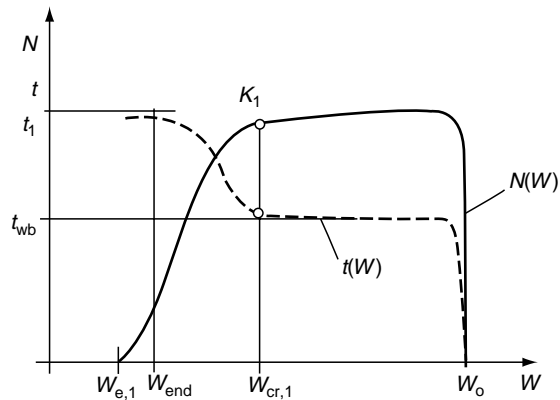


FIGURE 13.38 Typical drying rate and temperature curves of materials having constant drying rate period.

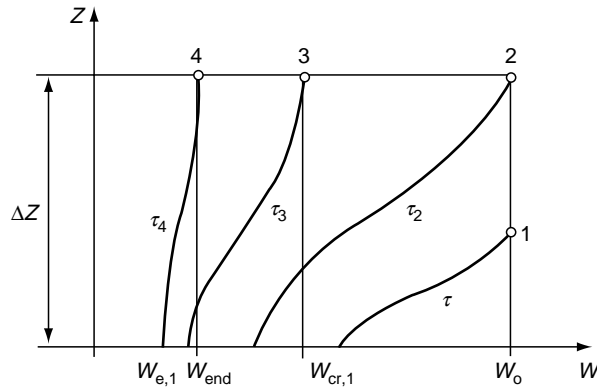


FIGURE 13.39 Moisture distribution in a static bed at the different stages of drying.

practically constant, not considering a short starting period. When $\tau > \tau_3$ (i.e., in the falling rate period of drying), t_m and t_2 will increase $t_2 \rightarrow t_2^x$ (see Figure 13.37) and, according to Equation 13.55, the exit energy loss of the dryer $\phi_L = \dot{m}c_p(t_2^x - t_2)$ will increase. Because of the increasing energy loss the effectiveness of the dryer will successively decrease in the falling rate period ($\tau_3 < \tau < \tau_4$).

13.7.3.2 Direction of Operation of Static Bed Dryers

Considering the drying process in a static bed described qualitatively before, conclusions for the direction can be summarized as follows.

1. In Figure 13.40 the $\phi_D(\tau)$ function is presented at constant inlet parameters t_1 , U_1 for two different bed thicknesses. In the case of a thin bed (Figure 13.40a) τ_2 is almost negligible (for τ_2 see also Figure 13.39), $\tau_3 - \tau_2$ is the time interval of the practically constant rate period, $\tau_4 - \tau_3$ is that of the falling rate period. By increasing the thickness of the bed τ_2 will increase, $\tau_3 - \tau_2$ and $\tau_4 - \tau_2$ intervals remain

practically constant (Figure 13.40b). The energy effectiveness of the dryer (taking ϕ_h as constant)

$$e = \frac{\phi_D}{\phi_h} \quad (13.56)$$

is the function of the time. For the time interval of drying τ_4 the average value of e is

$$e_{av} = \frac{1}{\tau_4 \phi_h} \int_0^{\tau_4} \phi_D(\tau) d\tau \quad (13.57)$$

The e_{av} can be increased by increasing the thickness of the bed (see the ratio of the dark and white areas of Figure 13.40a and Figure 13.40b).

2. Though by increasing the layer thickness the energy effectiveness of the dryer can be improved, another effect has also to be taken into consideration. Since the time interval τ_2 will be longer, this method is not advantageous for materials sensitive to quality deterioration because the loss of internal substances is directly proportional to the time passed from the harvesting to the preserved state. For the material situated in the upper layer of the bed, drying will start after a longer time and some deterioration may occur. To solve this contradiction in the economy requirements, drying should be started with a fairly thin layer thickness. The thickness of the bed should be increased from time to time by feeding successively a new fresh layer in at $\tau = \tau_3$. This multilayer feeding method can be applied when the harvesting can also be fulfilled successively (e.g., in case of drying meadow grass or alfalfa in a solar drying-storing barn). In this case the task of the direction is to determine the time points of τ_3 . Observation of τ_3 is possible with a fairly good approximation by measuring continuously the temperature t_2 and the relative humidity U_2 of the air leaving the bed.

3. Exit energy loss can be reduced by the multilayer feeding also in case of materials not having a

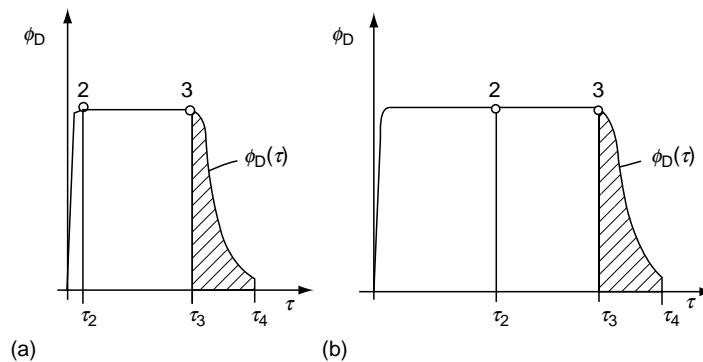


FIGURE 13.40 Energy flux consumed for drying in the function of time: (a) thin bed; (b) increased bed.

constant drying rate period at all. As it can be seen from Figure 13.41, by feeding new layers in after the time intervals of $\Delta\tau$, the value of e_{av} will be much higher than that of the single processes fulfilled one after the other.

4. Another well-known direction method is the application of intermittent operation in the falling rate period of drying (Figure 13.42). The first drying interval of $\Delta\tau_{1d}$ is followed by a break of $\Delta\tau_{1b}$ and so on ($\Delta\tau_d - \Delta\tau_b, \dots$). During the break the uneven moisture distribution inside the material tends to equalize and, in the next drying interval, the drying rate will be higher again. This way e_{av} can be increased and the electrical energy consumed for driving and fan can be reduced. To find the best combinations of the drying periods and the breaks, an optimization problem has to be solved. For optimization the drying curves of the material, the time function of the internal moisture equalization process at different integral moisture contents, and the technical data of the dryer itself are needed. The direction strategy can be elaborated by computer simulation and proved by experiments.

For solar dryers not having any heat storage the utilization of the solar energy collected in the breaks is a problem. One possibility is to time the break intervals in the night. This condition in some cases will not permit fulfillment of the optimal strategy.

5. Another possibility is to divide the drying space into individual cells (see Figure 13.10). Each cell should have its own fan and the construction should permit the distribution of the solar energy collected between the cells in an optional ratio. The intermittent drying processes in the different cells are shifted in time, making it possible to utilize more solar energy in one cell, whereas in another cell a break is in progress.

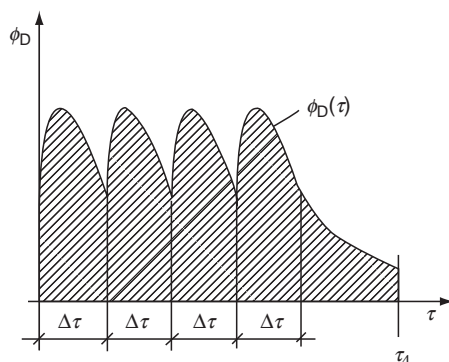


FIGURE 13.41 $\phi_D(\tau)$ function with multifeeding in for drying materials not having any constant rate period.

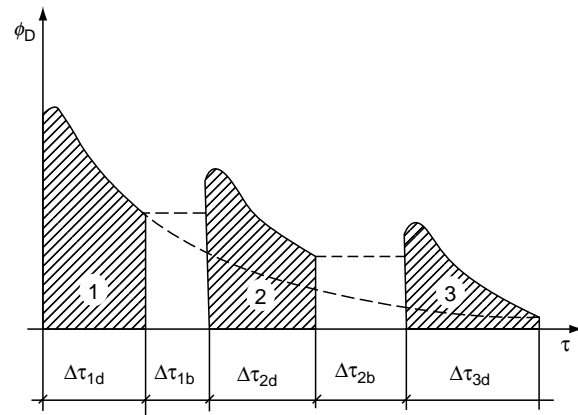


FIGURE 13.42 $\phi_D(T)$ function for intermittent operation.

6. In the falling rate period of drying, the energy effectiveness of the solar dryer can be improved by recirculating one part of the air leaving the material. This well-known method can be recommended, first of all, for drying materials of long drying time.

A simplified scheme of a solar dryer having a separated collector is presented in Figure 13.43. Part of the recirculated air can be regulated by valves 4 moving together built in the air ducts serving for the outlet air 5 and for the recirculating air 6. Airflow of the air duct 6 is mixed with fresh air before entering the collector 1. The mixed air will be preheated in the collector and transported by the fan 2 to the drying chamber 3. The recirculated air can also be mixed with the fresh air preheated by the collector before the fan (dotted line 6*).

The principles of the direction can be followed in an $h-x$ diagram (Figure 13.44). The state of the fresh air is represented by 0, the air preheated in the collector (without recirculation) by 1 and, the state of the air leaving the dryer by 2. In case of applying recirculation in a proportion of $(b/a) \dot{m}$, the state of the mixed air will be of M . Flowing through the collector the air will be preheated to t_1^* (point 1*). Supposing that the temperature of the material is t_2 (point 2

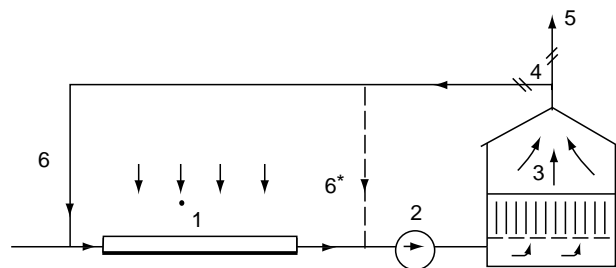


FIGURE 13.43 Simplified scheme of a solar dryer with recirculation.

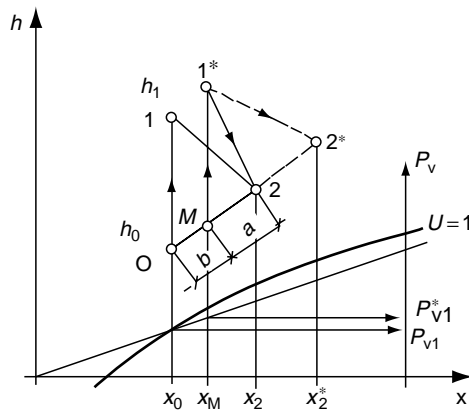


FIGURE 13.44 Change of state of the air with recirculation and mixing before the collector.

represents the state of air in equilibrium with the material), the direction of the change of state of the air will be of 1^*2 . As a result of the increased vapor pressure of the air (p_{v1}^*) the rate of drying will temporarily be decreased and the temperature of the material will increase. The state of the air leaving the material will tend to 2^* .

Applying recirculation the material can be heated to a higher temperature level than that without recirculation, where the rate of drying will be higher. The operation can be directed by regulating the b/a ratio. In the course of drying, temperature of the material tends toward 1^* . This fact has to be considered when regulating the recirculation.

In the case when the recirculated air is mixed with the air leaving the collector (see duct 6^* in Figure 13.43) the airflow rate in the collector will decrease and the temperature t_1^* will increase. Air temperature depends on the b/a ratio (see Figure 13.45). The mixing point M will represent the state of the air

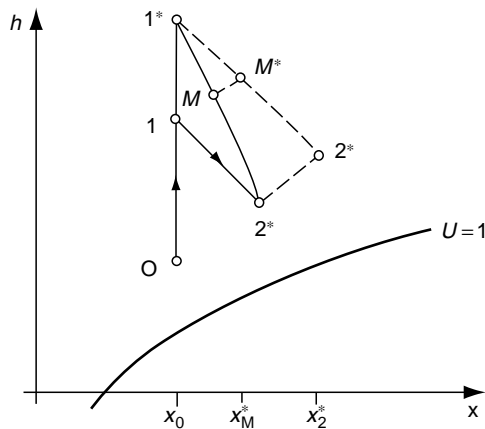


FIGURE 13.45 Recirculation with mixing after the collector.

entering the dryer and it will be situated between 1^* and 2. The position of M will change in the function of the material temperature (2 tends to 2^*). The possibilities for drying will be similar to the previous version. In both cases the energy effectiveness of the dryer will be higher in spite of the greater heat losses in the collector and the drying chamber. A disadvantage of this system is the additional cost of the air duct 6. By applying a special construction duct 6 can be eliminated. As an example the solar timber dryer presented in Figure 13.6 can be considered.

The direction of the operation of such a dryer is, principally, the same as that written before. Representing the process in an $h-x$ diagram, some differences arose from the facts that the stack is partially irradiated by the absorber and the airflow contacts a differing area of the absorber surface before entering the stack. To elaborate the direction strategy of solar dryers operating by recirculation, experiments are recommended.

13.7.3.3 Direction of Solar Dryers with Heat Transfer by Convection and Direct Irradiation

In Figure 13.46a, a simplified scheme of a solar tunnel dryer operating with convection and direct irradiation [184,203] is presented. A single-covered solar collector having a transparent covering is connected to the drying space. In the drying space the material to be dried is arranged in a fairly thin layer and, through the transparent covering, it is directly irradiated. Drying air is transported through the collector and the drying space by a fan. In rural areas the fan can be driven by the electricity produced with photovoltaic modules [226].

The operation of the dryer can be followed in an $h-x$ psychrometric chart (Figure 13.46b). In the collector the atmospheric air is heated to the drying temperature t_1 . In the drying space temperature t_1 is approximately constant if the heat demand of evaporation is satisfied by direct irradiation. The state of the air leaving the dryer is represented by point 2. The rate of drying is $N = \dot{m}\Delta x$, where \dot{m} is the mass flow rate of the air. During the sunny hours of the day the state of the atmospheric air is 0 and the solar irradiation is changing over time. As a consequence, t_1 and t_2 temperatures are also varying. Nevertheless, the character of the change of state process of the drying air can be represented by the line 012 in Figure 13.46b.

Let us suppose that the temperature of the material—and of the air in equilibrium with it—is t_2 when, in the evening, the solar irradiation stops. If the fan

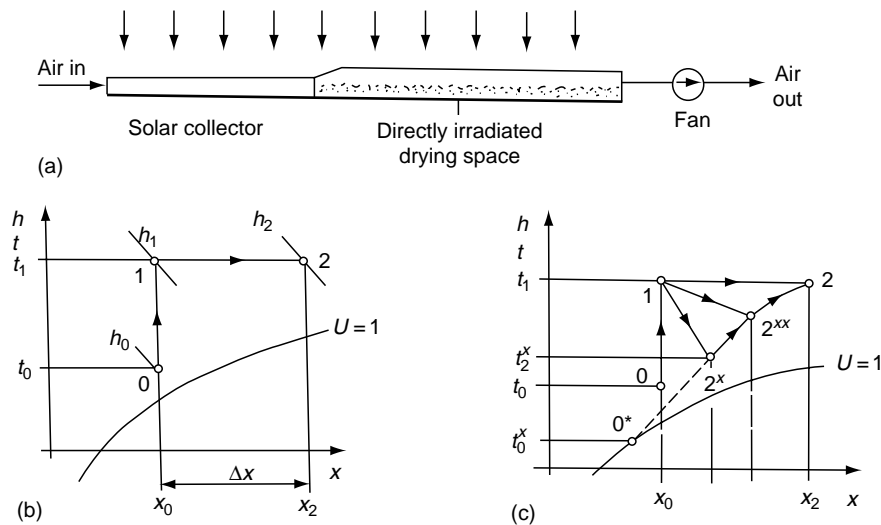


FIGURE 13.46 Solar dryer with heat transfer by convection and direct irradiation: (a) simplified scheme of the dryer; (b) change of state of the air in h - x chart; (c) night operation.

will be stopped, too, the material having a higher partial vapor pressure will keep evaporating. Without ventilation the water evaporated from the material will partially condense on the internal surface of the covering. In the night the atmospheric temperature will decrease (t_0^x , see Figure 13.46c) and the covering will cool down by convection and radiation. To avoid the condensation and the possible damage to the quality of the wet material it may seem to be advisable—at least in the first stage of drying—to keep the fan in operation also in the night. As a consequence of the evaporation and of the cooling effects the temperature of the material will decrease. If the temperature decreases below the atmospheric temperature that existed in daytime ($t_2^x < t_0$), next morning the partial vapor pressure of the preheated air will be higher than that of the material and, for a time interval, the material will absorb some water vapor from the air (rewetting effect). The temperature of the material will increase by the convective heat transfer from the air as well as by the direct irradiation of the sun (from t_2^x to t_2).

13.7.2.2.1 Possible Methods for Direction

During drying, the rate of evaporation on the surface of the layer will be higher than that in the bottom of the layer. In the falling rate period the temperature of the surface layer will increase over the temperature of the air stepping in the drying space from the collector. Heat transfer from the surface into the layer will increase and, at the same time, a part of the energy gained by direct radiation will be transferred into the air by convection. As a consequence, the energy effectiveness of the dryer will decrease. The

disadvantageous effects can be avoided by turning the layer over from time to time. This operation needs some handwork and can be applied only for materials not sensitive to the mechanical effects of turning or mixing. Without turning, the overheating of the surface should be prevented by the appropriate design of the dryer.

In the night, the cooling down effects can be moderated by reducing the mass flow rate of the air. It can be realized by a throttling or by using a driving motor of variable speed. Heat losses can be reduced by shading the transparent covering. A further possibility is the application of a solar collector with heat storage. A heat storage pebble bed or latent heat storage can be used. The surface of the storage layer may serve as absorber. Application of heat storage results in more sophisticated construction and higher investment costs. Economy aspects should be determined in each case separately.

13.7.3.4 Direction of the Operation of Tent, Greenhouse, and Cabinet-Type Solar Dryers

Tent, greenhouse, and cabinet-type dryers have simple and unsophisticated structure. The material to be dried is partially or totally irradiated during drying. Problems that may arise during the operation of such dryers are similar to those discussed in the previous section. The possible means of direction of the operation are also the same: turning the material over from time to time and to close and shadow the drying space during the night or, in some cases, to reduce the airflow in night operation.

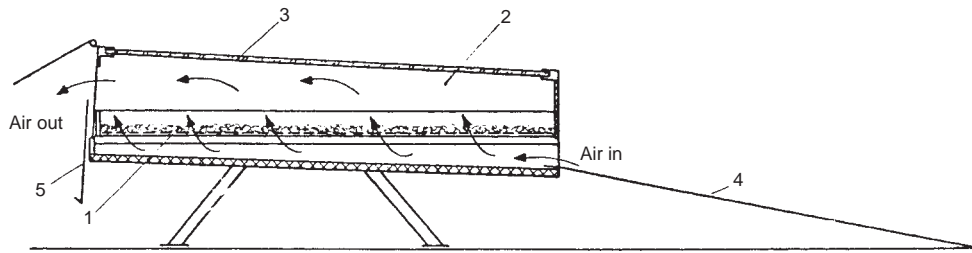


FIGURE 13.47 Scheme of a solar cabinet dryer for household use.

In Figure 13.47 a scheme of a cabinet-type dryer for household use is presented to dry fruits, herbs, and vegetables. Material to be dried is arranged in a thin layer on a tray 1 with perforated bottom inside the cabinet 2 with a single glass covering 3. The air is introduced to the dryer by free convection using a black-painted metal sheet 4 of the same surface area. The metal sheet serves as a secondary uncovered collector by increasing the inlet air temperature. The airflow can be controlled by the throttling device 5 at the air outlet. During the day the position of the dryer can be changed according to the position of the sun. At the first stage of the drying the slider 5 is open. The outlet cross-sectional area of the air should be decreased according to the progress of the drying process. Thus, the airflow rate will also be decreased and the temperature of the air increased. In the night, the material should be shaded by posing the black-painted sheet 4 on the transparent covering of the cabinet 2 and having the slider 5 closed.

13.7.3.5 Direction of the Operation of Chimney-Type Solar Dryers

In chimney-type solar dryers the driving force of the airflow is the hydrostatic pressure difference caused by the decreasing density of the preheated air (chimney effect). Since no conventional energy sources are needed, chimney-type solar dryers can effectively be used as country dryers. For keeping the airflow on in the night, chimney-type dryers usually have some kind of heat storage.

In Figure 13.5, the construction of a chimney-type solar dryer [6] is presented. During night operation the transparent walls 6 can be insulated by reflecting panels. The air duct of the collector should be closed and air ducts [5] below the drying chamber should be opened. The atmospheric air will be preheated by convection when flowing through the heat storage space and contacting with the water containers.

Another construction is presented in Figure 13.48 [178,214]. This dryer has a collector 2 with a latent

heat storage material ($\text{CaCl}_2 \cdot 6\text{H}_2\text{O}$) as absorber 4. Two additional latent heat storage plates are applied [5] that are pulled out from the covering of the collector and are directly irradiated in daytime. The southern walls, including the wall of the chimney, are made of transparent material; the northern wall and the bottom 6 are insulated. The material to be dried is arranged on trays 7 in the drying chamber 1 and partially irradiated. On the northern wall of the chimney 3 a layer of latent heat storage is built in 4 that produces additional preheating effects for the air in night operation.

As to direction of the operation the latent heat plates 5 should be pushed below the covering of the collector for night operation (see the figure). Since the heat storers are directly irradiated, they can be charged to a higher temperature level than they could be when heating by the air. The airflow through such dryers can be assisted by a windmill driving a fan built in the chimney.

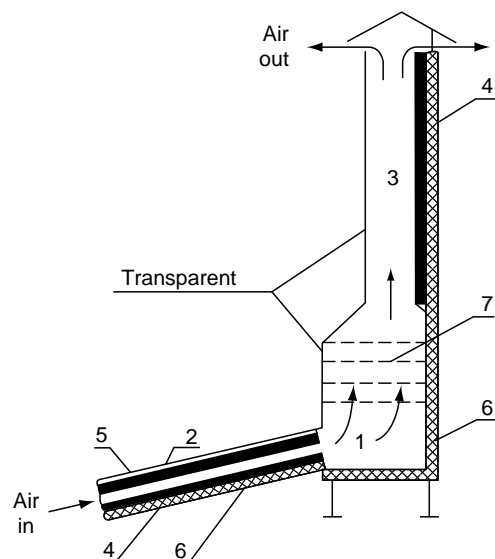


FIGURE 13.48 Simplified scheme of a chimney-type solar dryer with latent heat storage.

13.7.4 BASIC PRINCIPLES OF CONTROL

13.7.4.1 Application of Automatic Control

13.7.4.1.1 Economy Aspects

Values of the operational parameters required by the direction strategy can be effected automatically by using controllers. Application of automatic control can be justified by the possible reduction of workforce as well as by the higher reliability compared to manual control. Though the instrument cost is high, its effect on economics can be balanced by the better quality, by the better energy effectiveness, and, in case of sophisticated solar dryers of high performance, by the improved security of the drying operation. Economy analysis has to be done in each case separately.

13.7.4.1.2 Aspects for Designing the Control

Control systems generally consist of three main elements. A sensing element (sensor) is used for measuring the actual value of the parameter to be controlled. The control device serves for forming commands for intervention, if necessary. The element for intervention (e.g., a motoric valve) executes the command of the controller.

Process control systems generally operate with a constant set point. It should be emphasized that exact control is not always necessary and overinstrumentation should be avoided. Solar dryers generally have some parts of large capacities that reduce the effects of disturbances.

For the accurate design of the control system a detailed operational analysis is needed. The requirements for the design can be determined on the basis of the information produced by the analysis. In simple cases, the appropriate controller can be selected by using rules of thumb. Solar dryers having various input parameters can often be controlled by individual controllers. In cases when the control aims at optimizing the drying operation, commands for interventions should be formed by considering several input and output variables. For this purpose a micro-processing control device can be applied.

The detailed description of process control theory is beyond the scope of this chapter. In actual designing problems a consultation with experts in process control is recommended. A brief overview of the main types of controllers is offered here (for further details, see Refs. [179–181]).

13.7.4.1.3 Main Types of Control Systems

On–off control is the most simple and the cheapest method and it is widely used. If the value of the measured variable is less than the set point value, the controller is on. The output signal of the controller is a

given value. When the measured variable is above the set point, the controller is off. In solar dryers on–off temperature controllers are used (e.g., for control of the operation of auxiliary heaters in a water storage tank). A disadvantage of this control system may be the uncertain operation and, actually, some overshoot may occur.

Closed-loop or feedback control systems operate by adjusting automatically one of the input variables of the process by comparing a signal fed back from the output of the process with a reference input. The difference serves as signal for the controller. The system can be characterized by the transient response of the output of the process due to some specific variations in the input. The change in input may be either a change in the set point or in one of the load variables (e.g., uncontrolled flows and temperatures). Two different operations can be realized. With servo operation the aim is to follow changes in the set point. With regulator operation the output of the process should be kept constant in spite of some changes in load variables.

Open-loop control systems are used when every input variable of the process should be constant. Open-loop control can effectively be used when a closed control is not needed, when the change in inputs is not strong, or in cases when the feedback control is not good enough. The open-loop control is called feed-forward control when one of the input variables is measured and used for adjusting another input variable.

13.7.4.1.4 Main Types of Control Actions

In Figure 13.49, a block scheme of a control system is presented. In the case when the set point value x_0 is constant the control is called value keeping. If the set point is a function of time, that is, $x_0(\tau)$, a signal is needed for operating the set point device. Various principles can be used to form commands for interventions by the controller. The basis of the methods is the error e , which is the difference between the controlled parameter (control signal x_c) and the set point

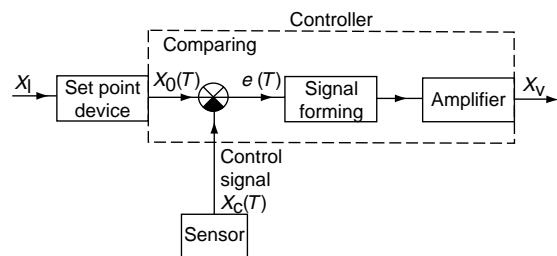


FIGURE 13.49 Block scheme of a control system.

x_0 : $e = x_0 - x_c$. Controllers have a signal-forming unit that produces the command signal as output x_i .

The proportional controller (P) produces its output proportional to the error

$$x_i = K/x_0 - x_c/ = Ke \quad (13.58)$$

where K is the gain of the controller. At a given working point, the change in the output related to a differential change in input is called the gain of the controller.

With integral control (I) output of the controller is proportional to the time integral average value of the error

$$x_i = \frac{1}{\tau_R} \int_0^{\tau_R} e \, d\tau \quad (13.59)$$

where τ_R is the reset time.

The two-mode proportional integral (PI) controller produces its output by addition:

$$x_i = K \left(e + \frac{1}{\tau_R} \int_0^{\tau_R} e \, d\tau \right) \quad (13.60)$$

Derivative control action (D) can improve the response of slow systems when coupling parallel to proportional control by adding an effect proportional to the time derivative of the error. This way some disadvantageous effects of large load changes and the maximum error can be reduced.

The three-mode controller has a proportional and an integral character with derivative action (PID). The output signal of a PID controller is

$$x_i = Ke + \frac{1}{\tau_R} \int_0^{\tau_R} e \, d\tau + \tau_D \frac{de}{d\tau} \quad (13.61)$$

where τ_D is the derivative time.

The output signal x_i of the signal-forming unit will be amplified and modified. The output of the controller is the signal for intervention x_v . As an example the scheme of the automatic temperature control of a liquid–air heat exchanger is given in Figure 13.50.

13.7.4.1.5 Selection of Control Systems

In the operation of control systems stability is required. Operation is stable when continuous cycling will not occur. Instability could be the consequence of the increase in the overall gain of the controller above a maximum value. The overall gain of the controller is the product of gain terms in a closed loop. The role of the time lag may also be considered. Different stability criteria have been elaborated and various rules developed. Integral control and derivative control

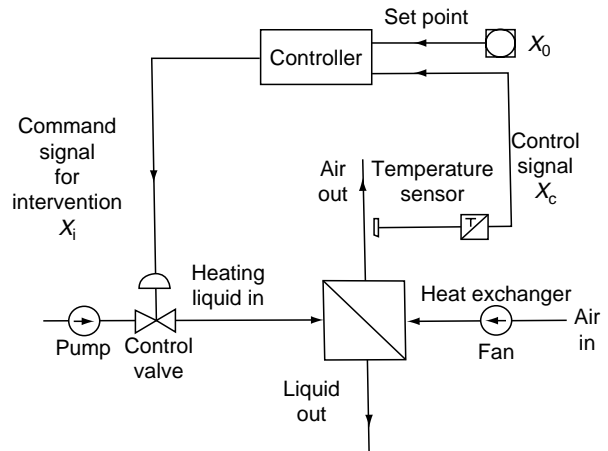


FIGURE 13.50 Control scheme of a liquid–air heat exchanger.

action can improve the stability of systems added to proportional control. The final performance of a system is affected by the characteristics of the process to be controlled, by the operational characteristics of the controller used, and by the nature of the disturbances to be expected.

Operational characteristics can be described by the response function of the system $x(\tau)$ to a step change in load. The system consists of the process to be controlled and of the controller. The mathematical relationship between input and output is called a transfer function. The time τ_c necessary to approximate the required value within a given difference Δx (mostly, $\Delta x = \pm 0.02 \dots 0.05$) is called control time. The response function of the controlled process (not having any integrating character) can be characterized by the time lag τ_1 and by the time constant τ_t .

For the selection of the appropriate control system a dynamic analysis is needed that can inform the designer about the type of controllers really needed. Some general recommendations follow. In cases when the expected disturbances are strong and the control time permitted is long enough, proportional (P) controller can be used. PI controllers can be applied with $\tau_c \tau_1 > 6$. PID controllers are recommended in the range of $4 \tau_1 < \tau_c < 6$. For flow rate control and for level control P and PI controllers are used; for temperature control, P, PI, and PID controllers are mostly applied.

13.7.4.2 Control of Solar Dryer Operating with Recirculation

With solar dryers of multicell construction the collector field serves for more than one drying chamber. Recirculation of the drying air can be directed for keeping the inlet air temperature constant. The solar energy saved by recirculation can be utilized for

preheating the air of another drying chamber (or for other technological purposes). In these cases the aim of applying recirculation is the improvement of the energy effectiveness of each dryer and, by this way, of the system. As it can be seen from Figure 13.44, constant inlet air temperature can be realized by the appropriate variation of the b/a ratio. A precondition of such operation is that the point 0 (i.e., the state of the atmospheric air) should be of lower absolute water content than that of the outlet air. The scheme of the control is presented in Figure 13.51 for a multi-cell solar dryer having a liquid-type collector and a liquid–air heat exchanger. (In the figure one drying chamber is indicated.)

Operation by recirculation can be realized by controlling the b/a ratio and the heat input to the heat exchanger. By controlling the air valves moving together (see also 4 in Figure 13.43) the absolute moisture content of the air x_1 can be ensured (according to the position of the mixing point M). The controlling signal for the recirculation is the wet bulb temperature T_{wb} of the inlet air. Inlet temperature of the drying air is controlled by the valves V of the liquid working medium of the collector serving as heating (primary) medium for the heat exchanger. The controlling signal is the dry bulb temperature ($T_{db} = T_1$). The control action can be realized automatically or manually.

13.7.4.3 Direction and Control of Solar Dryers with Rock-Bed Heat Storage

For dryers having separated rock-bed heat storage (see Figure 13.9) three main modes of operation should be applied:

1. Drying with air preheated by the collector
2. Drying and simultaneously charging of the heat storage with the air preheated by the collector

3. Drying with air preheated by the heat storage (discharging period) when no solar radiation exists

In mode of operation 1, damper 8 in Figure 13.9 is open whereas damper 9 closes the air duct below the heat storage 7. In mode of operation 2, damper 9 closes the upper air duct and the air flows from the drying space into the rock bed. In mode of operation 3, damper 9 is in a medium position, fan 2 is out of operation, and the air flows from the dryer into the rock bed; there it will be preheated and flown back into the drying space in the upper air duct.

The regulation of the mode of operation can be realized manually or automatically. As a signal for regulation, the temperatures of the drying space, the outlet air of the collector, and the rock bed can be used. As a controlling signal for the operation of the dampers 6, the wet bulb temperature or the relative humidity of the drying space can be applied.

13.7.4.4 Automatically Controlled Solar Dryer with Auxiliary Heater

13.7.4.4.1 Construction of the Dryer

The simplified scheme of an automatically controlled solar lumber kiln dryer with an auxiliary energy source of wood residue burner is shown in Figure 13.52 [182]. Collector 1 has a charcoal absorber and a gravel-bed storage is arranged below it. Airflow through the collector 1 is induced by two blowers 2. Four collectors are coupled in parallel to the kiln. The preheated air is distributed by the manifold duct 3 behind the four fans serving for the internal circulation in the drying chamber 4. Four blowers 6 exhaust humid air from the kiln through the stack 7. A part of the air is recirculated to the collector from the drying

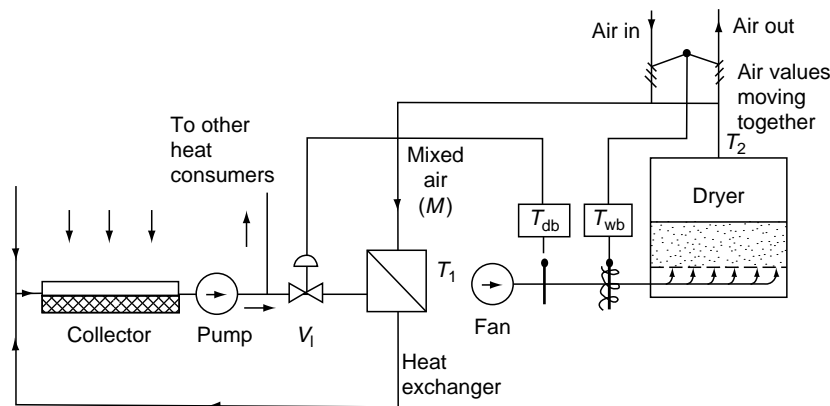


FIGURE 13.51 Control of a solar dryer with recirculation.

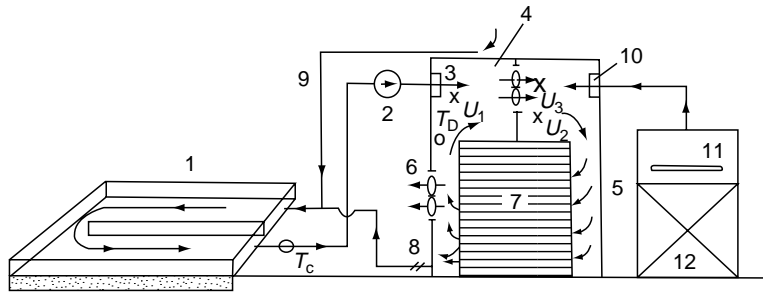


FIGURE 13.52 Simplified scheme of an automatically controlled lumber kiln dryer.

chamber through the dampered duct 8. Fresh air, slightly preheated by the black-painted roof surface, is led into the collector through duct 9. The wood residual burner 12 produces hot air for the drying space that is distributed by the manifold 10. A humidifier 11 is coupled to the burner for use when the humidity is below the minimum level.

13.7.4.4.2 Direction and Control of the Dryer

The operation of the dryer is directed and controlled as follows. Solar blowers 2 start when the temperature in the drying space T_D is lower than the collector outlet temperature T_c , dampers 8 are open. Blowers are activated by an on-off temperature control device. As a precondition of the operation of blowers the relative humidity U_1 should be lower than the set point value U_{1s} . Set point selection is manual. If $U_1 < U_{1s}$ the internal fans 4 are on. The operation of the exhaust blowers 6 is controlled by the signal of the relative humidity sensor U_2 situated behind the internal fans. The set point U_{2s} is high initially and should be reduced in the progress of drying. Exhaust blowers are activated if $U_2 > U_{2s}$. In the drying space the relative humidity should also be above a minimum level U_{3s} . When $U_3 < U_{3s}$ humidifier 11 will be activated by the control signal of the U_3 sensor. By applying water spray into the furnace chamber, humidification of the air led to manifold 10 will be effected. Operation of the burner is directed by U_3 or manually. The solar blowers 2, the internal fans, the exhaust vents, the humidifier 11, and the dampers 8 can also be controlled manually by using bypass switches. When solar blowers 2 are off, the internal fans 4 and the exhaust blowers 6 can be in operation if the state of the air in the dryer satisfies the requirements.

Drying time can be influenced by the operation time of the burner. For the sake of electric energy saving the number of internal fans in operation can be reduced, generally, in the final period of drying. In this period the humidifier 11 can be used for relief of drying stresses. The control of the drying process can be performed according to a schedule by applying a

timer that opens or closes the control relay at a determined time point. It can be bypassed manually. When the dryer is out of operation, dampers 8 are closed and the drying space can be isolated from the collectors (e.g., in the night).

13.7.4.5 Direction and Control of Solar Dryers with Water Storage

Solar dryers applying a water tank for heat storage have an indirect heat transfer system: the working medium of the collector is liquid (e.g., water) and the drying air should be preheated in a liquid-air heat exchanger. Two different constructions are discussed below.

13.7.4.5.1 Construction and Control of a Dryer with Water Storage

The scheme of a simple system is presented in [Figure 13.53](#). In this system three flow loops are applied. The first is that of the collector-tank loop in which the flow of the liquid working medium is maintained by pump P1. The second loop is that of the tank-heat exchanger with pump P2. The third one is the open loop of the air that is transported by the fan through the heat exchanger toward the drying space. Drying air is preheated by the heat exchanger using water from the heat storage tank.

Temperature required for drying T_D is controlled by control device CV using temperature sensor T_D and valve V. Fan is in operation when T_D is higher than the lowest temperature limit as set point value T_{Ds} . The fan is turned on by thermoswitch SD. The minimal temperature level of the water needed for ensuring T_D is T_s , which is the set point for the temperature sensor in the tank T_T . Operation of pump P2 is induced by the thermal switch SP2. Pump P1 and the collector are in operation if the liquid outlet temperature $T_L \geq T_T$. The operation of the pump P1 is induced by the control device CP1. The controlling signal is the temperature difference $T_L - T_T$.

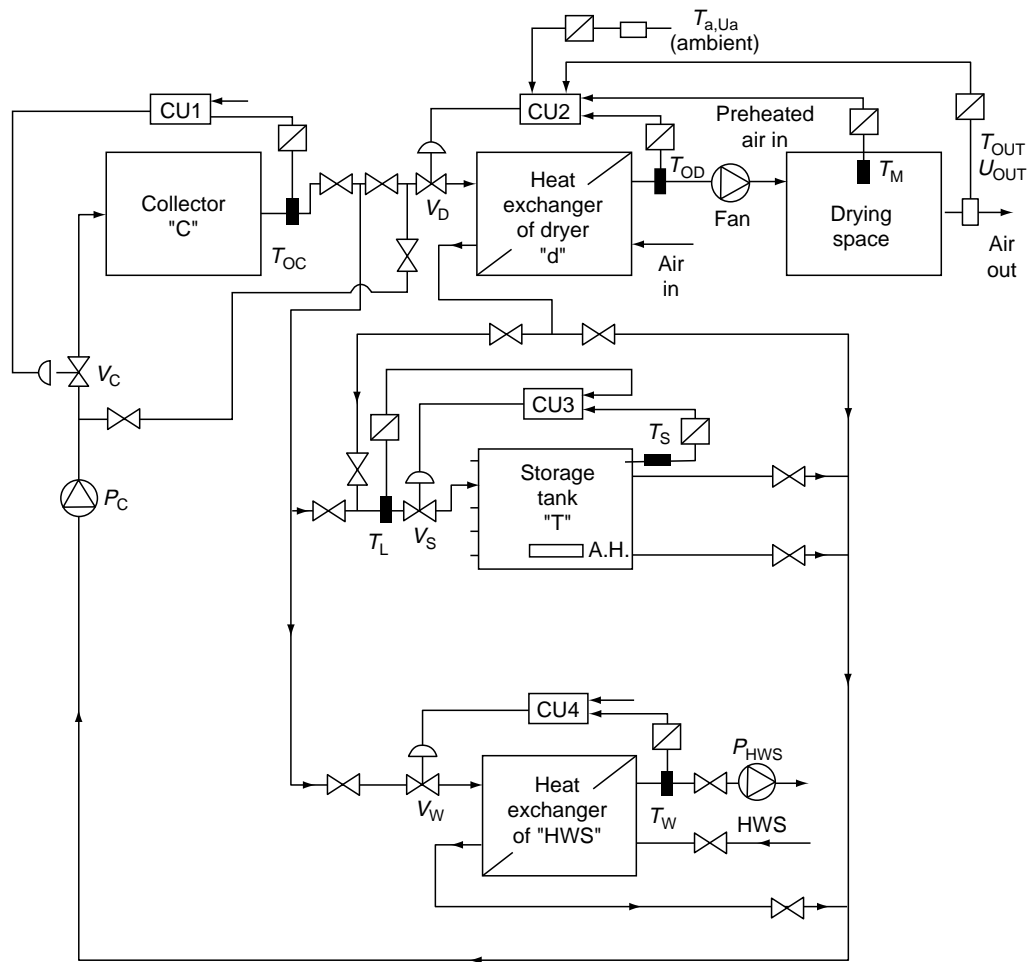


FIGURE 13.54 Scheme of the control system.

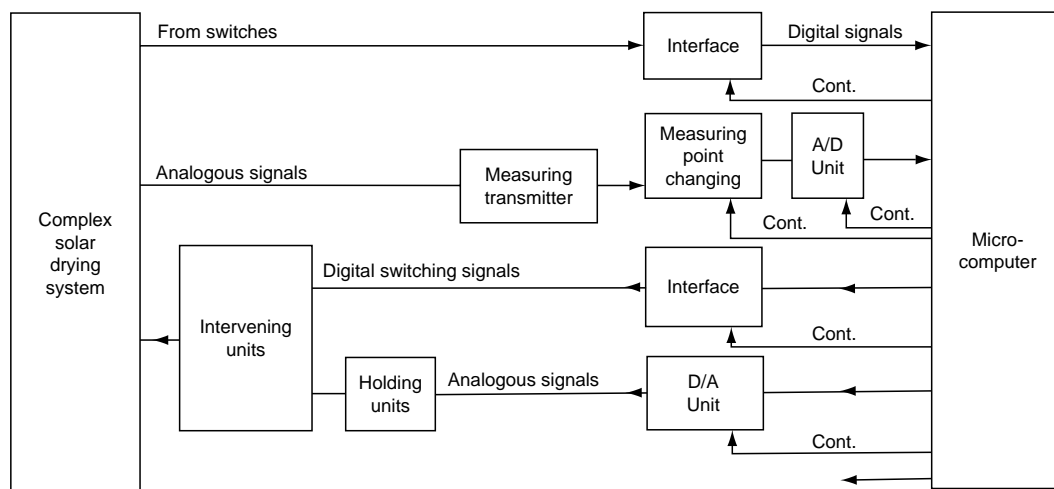


FIGURE 13.55 Block scheme of the microprocessor.

TABLE 13.4
Solar Drying of Different Materials

Material	Refs
Agricultural products	155, 166
Banana	18
Coffee	3
Crop	3, 129, 138, 153, 155, 156, 158, 159, 161, 163, 164, 168, 173, 205
Fruits	3, 6, 11, 41
Garlic	12
Grain crops	3, 13–15, 19, 31, 128, 135, 148, 155, 167
Hay, herbage, grass	3, 13, 16, 132, 133, 135, 136, 141, 142, 171
Jute	3
Peanuts	3, 17, 19
Raisins	19, 143, 170
Rice	146, 219
Sorghum	140
Soybeans	19
Timber	3, 8–10, 19, 20, 29, 124, 133, 134
Tobacco	19, 137, 144, 145, 169
Tomato	139
Vegetables	3, 193, 194, 198
Lumber	162, 171, 172

process is realized by computing parameters serving for the indication of the beginning and the time interval of the break. A microprocessor forms commands for the execution of the direction and control actions, indicates the main information about the actual state of the system, and calls attention to manual interventions when needed.

13.8 PROSPECTS FOR SOLAR DRYING

Further research and development of various components of solar drying systems continues to proceed internationally. Major current applications are confined to drying of agricultural and forest products. Table 13.4 summarizes selected references on solar drying of various materials.

Some significantly good payback times (1–7 years) have been achieved [14,17,133], mainly with simple and cheap dryers. The chances for the extensive use of high-performance systems may be improved by integrated construction and multipurpose operation. Using modern process control techniques, the efficiency of solar dryers can be increased. Due attention must be paid to system maintenance and training of the operating personnel.

In the design of solar dryers modern methods of modeling and simulation can play an important role in both the design and optimum operation. The renewable nature of solar energy is a definite asset in most parts of the world.

NOMENCLATURE

a	specific transfer surface by volume basis, m^{-1}
a	acceleration, m/s^2
A	area, m^2
b	biologic heat source, J/kg^1
c	specific heat capacity, $J/(kg/K)$
C	heat capacity, J/K
CU	control unit
\dot{C}	heat capacity flow rate, W/K
d_h	hydraulic diameter, m
e	effectiveness
e	error
E	irradiance, W/m^2
f	efficiency function
f	Fanning coefficient
g	acceleration of gravity, m/s^2
h	heat transfer coefficient, $W/(m^2 K)$
H	heat exchanger effectiveness
I	global irradiance, W/m^2
k	heat conductivity, $W/(m K)$
K	heat conduction, W/K
K	gain of controller
L	length, m
M	mass, kg
\dot{m}	mass flow rate, kg/s
N	mass flow rate of evaporation, kg/s
n	revolution, s^{-1}
p	pressure, Pa
q	radiant flux density, W/m^2

Q	radiant energy, J
r	heat of evaporation, J/kg
P	pump
R	resistance in a network
R	gas constant, J/(kg K)
S	switch
w	velocity, m/s
W	integral moisture content, kg/kg
W	water
z	height, m
t	transmittance
T	temperature, K
U	overall heat transfer coefficient, W/(m ² K)
U	relative humidity of the air, %
V	value
x	absolute water content of the air by dry basis, kg/kg
X	moisture content of the material, dry basis, kg/kg

GREEK SYMBOLS

α	absorbance
β	moisture transfer coefficient, m/s
ε	emittance
ϕ	heat flux, W
η	efficiency
η	dynamic viscosity, kg/(s m)
ζ	shape flow resistance coefficient
ξ	porosity
ρ	reflectance
ρ	mass density, kg/m ³
σ	Stephan–Boltzmann constant
σ	evaporation coefficient, kg/(m ² s)
λ	latent heat, J/kg
λ	wavelength, m
ν	frequency, s ⁻¹
τ	time, s

SUBSCRIPTS

A	auxiliary energy source
a, A	air
a	ambient
av	average
b	beam (direct)
b	bottom
b	break
c	collector
c	control
cr	critical
d, DIR	direct
db	dry bulb
D	drying
e	equilibrium

f, F	fluid
h	heat
$H-T$	horizontal–total
i	incident
i, j	nodal points in a network
in	inlet
k	component
l	lag
l, L	loss
L	liquid
lt	long term
m	number of nodes in a network
m, M	material
n	number of the discrete parts
n	normal
o	outside atmosphere
O	set point
out	outlet
p	at constant pressure
r	reflected
R	reset
s	space
s	sky
S	set point
t, TOT	total
T	temperature
th	threshold
u	useful
v	vapor
w	water
wb	wet bulb
pw	for humid air
(k)	time period
$'$	inlet
$''$	outlet

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14 Spouted Bed Drying

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14.1 INTRODUCTION

The applicability of the spouted bed technique [1–5] to drying of granular products that are too coarse to be readily fluidized (e.g., grains) was recognized in the early 1950s. Interest in this area received appreciable impetus two decades later as the energy-intensive drying processes were reexamined with renewed vigor. Spouted bed dryers (SBDs) display numerous advantages and some limitations over competing conventional dryers. Because of the short dwell time in the spout, SBDs can be used to dry heat-sensitive solids,

such as foods, pharmaceuticals, and plastics. With simple modification the so-called modified spouted beds can be designed to ensure good mixing, controlled residence time, minimum attrition, and other desirable features. Also, the operations of coating, granulation agglomeration, and cooling, among others, can be carried out by the same apparatus by varying the operating parameters. SBDs can be used for solids with constant as well as falling rate drying periods. Using inert solids as the bed material, SBDs have been used successfully to dry pastes, slurries, and heat-sensitive materials.

This chapter is devoted mainly to the generally less accessible results on spouted bed drying obtained at the Research Institute of Chemical and Process Engineering of the Hungarian Academy of Sciences (PE MÜKKI). Other results are readily available in literature.

14.2 EXPERIMENTAL DEVICES AND PROCEDURES

The classic or conventional spouted bed (CSB) is a cylindrical vessel with a conical bottom fitted with an inlet nozzle for the introduction of the spouting air (drying medium). This device suffers from limited capacity due to the maximum spoutable bed height and inability to scale up the apparatus beyond 1-m diameter. The introduction of a hollow, tall vertical tube (draft tube) some distance above the nozzle eliminates the former restriction by acting as a pneumatic conveyor.

Figure 14.1 displays a draft-tube SBD, which is a tremendous improvement over the classical SBD [6,7]. The tube may be impermeable, porous, or partly porous. It may be cylindrical or slightly tapered. The bed height can be increased several fold by inserting a suitable draft tube with solid walls. The solids in the annulus flow in a plug-flow fashion, guaranteeing a uniform residence time distribution (RTD) in the SBD. Experiments in two-dimensional beds showed that the RTD is more uniform in two-dimensional rather than circular cross-sectional SBDs. Table 14.1 and Table 14.2 show the geometries of the two SBDs studied.

The drying capacity of the laboratory-scale angular predryer for corn was about 80 kg/h of product, with a 10% decrease in moisture content. In the course of this calculation it was presumed that the particles on the average made two cycles within the bed.

The second part of the two-stage SBD is a conical-cylindrical device of traditional shape (Table 14.2). The drying capacity of this device was about 80 kg/h of dried corn with 150°C air at a flow rate of about 200 m³/h, which resulted in a moisture removal of about 5% in the falling rate period.

Efforts have been made to develop new solutions for the introduction of air, which will avoid the

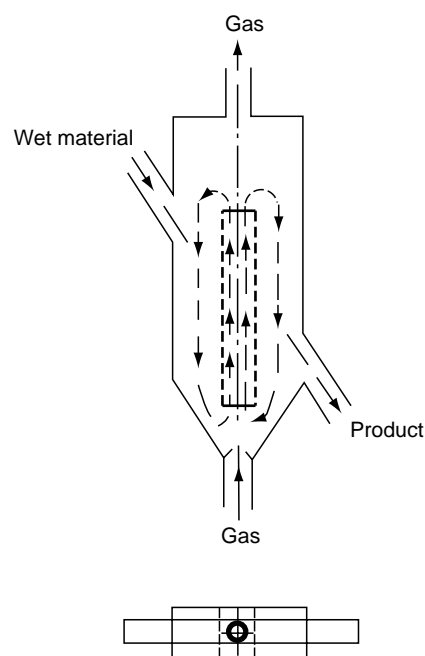


FIGURE 14.1 Longitudinal and cross-sectional schematics of the spouted bed predryer.

high-pressure loss caused by the central nozzle. The problem was solved by tangential air feeding with the use of horizontal slits (see Figure 14.2) [8] and by the so-called swirling rings (see Figure 14.3) [9]. The first type of equipment is referred to as a “vortex bed dryer”; the latter device ensures the maintenance of uniform particle circulation in the spouted bed, favoring the development of the spout channel. Figure 14.3 shows this device, a novel type of SBD equipped with an inner cylinder, a draft tube, and an attachment permeable to air. This type of dryer can operate with flow rates three to five times higher than those used in conventional SBDs and the start-up of the dryer also becomes simpler.

To select the optimal geometry of the draft tube it was necessary to carry out experiments in semicircular spouted beds for visualization of the flow patterns. It was then possible to measure the particle velocities in the annulus as well as the spout by introduction of marked (tagged) particles. High-speed photography was employed for this purpose [10].

TABLE 14.1
Dimensions of the Angular Spouted Bed Predryer

Longer Dimension (mm)	Shorter Dimension (mm)	Height (mm)	Nozzle Diameter (mm)	Cross Section of the Annulus (m ²)	Mean Sliding Velocity (m/s)
350	40	1000	20–40	0.0124	0.01

TABLE 14.2
Dimensions of the Cylindrical Spouted Bed Afterdryer

Diameter (mm)	Height (mm)	Nozzle Diameter (mm)
170	1500	17–30

Since the velocity of particle recirculation depends mainly on the velocity of the slow sliding in the annulus, the mean residence time of the particles is about $\bar{\tau} = H/w_a$. In order to investigate the possibility of controlling the circulation of particles, various sets of measurements were carried out with glass beads, with ground-activated carbon, and with plastics as model particles. It was found that the sliding velocity of the particles varies according to the following empirical correlations. It is affected by the velocity of the entering air, the bed height, and the nozzle diameter [10]:

$$w'_a = w''_a \left(\frac{v'}{v''} \right)^{3/2} \quad (14.1)$$

$$w_{aH} = w_{aH_{\max}} \left(\frac{H}{H_{\max}} \right)^{1/3} \quad (14.2)$$

and

$$w'_a = w''_a \left(\frac{D''_i}{D'_i} \right)^{2/3} \quad (14.3)$$

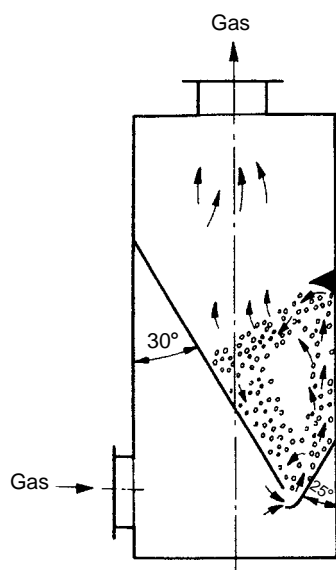


FIGURE 14.2 Dryer with a slit for gas introduction. (From Sulg, E.O., Mitere, D.T., Rashkovskaya, N.B., and Romanbore, P.G. *J. Appl. Chem (USSR)* 43: 2204 (1970). With permission.)

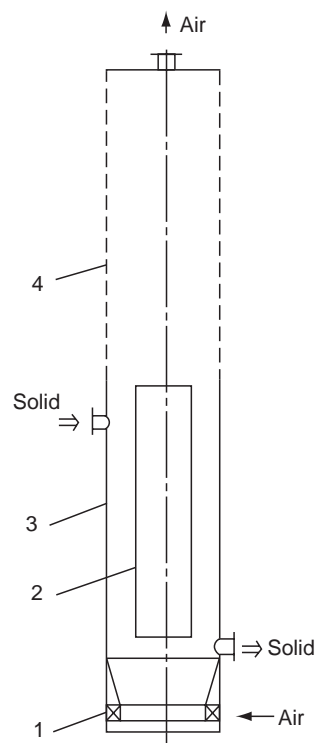


FIGURE 14.3 Spouted bed dryer with a draft tube and an air-permeable attachment: 1, tangential air feed; 2, draft tube; 3, dryer body; 4, air-permeable attachment.

However, the operation of the dryer depends not only on the mean residence time but also on the RTD of the particles. Therefore, complementary measurements were carried out to study this variable.

Polyvinyl chloride (PVC) granules of various colors were used in these experiments. To a white particle layer some black PVC particles were fed instantaneously onto the upper level of the annulus at the right side. During continuous feeding of the white PVC granules, the composition of the mixture (the proportion of white and black granulates) leaving the device could be continuously measured through a calibrated conveying belt and a movie camera.

The proportion of the particles of different color was recorded at intervals of 3 and 5 s. For example, the distribution of residence times is presented in Figure 14.4, in which the broken line denotes measured values and the solid line corresponds to the mean values of the distribution. The local peak values recurring periodically in the density curve may be explained by the fact that the marked particles are

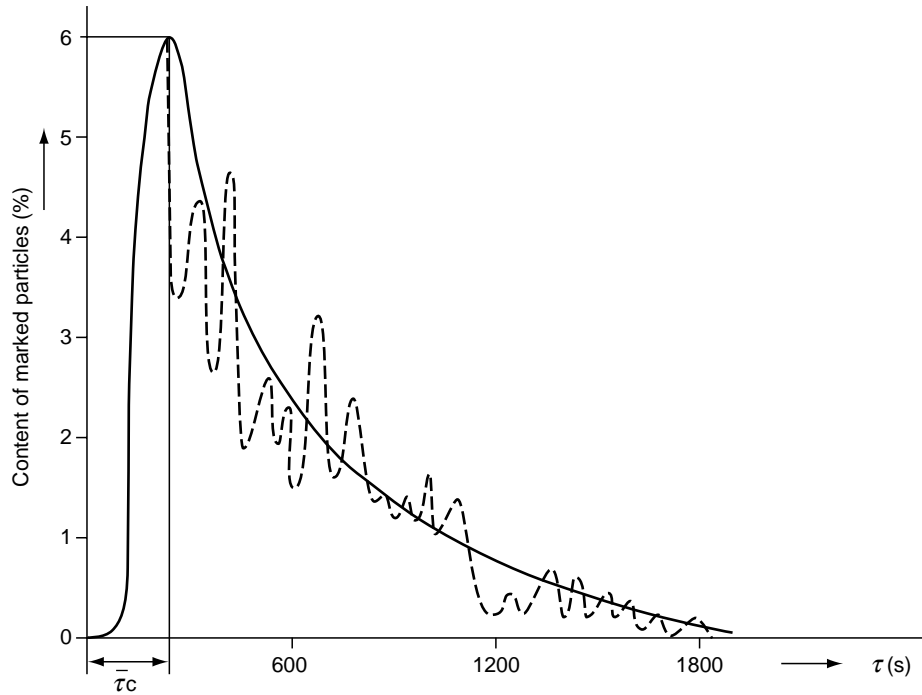


FIGURE 14.4 A typical distribution of the residence time of particles. $\bar{\tau}_c = 2\tau_1 + \tau_2$.

entering the outlet in different periods of the recirculating movement of particles, and by the fact that they are occasionally swallowed up from the annulus into the spout channel and thus fractions of the full recirculation period may occur.

From our experiments it appears that correlations that assume perfect mixing, as found in the literature [11,12], can be applied for the calculation of the distribution of residence times only under conditions that ensure intense circulation of particles in the bed. Even in these cases, the agreement is better at short dimensionless times ($\theta = \tau/\bar{\tau}$) corresponding to the initial section of the curve. According to the above-mentioned case, some modification of the physical model of the distribution of residence times appears desirable. Better agreement between the measured and calculated data was obtained with the plug-flow model, simultaneously taking into account the internal recirculation [13,14].

Using the symbols of Figure 14.5, the RTD function can be expressed by

$$\rho(\tau) = \left(\frac{1}{\tau_{in}} \right) \left(1 - \xi_a - \xi_b + \xi_a \xi_b \sum_{n=0}^{\infty} (\xi_a + \xi_b - \xi_a \xi_b)^n \right. \\ \left. \times \theta[\tau - 2\tau_1 - \tau_2 - n(\tau_1 + \tau_2)] \right. \\ \left. - \theta[\tau - \tau_{in} - 2\tau_1 - 2\tau_2 - \tau_3 - n(\tau_1 + \tau_2)] \right) \quad (14.4)$$

In this equation,

$$\tau_1 = \frac{S_a H}{2Q_1}, \quad \tau_2 = \frac{S_s H}{Q_2}, \quad \tau_3 = \frac{S_a H}{2Q_3}$$

and the recirculation fractions are $\xi_a = Q'/Q_2$ and $\xi_b = Q''/Q_3$.

From Equation 14.4 it can be estimated that these impulses appear at attenuated amplitudes with periods of $\tau_1 + \tau_2$. Because of the irregular pattern of recirculation of particles, the peak value does not appear as a sharp peak [6].

A spouted bed dryer with tangential air inlet equipped with an inner conveyor screw has also been developed [15] (see Figure 14.6). According to the developed and patented solution air is injected into the bed through specially designed “whirling” rings. Along the vertical axis of the device is a houseless open conveyor screw capable of ensuring, independently of the airflow rate, the typical spouted circulating motion even with materials of small particle size ($D_p/d_p > 500$) for which CSBs are not suited. The diameter of the screw is nearly equal to the diameter of the gas channel (spout) around which a similar dense sliding layer (annulus) is formed and a circulation motion (similar to conventional beds) can be visible.

Pressure drop is lower by 25–30% in comparison with air injection through a nozzle of CSB. Another advantage is that gas velocity can be regulated in a

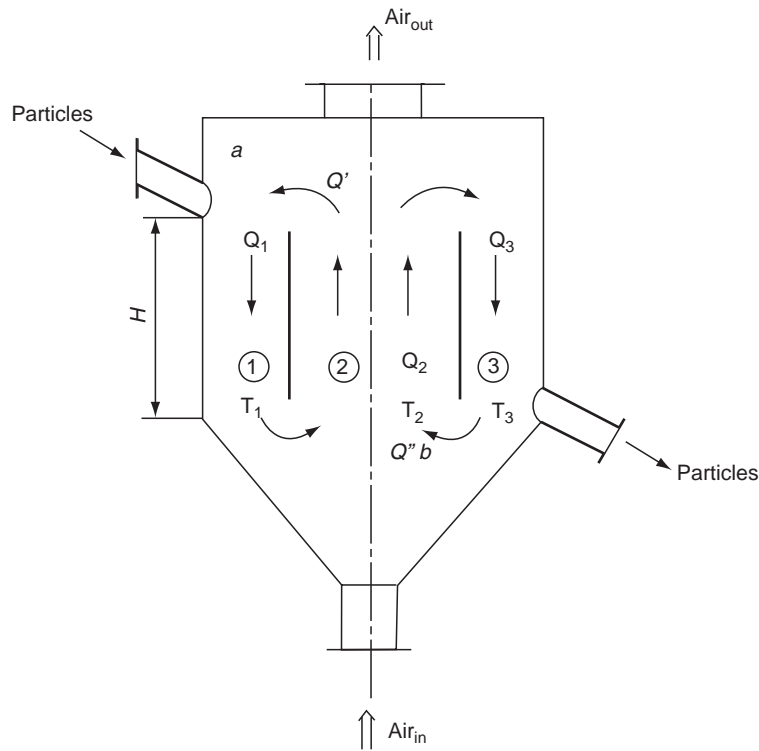


FIGURE 14.5 Proposed model of particle circulation.

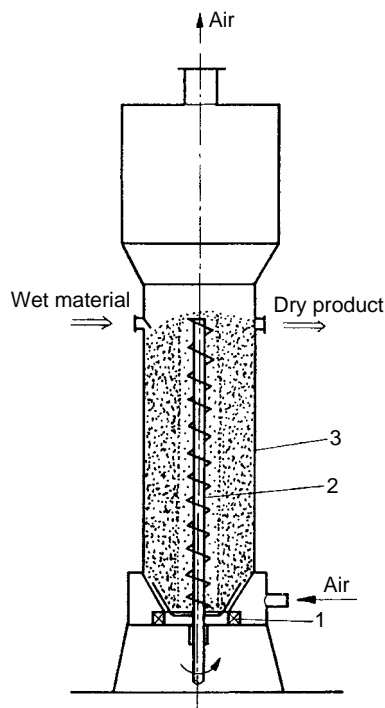


FIGURE 14.6 Spouted bed dryer with a tangential air inlet and a central conveyor screw: 1, tangential air inlet; 2, conveyor screw; 3, dryer body.

wide range due to the possibility of choice of proper size and number of slots. Due to mechanical particle circulation fan energy consumption can be reduced by 15–20% [16]. Moreover, using the inner screw, air volume rate can be chosen only from the point of view of drying requirements, resulting in optimum drying conditions. This mechanically spouted bed (MSB) construction offers further advantages in solving scale-up problems. Bed volume and diameter to height ratio can be selected quite freely, materials of wide particle size ranges can be circulated, and particle circulation time and rate can be controlled within wide limits [17].

Scale-up data of conventional SBDs are well known, but the relations cannot be applied to MSB dryers with inert packing due to different hydrodynamical and drying characteristics. Therefore, the drying mechanism itself, effects of various process and operational parameters, and the relevant relationship had to be investigated in more detail. Important dimensions as well as geometric, physical, and hydrodynamical characteristics of equipment and of the particles forming the spouted bed are summarized in Table 14.3 [18,31,32].

14.3 DRYING RESULTS

The inventors of the original patent concerning the spouted bed method and the construction of the

TABLE 14.3
Characteristics of Laboratory-, Pilot- and Industrial-Scale Mechanically Spouted Bed Dryers

	Laboratory	Pilot	Industrial
D_c (m)	0.138	0.380	1.0
Number of slots	2	6	8
h_i (m)	2×10^{-3}	4×10^{-3}	8×10^{-3}
A_i (m ²)	1.28×10^{-3}	2.1×10^{-3}	0.16
D_{sc} (m)	0.04	0.12	0.35
d_{sc} (m)	0.016	0.05	0.135
s (m)	0.028	0.09	0.180
s/D_{sc}	0.70	0.72	0.5
D_{sc}/D_c	0.29	0.31	0.35
	Ceramic spheres		Hostaform
Inert particles			
d_p (m)	6.6×10^{-3}	7.4×10^{-3}	12×10^{-3}
ρ_p (kg/m ³)	3640	3520	1340
ε	0.36	0.36	0.40
a (m ² /kg)	0.30	0.23	0.22
Re_{mf}	1100	1285	1260
u_{mf} (m/s)	2.6	2.7	2.4

apparatus [1] proposed primarily to solve the problem of the drying of cereals, such as wheat and corn, in a continuous and efficient way. However, the demand for an up-to-date drying apparatus that emerged in the course of the manufacture of products in various industries resulted in widening the field of application of the spouted bed dryer. Operational experience and frequent emergence of novel problems of drying initiated an activity to develop newer devices. As a result of this activity, various modified SBDs proved to be suitable for energy-efficient drying in the agricultural and food industry, the chemical industry, and many other industries for drying of powdered materials ($d_p < 100 \mu\text{m}$) and of centrifuged materials with high moisture content, as well as coagulating and adhesive pulps, and suspensions.

In the following sections, we will discuss typical SBD applications. We shall also describe briefly the operating conditions and results of spouted bed drying in laboratory- or industrial-scale units. These results may form the basis of industrial dryer designs when coupled with appropriate engineering judgment.

14.3.1 DRYING OF AGRICULTURAL PRODUCTS

The optimum parameters for drying of granular agricultural products are determined not only by their chemical composition and physical properties but also by their potential use. In the case of corn and

oats, for example, as cattle feed, when the valuable nutrients (including proteins and vitamins) must be saved, the temperature of drying air may be higher (the seed temperature can be as high as 60–75°C) than is the case of seeds for sowing. In this latter case, in order to conserve the ability of germination, the temperature must not exceed 40–45°C. However, the actual level depends on the seed type and on the residence time of the particle. In case of some seeds, a maximum seed temperature of 30–35°C is required. The quality of the dried product is determined, furthermore, by the rate of drying, the temperature, and the flow rate of the drying agent, mainly in the initial period of the drying.

Cereal seeds and, generally, also most other seed varieties are colloidal materials with capillary pores. This means that the walls of their capillaries are elastic and change their shape on absorbing moisture. They are composed of a skin, an endocarp, a so-called aleuron layer between these two parts and the embryo. The hygroscopic properties of these parts are different. At equilibrium, the embryo has the highest moisture content, followed by aleuron cells and fiber cells. Accordingly, a greater part of moisture is present in the external layer of the seeds, from which it can be removed relatively easily. On drying the seed is crumbled and becomes hard. At a drying temperature below 50°C, the drying rate of the seed decreases at first abruptly, then much more slowly. In the temperature range between 46 and 50°C, the aleuron cells form an almost closed, elastic skin in which the diffusion of the moisture is very slow. At a drying temperature above 50°C, the structure of various parts of the seed changes. The thermally labile cell content is hardened, the cell walls are split, and the moisture moves quickly through these splits toward the surface of the seed.

In the case of seeds for sowing, that is, when the temperature of seeds must definitely be maintained below 45°C, the drying rate is lower. Consequently, it is practical to increase, instead of the temperature, the velocity of the drying agent, obviously only to the optimal limit. At this limiting value of air velocity the moisture migrating from the interior of seeds to the seed surface is removed immediately. Beyond this limit it is not reasonable to increase air velocity. The drying curves of corn and wheat are shown in Figure 14.7 and Figure 14.8, respectively, at various drying temperatures; the critical moisture contents are also indicated.

The drying rate is affected by the manner of drying and the flow rate and temperature of the drying agent. Two grass varieties and two herbs (scarlet clover and coleseed) were dried in a flow-through oven and in a laboratory-size SBD with a central

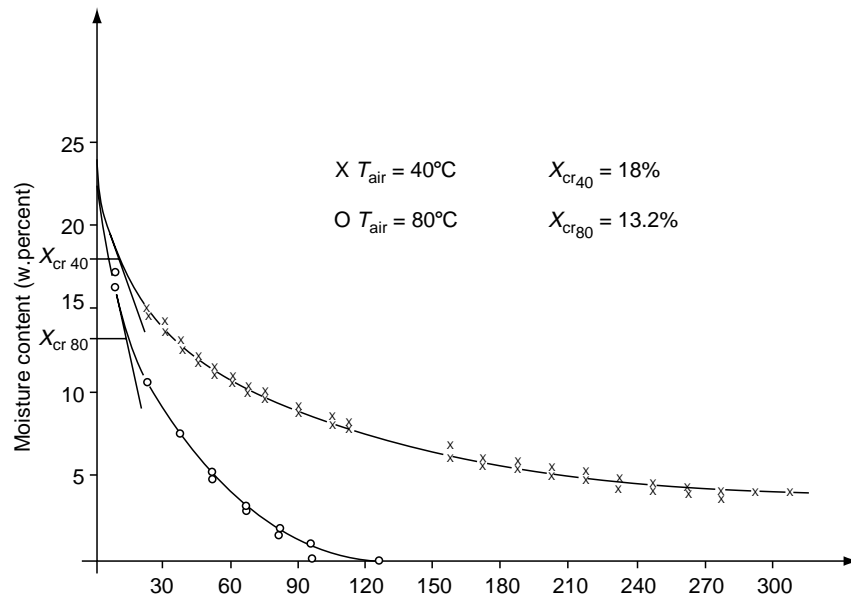


FIGURE 14.7 Drying curves for corn plotted against the drying temperature measured in a flow-through drying oven.

screw conveyor. Significant deviations are observed in the times required for attaining the final moisture content of 14% prescribed for storage (see Table 14.4).

For determination of the drying rates, airflow rates used in the SBD experiments varied from 0.6 to 1.0 m/s for various seeds. The typical drying rate curves also present some information concerning the structure of the material. The curve labeled “grass seed I” in Figure 14.9 is used as an illustration.

It can be seen in Figure 14.9 that, for oven drying, constant rate drying extends from a moisture content of 32.0 to 26.5% (by weight), characterizing the removal of moisture from the surface layers. The

section of falling drying rates points to a complex structure of the internal pore. However, it is quite striking that in dryings carried out in the spouted bed the shape of the rate curve is different. The protracted shape of the constant drying rate (i.e., from 32 to 13.5% by weight) allows the conclusion that inside the seed there are open pore-spaces, between the fibers—from which the moisture can be removed more quickly by increasing the flow rate of the drying agent. The critical moisture contents of the seed varieties listed in Table 14.4 are given in Table 14.5.

The equilibrium moisture contents are important in the design of drying equipment. For example,

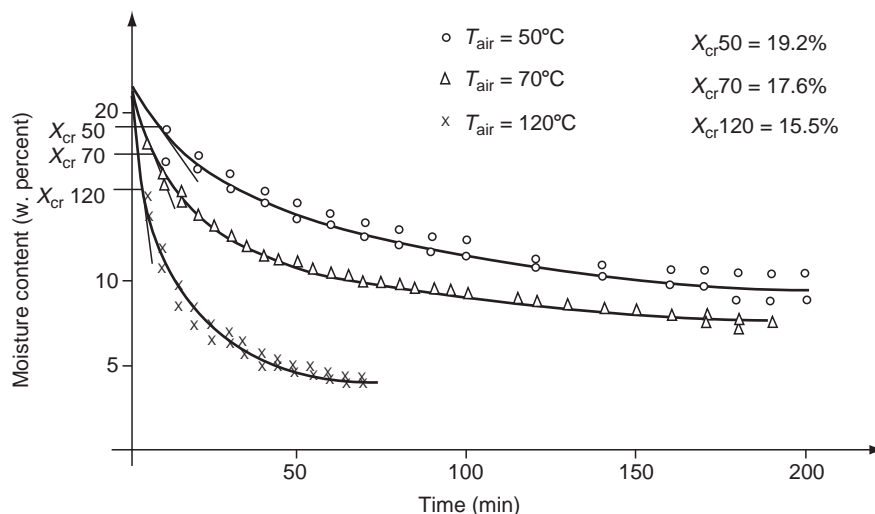


FIGURE 14.8 Drying curves of wheat plotted against the drying temperature measured in a flow-through drying oven.

TABLE 14.4
Drying Times of Seeds for Sowing to a Final Moisture Content of 14%

Seed Variety	Drying Time, $t_{14\%}$ (min)		
	Drying Oven	Drying by Airflow	Spouted Bed Dryer with Inner Screw
Grass seed I (319.0 kg/m ³)	132.0	61.2	27.0
Grass seed II (195.2 kg/m ³)	147.0	81.0	39.0
Scarlet clover (604.4 kg/m ³)	132.0	55.5	33.0

Figure 14.10 presents the adsorption isotherm of grass seed I at 20°C. The shape of this isotherm corresponds to the rare type III, according to the Brunauer–Emmett–Teller (BET) classification. It appears from this isotherm that, although it is also possible to dry grass seeds to a moisture content required for storage by atmospheric air of 70–80% humidity, time for drying is very high. About 24 h are needed to attain equilibrium at moisture content of about 14% by weight. In a spouted bed dryer using hot air at 45°C, the drying period is reduced to tenfold.

14.3.1.1 Drying of Wheat in Spouted Bed

The main characteristics of wheat used in the drying experiments are given in Table 14.6. It must be noted that the desired moisture content depends on the conditions of storage. At a moisture content of 14% by weight, wheat can be stored for a long time in bulk in large quantities, at 15% in bags for about 1 y, and at 16–18% in bags for some weeks, whereas at 19% by weight at most for only a few days.

For drying wheat to be used as seed for sowing, for example, the removal of the superficial moisture

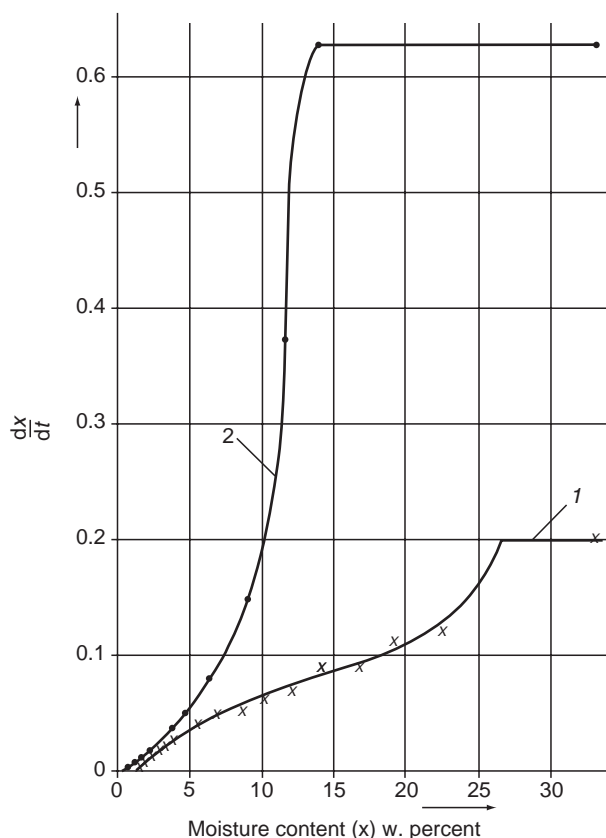


FIGURE 14.9 Drying rate curve of grass seed I on the basis of drying experiments carried out in a drying oven (curve 1) and in a spouted bed dryer (curve 2).

can be attained by air at 50–70°C (see the values of critical moisture content given in Figure 14.8); by parameters to achieve a particle residence time of about 25 min, the moisture content is reduced to 19 and 17.5% m/m with air at 50 and 70°C, respectively. Such highly efficient predrying may be needed to prevent damage of seeds before shipping to warehouses or processing plants. However, wheat to be stored over long periods in bulk must be at the prescribed moisture content of 15% m/m.

TABLE 14.5
Critical Moisture Contents of Seeds for Sowing Obtained from Drying Rate Curves

Seed Variety	Critical Moisture Content X_{cr} (% by weight)					
	X_{cr1}		X_{cr2}		X_{cr3}	
	Drying Oven	Spouted Bed	Drying Oven	Spouted Bed	Drying Oven	Spouted Bed
Grass seed I	26.5	13.6	16.0	11.5	—	—
Grass seed II	27.0	16.2	20.0	9.6	8.0	—
Scarlet clover	26.0	17.0	14.0	11.0	—	—
Coleseed	23.5	17.0	17.0	11.0	5.4	—

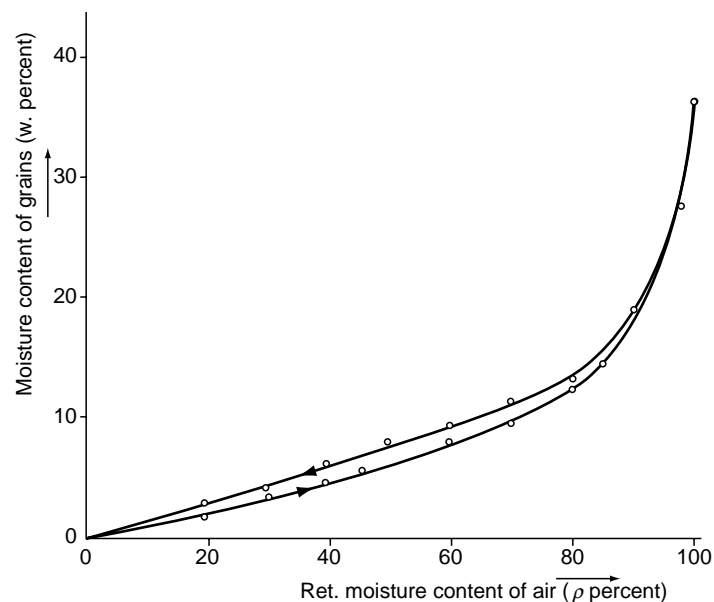


FIGURE 14.10 Adsorption isotherm of grass seed at 20°C.

It is practical to dry wheat for processing in mills in one step in a spouted bed dryer by air at 100–140°C without any damage to its composition. According to Figure 14.8, the value of x_{crit} is 15.5% by weight, which exceeds only marginally the desired final moisture content of 14% by weight. Thus, much of the drying may be carried out essentially at a constant drying rate.

The wheat drying studies were carried out in a spouted bed predryer of rectangular cross section equipped with a draft tube, and also in a cylindrical afterdryer (Table 14.1 and Table 14.2). The experimental conditions and the data of the batch- and continuous-drying experiments are presented in Table 14.7 through Table 14.10.

It appears from the data of Table 14.7 that wheat can be dried without any damage to the embryo to 17% m/m of moisture content in a spouted bed within 20 min by air at 100–102°C. In the afterdryer, the

moisture content of wheat was decreased from 25.5 to 11.7% m/m (see Table 14.8) at a mean temperature of 105°C of the drying air within 35 min without any damage of the embryo.

14.3.1.2 Drying of Corn in a Spouted Bed

A number of prototype SBDs have been used to dry shelled corn. One pilot-plant dryer has a designed capacity of 100-kg water/h, based on 10% moisture removal. This dryer is suitable for drying at two levels and for carrying out cooling after the drying. Accordingly, the equipment consists of a unit of smaller capacity (300 × 400 mm) and is equipped with a gas inlet nozzle and a draft tube, and of a spouted bed dryer of a cross section of 0.32 m² (800 × 400 mm) developed with two gas inlet nozzles [19]. The main

TABLE 14.6
Main Characteristics of Wheat

Mean particle size (\bar{d}_p)	6.5 mm (longer diameter), 3.5 mm (shorter diameter)
Density	798 kg/m ³
Initial moisture content	22–25.5% m/m
Moisture content of the product	11.5–19% m/m
Minimum fluidizing gas velocity	1.22 m/s
Operating gas velocity	3.19 m/s
Terminal gas velocity	10.2 m/s

TABLE 14.7
Drying Experiments for Wheat in a Spouted
Bed Predryer

Time (min)	Moisture Content of Wheat (% m/m)	Air Temperature		Embryo Number (%)
		T_{in} (°C)	T_{out} (°C)	
0	25.5	102	98	93
5	23.0	101	91	94
9	22.5	101	75	92
14	21.4	102	76	94
18	19.1	100	77	94
20	17.0	102	78	94

TABLE 14.8
Drying Experiments for Wheat in a Cylindrical Spouted Bed Afterdryer^a

Time (min)	Moisture Content of Wheat (% m/m)	Air Temperature		Grain Temperature (°C)	Quality ^b (%)
		T_{in} (°C)	T_{out} (°C)		
0	25.5	105	99.5	20.4	93
15	20.7	105	73.0	40.1	94
25	15.9	103	80.0	58.5	96
30	14.0	104	81.0	70.2	94
35	11.7	105	88.0	72.8	96

^aWeight of charge, 8 kg of moist wheat; volume of drying air, 120 m³/h (at 20°C).

^bPercentage of viable seeds.

characteristics of the shelled corn investigated in these drying experiments are given in Table 14.11, whereas the results of the laboratory and pilot-plant experiments are listed in Table 14.12 through Table 14.14.

The bed height was 1200 mm in order to utilize more completely the drying medium. The total pressure drop across the dryer was 1260 mm of water column (12.4 kPa); that is, drying that may be considered very good from the aspect of drying efficiency and of specific energy consumption represented in fact a significant energy requirement for aeration. It must be noted further that the products of the drying experiments 1 and 2 were also satisfactory from the aspect of the capability of germination; the bed

temperature did not exceed 80°C even at the highest temperature used.

The aims of the corn-drying experiments carried out in a large laboratory-size SBD (see Figure 14.6) were to decrease the pumping power requirement and to study the effects of various modifications on SBD performance. Table 14.14 summarizes the relevant results. This novel SBD has improved drying efficiency and favorable specific energy consumption; the bed pressure loss is reduced significantly from the earlier pressure drop of 1000–1400 mm of water column (9.8–13.7 kPa) to as low as 300–400 mm of water column (2.9–3.9 kPa).

TABLE 14.9
Continuous Wheat Drying in a Spouted Bed Predryer

<i>Experimental conditions</i>	
Initial charge	15 kg wheat of a moisture content of 17% m/m
Feed rate	15 kg/h wheat of a moisture content of 25% m/m
Mean residence time of grains	20 min
Air volume rate	50 m ³ /h at 20°C
Temperature of drying air	105°C
<i>Drying results</i>	
Mean temperature of air leaving the dryer	68°C
Mean moisture content of the dried product	16.9% m/m
Amount of removed moisture	1.53 kg/h
Water evaporating capacity referred to the dryer cross section	110 kg water/m ² h
Specific energy consumption	3360 kJ/kg water

TABLE 14.10
Continuous Drying of Wheat in a Spouted Bed Afterdryer

<i>Experimental conditions</i>	
Initial charge	7 kg wheat of a moisture content of 14%
Feed rate	m/m 40 kg/h wheat of a moisture content of 17%
Mean residence time of grains	m/m
Air volume rate	10 min
Temperature of drying air	120 m ³ /h at 20°C
<i>Drying results</i>	
Mean temperature of air leaving the dryer	100°C
Mean moisture content of the dried product	72.5°C
Amount of removed moisture	13.8% m/m 1.48 kg/h
Water evaporating capacity referred to the dryer cross section	74 kg water/m ² h
Specific energy consumption	7683 kJ/kg water

TABLE 14.11
Main Characteristics of Shelled Corn

Mean grain size (\bar{d}_p)	10.5 (longer distance), 7.5 mm (shorter diameter)
Bulk density	677.6 kg/m ³
Initial moisture content	22–30% m/m
Desired final moisture content	14% m/m
Minimum fluidizing gas velocity	1.58 m/s
Terminal gas velocity	14.6 m/s

14.3.2 DRYING OF OTHER FOOD PRODUCTS

In this section, the solution of some special drying tasks will be presented, such as the drying of preg-round red paprika, of various cut, chopped vegetables containing significant moisture (of about 50% by weight) (e.g., cubed carrots), and of various pulpy materials (e.g., potato pulp).

14.3.2.1 Drying of Washed and Centrifuged Tomato Seeds

Drying of agglomerated granular materials of high (45–50% m/m) moisture content causes difficulties. A further problem emerges in the treatment of the washed and centrifuged seeds with a moisture content of about 50% m/m, when a large amount of water must be removed at low temperatures in order to avoid damage at higher temperatures.

The drying of washed and centrifuged tomato seeds for sowing presents a problem of this type. On the basis of drying experiments performed in the laboratory-scale dryer, an industrial-scale SBD (with a capacity of 25 kg water/h) has been designed, manufactured, and put into operation [20].

TABLE 14.12
Drying of Shelled Corn in a Continuous Operation in a Laboratory-Scale Spouted Bed

<i>Experimental conditions</i>	
Charge	5 kg of shelled corn of a moisture content of 14% m/m
Feed rate	18 kg/h of corn of a moisture content of 23.7% m/m
Mean residence time of grains	16.6 min
Air volume rate	70 m ³ /h at 20°C
Temperature of drying air	90°C
<i>Drying results</i>	
Mean temperature of air leaving the dryer	47°C
Mean moisture content of the dried product	14.2% m/m
Amount of removed moisture	2.0 kg/h
Evaporational rate referred to the cross section of the dryer	143 kg water/m ² h
Specific energy consumption	3290 kJ/kg water

Tomato seeds for sowing are one of the end products of the tomato-processing technology, containing significant amounts of fibrous pulp and skin fragments. These are removed by washing combined with repeated sedimentation. The seed for sowing obtained in this way is adjusted by centrifugation to a moisture content of 45–50% m/m. Tomato seeds for sowing treated in this way are still very moist to the touch and in lump form. Uniform drying of this material can be carried out only by ensuring a mobile state of the particles. A further task is abrasion of the dry seeds, that is, the removal of the undesirable

TABLE 14.13
Drying of Shelled Corn in a Continuous Operation in a Pilot-Plant-Scale Spouted Bed Dryer of 0.16 m² Cross Section^a

Drying Results	Experiments			
	1	2	3	4
Temperature of the inlet air (°C)	150.0	120.5	178.5	190.0
Temperature of the outlet air (°C)	68.0	46.6	67.5	76.3
Mean residence time of grain (min)	20.0	14.6	11.1	10.0
Mean moisture of product (% m/m)	14.8	17.2	16.8	16.0
Amount of removed moisture (kg/h)	65.6	25.4	42.3	48.6
Capacity referred to the cross section of the dryer (kg water/m ² h)	396.0	132.3	220.3	252.1
Specific energy consumption (kJ/kg water)	2884	3868	3811	3485

^aExperimental conditions: Charge, 120 kg shelled corn of a moisture content of 14% m/m; feed rate, 360–700 kg/h corn of a moisture content of 29.7% m/m; mean residence time of grains, 10–20 min; amount of drying air, 1100 m³/h at 20°C; temperature of drying air, 120.5–190°C.

TABLE 14.14
Drying of Shelled Corn in a Continuous Spouted Bed
Dryer Equipped with a Draft Tube and an Inner
Conveying Screw

Experimental conditions (diameter of the dryer, 138 mm)

Drying airflow through a swirling ring	
Feeding of the wet material through the feeding tube moving circularly above the bed surface, distributing the material uniformly on the surface of the sliding bed (annulus)	
Draft tube	
Diameter	70 mm
Length	1000 mm
Conveyor screw	
Diameter	32 mm
Length	1200 mm
Speed of rotation of the screw	700 rpm
Weight of bed	13 kg of shelled corn with a moisture content of 14%
Mean residence time of particles	20 min
Air volume rate	120 m ³ /h at 20°C
Temperature of drying air	160°C

Drying results

Temperature of the outlet air	67.5°C
Mean moisture content of product	13.8% m/m
Amount of removed moisture	5.64 kg/h
Evaporational rate referred to the cross-sectional area of the dryer	376 kg water/m ² /h
Specific energy consumption	3535 kJ/kg water

residue on the seed surface to permit the use of a mechanical technique of sowing single seeds. This operation can be carried out practically in combination with the drying process in an SBD.

For this purpose, an SBD with a draft tube was chosen (enabling the application of large amounts of air), together with swirling-ring air entry (tangential air feeding). In order to prevent external damage to seeds, the removal of residue was carried out simultaneously with the drying procedure. The velocity of particles was increased to about three times the usual level. The particle velocity in the annulus was 0.05–0.07 m/s. The upper part of the equipment was made from linen permeable to air (see Figure 14.11).

Table 14.15 presents the calculated fundamental parameters of design based on laboratory measurements and also the main dimensions of a dryer of industrial scale. The operation of this type of SBD is as follows. Air for the dryer is supplied by a fan, heated by an air heater, and then led to the swirling ring along the periphery and fed tangentially. The cylindrical draft tube stretches about 20–30 mm to the linen part, which retains the dried residue. Figure 14.11 shows this dryer schematically. Table 14.16

gives other pertinent characteristics of this dryer. Quality tests indicated that the dried product displayed no damage to the seed embryo.

14.3.2.2 Drying of Paprika

Dried paprika pods are ground in a hammer mill. According to the practice followed so far, breaks of a moisture content of 10–16% were transported directly to a paprika mill, to be passed through several special rolls and runs in order to obtain grits suitable for use as a spice. If the initial moisture of paprika breaks is no more than 6–8%, the paprika mills can double their production capacity.

A spouted bed apparatus suitable for maintaining the recirculation of paprika breaks of a wide range of particle size with minimal clogging tendencies was developed for this purpose. Its efficiency was improved by using an air-fed annulus (swirling ring) of special design and by simultaneous application of a mixer and a draft tube. The spouted bed dryer is shown in Figure 14.3. The main parts of the device are the base, the drying column, and the separating column.

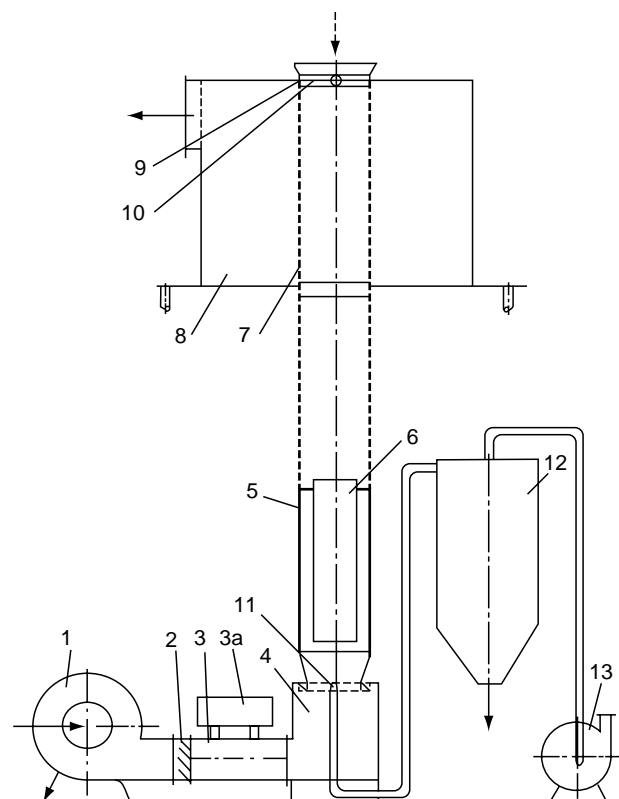


FIGURE 14.11 Spouted bed dryer of batch-type operation for drying of tomato seeds: 1, fan; 2, air feed control; 3, heater; 3a, regulator; 4, air chamber; 5, drying tank; 6, draft tube; 7, sieve; 8, dust chamber; 9, feed of wet material; 10, slap spindle; 11, discharge of the dried product; 12, cyclone; 13, ventilator.

TABLE 14.15
Main Dimensions of Spouted Bed Dryers
and Experimental Parameters

Dimensions of the Dryers (mm)	Laboratory Scale	Industrial Scale
Air dashpot		
Diameter	128	633
Length	650	900
Cylindrical part		
Diameter	140	538
Length	450	1830
Draft tube		
Diameter	75	317
Length	740	1220
Linen cloth permeable to air		
Diameter	200	538
Length	1500	1460
Bronze sieve cloth		
Diameter	—	538
Length	—	1500
Size of air inlet slits	10	5 rows of rings; each has 12 slits of diameter 8 mm (slit height: 8 mm)
Characteristic air velocity data		
Minimum fluidization velocity	0.94 m/s; (dry seed)	
Velocity	9.60 m/s	
Gas velocity in the draft tube	11.90 m/s	
Gas velocity in the annulus	1.30 m/s	

An electric heater, a stirrer, and the air inlet swirling ring are located in the base. The drying column includes the draft tube controlling the recirculation of particles. The device is completed by a wet-feed hopper, an over-

TABLE 14.16
Experimental Results Obtained with Tomato Seeds
for Sowing in a Dryer of Industrial Scale

Drying airflow rate	3500 m ³ /h at 20°C
Temperature of drying air	45–60°C
Temperature of outlet air	26–41°C
Pressure drop across the dryer	1.76 kPa
Drying period	
(total time of drying and abrasion)	85 min
Weight of the charge	200 kg of wet seeds
Initial moisture content	50–52% m/m
Evaporational rate	200 kg water/m ² h
Specific energy consumption	3156 kJ/kg water

flow chamber, a cyclone, and an air blower. The drying column has a cross section of 0.015 m².

In Table 14.17, note that the dryer capacity was in the range of 1000–1400 kg/m²h with a water evaporating capacity of 105–115 kg/m²h. Although the drying air temperature was rather high, the temperature of outlet air and the dried product was in the range of 39–62°C, and the product quality was good at a moisture level of 6–7%. The amount of air used here is several times that used in a conventional SBD. In test 4 of Table 14.17 carried out with a bed height of 1600 mm (approaching industrial dimensions), the specific heat consumption was found to be 4091 kJ/kg water. The pressure drop is 4.4–4.9 kPa, which is significantly lower than that for SBD using a nozzle.

14.3.3 DRYING OF PULPS AND PASTELIKE MATERIALS

For drying of particulate solids many well-proved processes are known, but during drying of pastelike materials and suspensions of high moisture content, due to their consistency, a lot of difficulties come up. In many cases, the local overheating, crusting also make impossible to provide a product of good quality. To obtain a uniform particle size often some disintegration or grinding are needed. In certain cases, spray drying is well applicable, but it demands very high costs of energy and investment.

Pastelike materials occur in many technological processes in the food processing, chemical industry, and so on. They are involved in the production of foodstuffs, organic intermediate products, pigments, pharmaceuticals, inorganic salts, and the like. Due to the variety of occurrence of these materials in engineering, the process of drying the pastelike materials can be considered an important stage of process technology. The drying process may greatly influence the quality of the product. The specific properties of the pastelike material affect the drying and general process requirements. Thus, in organizing the drying process it is necessary to take into account the properties of the materials that are dried. A knowledge of these properties and the laws governing the changes in parameters during drying is the basis for choosing the most practicable method of drying and the optimum conditions for carrying out the processes.

In view of the high moisture content of suspensions, economic drying can be performed only in dryers with intensive heat and mass transfer. The inert bed dryers provide good conditions for this purpose since the drying process is performed:

1. On a large and continuously renewed surface
2. In a thin layer formed on the surface of inert particles

TABLE 14.17
Drying of Paprika Breaks in a Spouted Bed

Conditions	Measurement			
	1	2	3	4
Initial moisture content (% m/m)	9.63	14.4	15.0	18.8
Bed height (mm)	490	490	490	1665
Mean residence time of particles (min)	12	13.5	11.5	48
Air feeding rate (m ³ /h)	56	61	78.8	63
Air temperature (°C)	120	147	144	122
Water evaporating efficiency (kg/h)	1.64	1.57	1.68	1.73
Specific rate of evaporation (kg water/m ² h)	109	104.6	112	115.3
Specific air utilization (kg air/kg water)	37.9	43.9	53.0	40.7
Specific moist-material efficiency (kg/m ² h)	1400	1116.8	1288.6	1001.6
Specific energy consumption (kJ/kg water)	4318	5644	6158	4091

3. With intensive contact between the wet material and the drying agent of high flow rate

For drying of materials of high moisture content, which cannot be directly fluidized or spouted, drying on inert particles can be advantageously applied. The principle of such drying is that the inert particles as an auxiliary phase form the fluidized or spouted bed. In such a case the suspension is fed into the moving or circulating bed of the inert particles, which provide a large surface for contacting. The wet solid distributed on the large surface of the inert particles forms a thin layer in which a very short drying process occurs. Due to the friction of the inert particles the dried fine coat wears off the surface, and then the fine product is carried out by the airstream.

Intensive, well-controlled heat and mass transfer can be carried out in MSB dryers in which the characteristic circulating motion of the particulate material is ensured by a vertical, houseless screw conveyor. This type of dryer and procedure can be advantageously used for continuous drying of materials of high moisture content (suspensions, sludge, etc.) by using inert charge.

In this case pastes, pulps, or suspensions of high moisture content are fed into the bed of inert particles circulated by the inner conveyor screw. In this way, an almost uniform, film-like coating is formed on the surface of the particles, the thickness of which in optimum case (short time drying) is 2 to 4 times higher ($\delta = 20\text{--}40\ \mu\text{m}$) than that of the primary particle size of the material to be dried. Since the inert particles provide a large contact surface, the heat and mass transfer processes of drying are short even at relatively low wet bulb temperature. Therefore, moisture diffusion resistance in solid can be considered negligible and drying takes place at

“quasiconstant” rate. This provides a very gentle drying, as well.

The spouted bed of circulating particles consists of three zones, which are separate and differ significantly in their flow characteristics, and are as follows:

1. The zone characterized by turbulent particle flow, enabling intensive gas–solid contact in the vicinity of the gas inlet
2. The zone of particles transported vertically upward by the screw-conveyor in cocurrent to the drying airflow
3. The dense annular part sliding downward in countercurrent to the airflow

There is no significant particle mixing between the zone of the sliding layer and that of the vertical particle transport.

From the point of view of drying the following subprocesses are of basic significance: formation of coating, drying of coating, and wearing of coating. It has been found that these partial processes of the inert bed drying occur in the following bed zones, that is

1. Formation of the coating in the annulus section of sliding down inert particles
2. Drying of the coating in the vicinity of the air inlet, in a bed height of a few (6 to 8) cm
3. Wearing mainly in the rotation area of the conveyor screw

Steady-state condition for drying can be achieved when the total operational time of the partial processes does not exceed the cycle time of the inert particles, that is the partial processes must take place during one circulation. It can be achieved by adequate coordination of the partial processes. In

order to clarify the mechanism of drying on inert particles in the MSB dryer it is important to know the time required by each of the partial processes as well as the residence time of the inert particles in the various zones of the bed [33].

The cycle time distribution (CTD) of the inert particles was measured as a function of the operational parameters of drying by use of ^{60}Co radioactive isotope as tracer. It was found that the inert particle circulation in the MSB dryer can be characterized by nearly plug flow except for the air inlet region at the bottom of the bed, in which the partial process drying takes place. However, the volume of this bed section is negligible in comparison with that of the whole bed. The time necessary for an entire particle cycle varies between 10 and 50 s, depending mainly on the bed volume and the rpm of the screw. From the particle velocities (residence times) measured in the important zones of the bed it could be calculated that a particle spends 70% of the cycle time in the sliding layer, 10% in the screw, and 15–20% in the intensive drying zone. This practically means that a particle can reside an average 2–10 s in the drying zone.

14.3.4 PRODUCTION OF POWDERLIKE MATERIAL FROM SUSPENSION BY DRYING ON INERT PARTICLES

14.3.4.1 Drying of Potato Pulp

Potato is a colloiddally dispersed material with a capillary porous structure. The mean composition of potato flour is presented in Table 14.18. Both raw and boiled potatoes are used in the production of potato flour. With potato flour prepared from boiled potatoes, the peeled washed potatoes are treated with sulfite and boiled for 30 min under pressure. Pulping may be carried out in a hammer mill before the material is dried on a cylindrical dryer at a temperature

of 60–70°C to a final moisture content of 7–9% by weight.

Drying of the potato pulp has been carried out in a large laboratory-size SBD fitted with a central screw conveyor, with an inert charge and tangential air feeding. A schematic of such an SBD, which can be used to dry pulp materials, pastes, and suspensions, is presented in Figure 14.12.

Potato pulp (moisture content, 75–80%) is fed through a suitable feeding system into the sliding part of the inert charge, which recirculates into the lower third part of the bed. (In the laboratory-scale SBD, the diameter of the feeding pipe for the pulp was 3 mm.) The size of this diameter plays an important role in the development of the product grain size. The uniform distribution of the pulp fed along the entire cross section is ensured by stirring elements attached to the inner screw axis at appropriate distances above and below the feeding site.

The recirculating motion of the inert particles, which are thinly coated with pulp, dries, grinds, and backmixes the dried powder with pulp to obtain good texture. The fine dry product is collected in a cyclone or in a bag filter. The grain size of the product can be controlled by the location of the discharge pipe and by the flow of air. The conditions and results of drying potato pulp in an SBD are presented in

TABLE 14.18
Composition of Potato Flour

Components	Content (% m/m)
Water	75.00
Total carbohydrate	21.00
Sugar	1.50
Starch	18.00
Protein	2.00
Fibrous material	1.00
Ash	1.00
Vitamin C (mg/100 g)	1.00
Vitamin B—thiamine (mg/100 g)	0.11

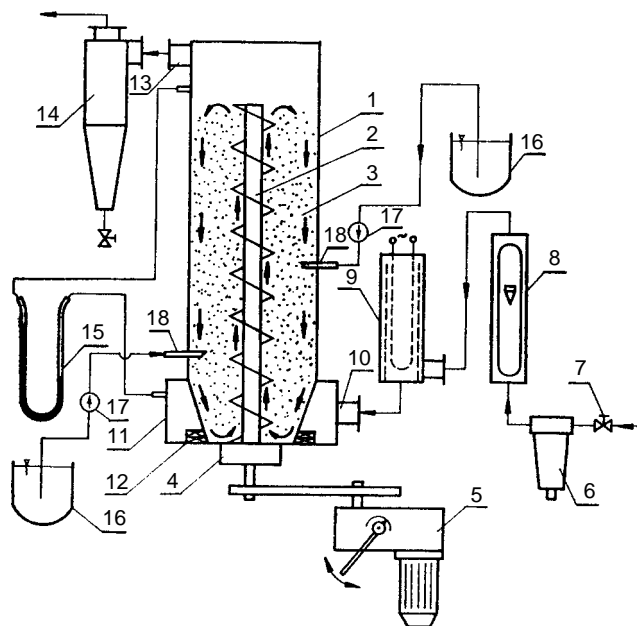


FIGURE 14.12 Flow scheme of mechanically spouted bed (MSB) dryer with inert particles for drying of materials of high moisture content: 1, dryer; 2, vertical conveyor screw; 3, inert packing; 4, bearing; 5, drive; 6, air filter; 7, valve; 8, rotameter; 9, heat exchanger; 10, air inlet; 11, air chamber; 12, slots for air inlet; 13, air outlet; 14, cyclone; 15, U-manometer; 16, suspension tank; 17, pump; 18, suspension feeding tube.

TABLE 14.19
Drying of Potato Pulp in a Mechanically Spouted Bed Dryer with an Inert Charge

<i>Experimental conditions</i>	
Wet solid	Potato pulp of moisture content 78% m/m
Feed rate	2.88 kg/h
Drying airflow rate	110 m ³ /h at 20°C
Mean temperature of the inlet air	114.5°C
Temperature of the outlet air	63°C
Speed of rotation of the screw	240 rpm
<i>Drying results</i>	
Removed moisture	2.87 kg/h
Mean temperature of the outlet air	42.5°C
Rate of evaporation referred to the cross section of the dryer	191 kg water/m ² h
Specific energy consumption	3025 kJ/kg water
Mean moisture content of the product	0.2% m/m

Table 14.19. In some areas, the SBDs for solutions and suspensions can be an economically interesting alternative to the spray dryer.

14.3.4.2 Drying of Brewery Yeast in Mechanically Spouted Bed Dryer

The task was to perform the drying of the brewery yeast suspension of 4.5 kg/kg db water content with density of 890 kg/m³.

The inactive brewery yeast suspension as the by-product of the beer production contains vitamin B and trace elements in relatively high concentration therefore the dried powder, after tableting can be circulated as roborant. It was very important to preserve its vitamin content whereas the moisture content of the dried product should be less than 5% m/m; moreover, the particle size of 90% of the product should be smaller than 0.4 mm and the size of remains should not exceed 1 mm.

Drying experiments were performed in a laboratory-scale MSB dryer with inert particles and the optimum process parameters were determined. The dimensions of the dryer and the operational parameters of drying are summarized in Table 14.20. From these data an industrial-scale dryer with a capacity of 100 kg water/h was designed (see Figure 14.13). The dimensions of the dryer and the operational parameters of drying are summarized in Table 14.20.

14.3.4.3 Drying of Some Other Products

Drying of pastelike materials of high thermal sensitivity such as animal blood and liquid vegetable

TABLE 14.20
Drying of Brewery Yeast in a Mechanically Spouted Bed Dryer with Inert Particles

<i>Technical parameters</i>	
Diameter	1.0 m
Height	1.8 m
Diameter of the screw	0.30 m
Diameter of particles	12 mm
Bed height	1.0 m
<i>Drying conditions</i>	
Moisture content	5.0 kg/kg db
Initial	0.05 kg/kg db
Final	5300 Nm ³ /h
<i>Airflow rate</i>	
Temperature	120°C
Inlet	70°C
Outlet	8–10 s
<i>Residence time of the wet material in the drying zone</i>	
Rate of evaporation	95–105 kg water/h
Specific drying rate	120–130 kg water/m ² h
Specific energy consumption	3000–3500 kJ/kg water

extracts in spouted bed of inert bodies is technically feasible [21]. Experimental investigations were carried out in a drying chamber consisting of a 60° conical base followed by a 0.14-m diameter cylindrical section. The chamber contained polypropylene beads as inert particles ($\rho = 820 \text{ kg/m}^3$ and $d = 3.9 \times 10^{-3} \text{ m}$). It has been observed during animal blood drying that when the air outlet temperature was kept at 74°C a dry product with soluble protein content of about 85% of its original amount can be obtained. At this temperature the process thermal efficiency is about 55%.

The dried vegetable products have shown the same organoleptic properties and moisture content as commercial vegetable powders for pharmaceutical use. Therefore, their active compounds do not lose their desirable characteristics after drying.

The product moisture can be adopted as a quality control parameter. It is recommended to keep it around 5% for dried vegetable extracts and around 6% for dried animal blood. Drying of agricultural products and by-products has received a great deal of attention in Brazil in recent years. The spouted bed with inert particles has also been investigated as a paste dryer for bovine blood and tomato paste, as well as for solutions of various pharmaceutical products. Another by-product that has been underutilized is the yeast produced in large quantities in the alcohol industry.

Since yeast is a living organism, great care must be taken in its drying to ensure a viable product for use



FIGURE 14.13 Industrial-scale MSB dryer with inert particles (photo).

in the baking industry. Owing to this heat sensitivity, the SBD with inert particles, in which the particle temperature stays well below the gas temperature, is considered to be a suitable dryer [22]. The SBD consisted of a conical base with 60° angle, followed by a cylindrical section with diameter of 500 mm. The diameter of the gas inlet was 50 mm. Drying was carried out on inert glass beads ($d_p = 1.8$ mm and $\rho = 2.5$ g/cm³). The inlet air temperature values were 100, 120, and 140°C , whereas the outlet temperature values were 60, 79, and 80°C . The airflow rate was kept at twice the minimum spouting value. It was found that the viability of the product was in the 50–70% range, and that a feed concentration of 10% was satisfactory.

An SBD for PVC tested in the Technological Institute in St. Petersburg, Russia has the following dimensions: $D = 400$ mm, $D_i = 70$ mm, and $\alpha = 40^\circ$. The cross section of the dryer was 1 m^2 , with a capacity of 400 kg/h of moisture and 1600 kg/h of dry material. Additional data for this operation are given in Table 14.21 [23]. Some of the examples described below relate to products of high moisture content, ranging from pasty or pulpy products to granular, crystalline substances.

TABLE 14.21
Drying of Polyvinyl Chloride in a Spouted Bed Dryer

Initial moisture content	60% m/m
<i>Temperature of drying air</i>	
Inlet	120°C
Outlet	60°C
Specific energy consumption	9000 kJ/kg
Specific rate of evaporation	$6.4\text{ kg/m}^2\text{h}$
Final moisture content	0.5% m/m

14.3.4.4 Drying of Calcium Carbonate of High Purity and Fine Grain Size

The moisture content of limestone produced in the desired grain size of $2\text{--}3\text{ }\mu\text{m}$ and in desired crystalline form, upon filtration or centrifugation, is 45–50% m/m. This pulpy product can be dried in an MSB dryer with inert charge. Here, the conveying screw was made up of Teflon because metallic moving elements would have caused abrasion and contamination of the product.

The diameter of the drying chamber was $D = 140$ mm; its height was 450 mm. The upper part had a disengagement zone with a threefold increase of area. The conveying screw, diameter of 75 mm and 650 mm long, extended into the upper-enlarged duct. The surface of the bed, consisting of inert grains, must be lower by about one screw profile than the end of the conveying screw. This way larger bits of grains do not accumulate at the top but are transported back into the bed.

The results of these drying experiments are presented in Table 14.22. Note that here the specific

TABLE 14.22
Drying Fine-Grain Calcium Carbonate in a Mechanically Spouted Bed Dryer with Inert Particles

<i>Experimental conditions</i>	
Inert charge	Glass beads ($d = 4$ mm)
Feeding	With a pump
Feed rate	3 kg/h of suspension of 50% m/m
Drying airflow rate	$110\text{ m}^3\text{/h}$ at 20°C
Temperature of drying air	135°C
<i>Drying results</i>	
Temperature of the outlet air	75°C
Mean moisture content of the product	0.4% m/m
Removed moisture	1.49 kg/h
Rate of evaporation referred to the cross section of the dryer	$95\text{ kg water/m}^2\text{h}$
Specific energy consumption	9715 kJ/kg water

energy consumption is rather high. The result is considered favorable as one must take into account also that after drying the dried product saved its primary grain size hereby, the grinding and screening operations became unnecessary.

14.3.4.5 Drying of Potassium Permanganate

Large-grain crystalline potassium permanganate after centrifuging can be dried in an SBD [24]. The 0.5–1.2 mm grain size fraction varied between 75 and 80%, but the fraction under 0.5 mm was around 20–25%. To dry this highly sticky material, 50 Nm³/h air of 260°C inlet (100°C outlet temperature) was used, bringing down its original 4% moisture content to 0.22%. Gas velocity in the gas inlet nozzles was 16.1 m/s; in the annular part of the drier it was 0.66 m/s. Specific air and heat consumption values were 16.6 Nm³ air/kg water and 7975 kJ/kg water. The results of similar drying experiments conducted in a fluidized bed dryer were 50–55 Nm³ air/kg water and 7350 kJ/kg water, that is, much higher air consumption with marginally lower heat requirement.

14.3.4.6 Dehydration of Salts

The dehydration of moist inorganic salts with water of crystallization content is, in general, a complex task. The loss of crystalline water usually consists of several components with different temperature or residence time requirements for each case. In the course of dehydration, the material to be processed (e.g., ZnCl₂ or MnCl₂) may form undesirable compounds. In numerous cases, as with magnesium chloride and iron chloride, oxygen impedes the full loss of crystal water as a result of heat effects. On the basis of these problems, publications dealing with successful dehydration experiments conducted in spouted bed equipment are of particular interest.

A diluted solution of manganese sulfate forms granular particles free of water if the solution is sprayed into the bottom of the spouted bed of manganese sulfate particles [24]. The SBD dimensions were $D = 196$ mm, $D_i = 60$ mm, and $\alpha = 40$. The operating data for this experiment are given in Table 14.23. The output of the dryer is 50 kg solution/h at a concentration of 20–25%. At concentration levels of 40–45%, an insufficient number of nuclei can be formed and the spouting motion stops [24].

Rabinovich [25] dried a diluted solution of MnCl₂ by spraying the 24–25% solution onto a 3.5–5 mm aluminum silicate catalyst acting as an inert packing bed or, alternatively, anhydrous MnCl₂ was the bed material.

TABLE 14.23
Drying of Solution of Manganese Sulfate

Airflow rate	180 Nm ³ /h
Gas velocity in the nozzle	80.6 m/s
Gas velocity in the cylindrical part	2.49 m/s
<i>Temperature</i>	
Inlet	600°C
Outlet	140°C
Specific requirement of gas	4.76 Nm ³ /kg water
Specific requirement of heat	3840 kJ/kg water
Specific evaporation	1260 kg water/m ² h

Ermakova [26] developed an SBD for the washing medium drained from an electrolysis bath containing manganese chloride and iron chloride. The inert packing bed was an aluminum silicate catalyst. To prevent hydrolysis, aluminum chloride was added to the washing medium (30–50 kg/m³). The inlet gas temperature was 500–550°C; the outlet temperature was 140–150°C. Gas velocity in the annular part of the SBD was in the range 3.2 to 3.6 m/s.

The importance of the residence time was observed by Mitev in the course of vortex bed dehydration of plaster of paris. At a bed temperature of 128–136°C, a residence time of 130 min was required to remove 1.5 molecules of water of crystallization. The average residence time of 30 min even at an inlet gas temperature of 800°C was not sufficient to attain the final desired moisture level. The cross section of the vortex bed dryer was 3.9 m², with an output of 6000 kg/h and a specific evaporating capacity of 1500 kg/m³h. The inlet temperature of the 2000 m³/h drying medium was 650°C; the outlet temperature was 150°C.

14.3.4.7 Drying of Pigments and Dyes

The treatise by Romankov and Rashkovskaya [23] outlines SBD procedures for a number of inorganic pigments and organic dyes. The authors detail methods of drying pastelike materials with 40–70% moisture content. No deterioration in the quality of the product was observed at relatively high (150–390°C) inlet temperatures. Some significant specific data about these dryers of 200-, 300-, and 870-mm diameter are as follows: 14.2–24 kg gas/kg water; 3250–4310 kJ/kg water; and capacity 55–90 kg water/m²h.

Similar values were attained in a spouted bed of industrial scale. For example industrial-scale drying of 200 kg/h “Z-type black” dye resulted in 11.8 kg gas/kg water and 4600 kJ/kg water; the diameter of the dryer was 1.6 m. The inlet temperature of 330°C in the spouted bed decreased to 100°C in the process

of drying. The initial moisture content of the wet material was 5%. Gas velocity in the gas inlet nozzle was 30 and 0.75 m/s in the annulus. Pressure drop over the 400-mm bed at this velocity was 5 kPa.

Drying of NaCl solution and alumina suspension was investigated [27] in laboratory-size SBD inert packing. Experiments have been carried out in a cylindrical chamber 150 mm in diameter with conical base and an air inlet of 20 mm. The inert particles were mainly glass and plastic beads of 2 to 5 mm size and different shapes.

To conclude, solutions and suspensions can be dried by spraying onto the outside of large inert particles. The drying proceeds at a constant rate until nearly all the water has evaporated. The drying rate then falls rapidly. The dried product does not attrite from the surface of the particles until the drying passes into the falling rate period.

14.3.4.8 Drying of Cobalt Carbonate

For the drying of cobalt carbonate suspension, bed height to column diameter ratio (H/D) has been changed at constant airflow rate and temperature, comparing specific drying performances related to equipment cross section and total inert particle surface, also calculating drying efficiency from theoretical and actual heat consumption necessary for the evaporation of water. The measured and calculated data are summarized in Table 14.24 [18].

It can be seen from the data that specific drying performance P increases and the heat consumption r_r necessary for the evaporation of water decreases with reducing ratio H/D_c . If the ratio H/D_c is large, only a definite fraction of the inert particles participate in the drying process, that is, their surface is coated by the wet material, which dries on it and subsequently abrades. This means that the fraction of particles not participating in drying process is relatively high. This fraction can be reduced with a decreasing H/D_c ratio, and the amount of heat relieved in such a way can be used for removing the moisture. Drying efficiency was highest at lowest bed height to diameter ratio ($H/D_c = 1.4$). The experiment performed on the pilot-plant dryer ($D = 0.38$ m) proved that drying can be performed at low H/D_c ratio with the same good efficiency, as under laboratory conditions (see Table 14.24).

The experimental results are interesting from the point of economic operation, that is, that one can obtain very high specific drying performance and low specific energy consumption at packing heights much lower than with conventional SBDs. Energy consumption for airflow can be reduced considerably by decreasing the volume of the inert particles of relatively high density.

TABLE 14.24
Drying Experiments in the Laboratory-Scale and Pilot-Plant Dryers

		Lab 1	Lab 2	Lab 3	Pilot
D	(m)		0.135		0.38
m_p	(kg)	9.2	7.0	5.2	55
A_p	(m ²)	2.8	2.1	1.6	12.7
H/D_c		2.5	2.0	1.4	1.2
V''	(Nm ³ /h)	80	80	80	620
v''	(m/s)	1.5	1.5	1.5	1.5
x_{in}	(kg/kg)	5.5	5.3	5.3	5.0
x_{out}	(kg/kg)	0.05	0.05	0.03	0.02
G	(kg/h)	3.7	4.0	4.7	37.0
T_{in}	(°C)	174	178	178	178
T_{out}	(°C)	64	59	61	61
W	(kg water/h)	3.1	3.4	3.9	30.7
P_w	(kg water/m ² h)	209	227	264	270
P_d	(kg dry/m ² h)	40	40	53	56
r_r	(kJ/kg)	3350	3350	2950	2900

14.3.4.9 Drying of Sludge from Metal Finishing Industries' Wastewater Treatment Plants

Metal-finishing industries, in particular the plating industries, typically produce large volumes of metal hydroxide sludges from treatment of rinse and wastewater [28]. These sludges typically contain 20–25% of solid after dewatering in a filter press. Spouted bed is potentially an excellent alternative for drying such materials. Tests were carried out in a 154-mm diameter, half-cylindrical spouted bed designed for operation at high temperature. The bed has a total height of 1156 mm and a conocylindrical base with an included angle of 60°. The sludge used for the trials is a mixture of ferric and zinc hydroxides containing 41% iron, 7.5% zinc, and 4.0% lead on dry basis with the balance represented by oxide and hydroxide. The nonbound moisture percentage of the sludge was 83.4%.

In all cases, sand was used as the bed material. Spouting air preheated to a controlled temperature up to 500°C was introduced through a single orifice of 19-mm diameter. After steady spouting had been attained at a bed temperature of 350°C and a superficial gas velocity of 1.4 m/s, sludge was fed approximately 3 mm into the spout. Particle size distribution of material was between 15 and 30 mm. The product had moisture content of less than 3%. The sand discharged from the spout was almost uniformly coated with a thin layer of dried sludge.

A conservative capital cost estimate was made for a 610-mm diameter SBD designed to dry 220 kg/h of wet sludge with 20% by weight solids. The cost compared favorably with available kiln-type technologies

and shows that the spouted bed is cost competitive with conventional technology.

14.4 ASSESSMENT OF DRYING RESULTS

On the basis of extensive results obtained at the Research Institute of Chemical and Process Engineering, University of Kaposvár (KE MŰKKI) on SBD techniques, it is established that with proper design and selection of proper operating parameters the SBD lends itself to a very wide range of applications in various industries, such as drying granular, pastelike, or pulpy materials with a wide range of possible particle sizes. SBDs are especially suited for drying of heat-sensitive materials with surface moisture, as well as those with bound moisture, such as seeds, foodstuffs, pharmaceuticals, and synthetic products, in one or two stages. Among advantages of the SBD are:

Intense particle motion: Good particle mixing prevents localized overheating and ensures the product's uniform moisture content.

The recirculating motion of particles ensures that during the residence time the drying particles contact the inlet warm air at regular intervals. The velocity of this recirculatory particle motion can be adjusted as required by varying the operating parameters, such as gas velocity and bed height, and geometry, such as size of the gas inlet nozzle, the use of an inner transport screw, and draft tube.

The residence time of particles may be changed and regulated within very wide limits, for example, by changing bed height or the use of suitable internal elements, such as partitions or draft tubes.

To dry materials with bound moisture (e.g., plastics), tangential air inlet and an inner transport screw are highly recommended since this way the volumetric gas velocity can be adjusted as required by the drying process regardless of the gas velocity requirement for particle motion.

14.5 DEVELOPMENT OF SPOUTED BED SYSTEMS

Development activities at the Pannon University Research Institute of Chemical and Process Engineering were carried out in two ways:

1. Development of dryer construction to improve the control of the dried product characteristics
2. Development of dryer construction to improve the drying conditions

14.5.1 DEVELOPMENT IN CASE 1

Particulate materials of high moisture content (suspensions and pulps) can be advantageously dried without heat damage in the MSB dryer with inert packing. The particle size of the product, one of the most important quality requirements, is controlled by the wearing time in the inert conveyor screw by its rotation speed and conveying length. However, the latter can be increased only simultaneously with the spouted bed height, requiring higher ventilation energy. To avoid this disadvantage a modified MSB dryer was developed. In order to increase the effectual conveying length of the inner screw independently of the spouted bed height, a tube of changeable length was built in the dryer to the top of the bed [34]. This tube serves as a house for the screw and hereby, the screw can carry the inert particles over the spouted bed surface, improving the wearing, grinding effect. The results were demonstrated by drying tests carried out with microwave pretreated potato pulp. The results proved that by this way the particle size can be controlled in wider range, independently of the bed height.

14.5.2 DEVELOPMENT IN CASE 2

On the basis of concerning publications and experimental results, heat treatment processes can be successfully (better product quality, economically) performed by the combination of microwave and convective heating. For this reason a combined (microwave + spouted bed) dryer was developed on a big laboratory scale. The prototype dryer consists of a so-called MSB device and of an equipment part for the microwave energy supply (see Figure 14.14). The combined dryer can be advantageously used for continuous drying or batch drying and for heat treatment of particulate agricultural and food products, as well as of heat-sensitive pulps and suspensions. The convective and microwave drying and heat treatment processes can be realized simultaneously or subsequently.

The prototype was put in operation successfully. Experiments were carried out to heat and dry germinated pea and moistened rice. In this way a good product quality and the required final moisture content could be reached [35].

14.6 CONCLUSION

Spouted beds for drying of granular materials, pastes, and slurries are an emerging technology. Although there are few large-scale industrial applications reported in the literature, the modified spouted beds display significant potential for the future applications.

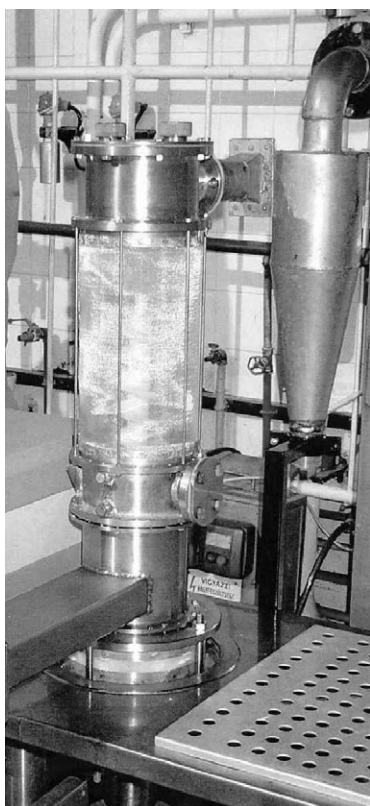


FIGURE 14.14 MSB dryer combined with microwave energy introduction (photo).

Scaling up of axisymmetric and three-dimensional spouted beds remains a difficult area. Two-dimensional spouted beds, such as those proposed by Mujumdar [29], have a decided advantage as far as scale-up and modular design are concerned. Also, spout-fluid beds (which combine the advantageous features of spouting with fluidization) with internal heat exchange surfaces are expected to find special applications. For a more comprehensive survey of the literature along with a discussion of the design of conventional as well as several modified spouted beds for drying, the reader is referred to Passos et al. [30]. This chapter is based to a great extent on results obtained at the Hungarian Academy of Sciences; other investigations reported in the literature (see Ref. [30]) are consistent with the findings and conclusions presented here.

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NOMENCLATURE

A_c	cross-sectional area of column, m^2
A_i	surface of inlet slots, m^2
A_p	surface of particle, m^2
A	specific surface, m^2/kg
D_c	column diameter (diameter of the dryer), m
D_i	diameter of the air inlet nozzle, m
D_{sc}	diameter of the conveyor screw, m
d_p	diameter of particles, m
G	wet feeding rate, kg/h
H	bed depth, m
h	height of slots, m
m_p	bed weight, kg
n_{sc}	speed of rotation of the conveyor screw, rpm
P_w	rate of evaporation related to the cross section of the dryer, kg/m^2h
P_d	specific drying performance (dry product) related to the cross section of the dryer, kg/m^2h
Q_1, Q_2, Q_3	bulk velocities in spaces 1, 2, and 3, respectively (see Figure 14.5), m^3/s
Q', Q''	bulk velocities in points a and b , respectively, m^3/s
q	conveying rate of the screw, kg/s
Re_{mf}	Reynolds number at minimum fluidization velocity
Re_p	particle Reynolds number
r_r	real heat of consumption, kJ/kg
s	pitch of the conveyor screw, m
T_{in}	inlet air temperature, $^{\circ}C$
T_{out}	outlet air temperature, $^{\circ}C$
v', v''	inlet air velocity, m/s
v_{ph}	peripheral speed of the screw, m/s
w_a	sliding velocity of particles in the annulus, m/s
w'_a, w''_a	sliding velocity of particles in the annulus at inlet air velocity v' and v'' , Equation 14.1, and air inlet nozzle D'_i and D''_i , respectively (Equation 14.3), m/s

$w_{aH}, w_{aH_{\max}}$	sliding of particles in the annulus at bed depth H and H_{\max} , respectively (Equation 14.2), m/s
W	evaporated water, kg/h
x_{in}	initial moisture content, kg/kg db
x_{out}	final moisture content, kg/kg db
Δp	pressure drop
ε	voidage of bed
ρ_f	density of fluid
ρ_p	density of particles
τ	time, residence time
τ_1, τ_2, τ_3	residence time in spaces 1, 2, and 3, respectively (see Figure 14.5)
τ_{in}	time of feed of material
τ	average residence time of particles

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15 Impingement Drying

Arun S. Mujumdar

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15.1 INTRODUCTION

Impinging jets of various configurations are commonly used in numerous industrial drying operations involving rapid drying of materials in the form of continuous sheets (e.g., tissue paper, photographic film, coated paper, nonwovens, and textiles) or relatively large, thin sheets (e.g., veneer, lumber, and carpets), or even beds of coarse granules (e.g., cat or dog food). In this chapter, we will not examine the last-mentioned application, which is a novel operation in which hot jets are directed normally onto thin beds of pellets transported on a slow-moving conveyor. The jets *pseudofluidize* the bed to ensure good gas–solid contact needed for effective drying.

Specific, large-scale applications of impingement drying, such as drying of tissue paper on Yankee

dryers, combined impingement and through-drying of newsprint (the so-called Papridryer process), and drying of wood, are covered elsewhere in this handbook. The objective of this chapter is to review the subject in more general terms and to provide empirical guidelines for the designing impingement systems on the basis of recent published literature on this subject.

Since impingement yields very high heat or mass transfer rates, it is a popular system for convective drying when rapid drying or small equipment is desired. High production capacities are attained at the expense of increased capital and operating expenses because of the more complex fabrication and increased air-handling requirements. Impinging jet drying is recommended only if a major fraction of the moisture to be removed is unbound. If the drying rate is internal diffusion controlled, the high heat transfer

rates of the impingement system can often result in product degradation if the product is heat-sensitive. For rapid drying of very thin sheets (e.g., tissue), high-velocity impingement of hot air jets is very effective. On the other hand, for drying of heavier grades of paper and textiles, for example, impingement drying is effective only to remove surface moisture. Another important industrial application of impingement drying is in the printing, packaging, and converting industry, in which printing techniques are used to deposit a thin film of coating onto a moving substrate.

It should be noted parenthetically that although most current applications use hot air jets for drying, the use of superheated steam in impinging jet configuration is seriously considered. At least one firm in India markets stenter dryers for textiles using steam jets. A combined impingement and through-drying scheme has also been proposed by the author for drying of permeable paper grades and textiles. Potential for the use of steam drying of veneer has also been explored and found to be favorable. Much developmental work needs to be done in the area of drying with steam jets.

It should be pointed out at the outset that since impinging jets are recommended only to remove surface or unbound moisture, the process calculation of such dryers is based essentially on the external heat or mass transfer rates, which can be estimated empirically for a wide variety of jet configurations. Much of the empirical information available in the literature was motivated by other than drying applications, such as heating or cooling of glass, metal sheets, turbine vanes, and furnace walls. This information can be readily adapted for dryer calculations if the geometric configurations are similar. It should be stressed that when the moisture removal is internal diffusion controlled, the design can be based only on experimentally determined drying rate data or on a validated mathematical model of the falling rate drying process.

If the internal resistance is large, designing a dryer to operate in a combination mode takes advantage of the high external transfer rates achievable with jets. Thus, for drying of foodstuffs extruded in the form of thin sheets, one may combine impingement drying with a microwave heat source that heats the solid volumetrically without depending on the development of large thermal gradients required by Fourier's law of heat conduction. Impingement may be combined with a through-flow to dry thin-packed beds or permeable sheets. In this case, application of suction has a synergistic effect: aside from through-drying, suction also augments the impingement heat or mass transfer rate.

In a few cases, the jet velocity is governed by the fluid mechanics of the impacting jet rather than by the heat or mass transfer rates. In drying of coated papers (films), very high shear stresses generated by jet impingement can cause the high-velocity ink or coating to flow, causing undesirable streaks (railroading) on the product. The jet velocity and spacing of the nozzle from the target surface should be adjusted to avoid such problems. Extremely high drying rates obtained by the use of high-velocity jets at elevated temperatures may cause such problems as case hardening or cracking due to rapid shrinkage. For combustible products, there is a significant fire hazard unless due care is taken in design and operation. Appropriate controls must be built in to avoid fire or explosion hazard.

In most industrial applications, the jet temperatures are such that the radiative heat transfer contribution is very small (a small percentage) of the convective. Thus, a design based purely on convective heat or mass transfer is conservative. This chapter will therefore consider only convective heat or mass transfer correlations for design. The reader is referred to the literature for full details.

15.2 DESIGN OF IMPINGEMENT DRYERS

15.2.1 GENERAL OBSERVATIONS

An impingement dryer system should be chosen only if the product is in a form that is amenable subjected to a multiplicity of hot jets directed normally (or nearly normally) onto one or both of its surfaces. In some cases (e.g., pulp or double-coated paper), it is possible to support the web to be dried entirely aerodynamically by strategically locating the blow boxes from which the jets emerge. In most other cases, the material that is dried is supported on a conveyor or roll, which may be solid or permeable. Because of their complex fabrication and high air-handling costs (due to the high nozzle pressure drop and high recycle ratio needed to achieve reasonable thermal efficiencies), impingement dryers are recommended only (or primarily) to remove unbound moisture. These should not be used to remove moisture well below the critical moisture level. Simpler parallel airflow dryers are just as effective and more economical under such conditions (e.g., wood and carpet).

Typically, the jet velocity and temperature may range from 10 to 100 m/s and 100 to 350°C, respectively, depending on the product. Because of the extremely high evaporative cooling in the constant rate period, the product surface temperature can be held well below its degradation or ignition temperature.

Care must be exercised in drying bound moisture under such conditions since the product temperature can rise rapidly. If the feed moisture or rate is susceptible to random variation or different products are to be dried in the same dryer, a suitable control system sensitive to the product surface temperature rise is recommended strongly.

If very high jet temperatures are needed, combustion gases may be used directly. Otherwise, steam-heated air may be used. When using high-velocity, high-temperature jets, up to 90% of the exhaust air may be recycled after reheating or directly mixed with the inlet jet air. Under these conditions, the jet contact time is too short to utilize the total moisture uptake capacity of the jet. Hence the need for such large recycle ratios, which result in expensive air-handling systems. For low-performance impingement dryers, the recycle ratios are typically low (less than 50%). In impingement dryers consisting of distinct compartments, each with its own operating temperature and air velocity, exhaust from the upstream compartment can be introduced directly into the downstream compartment, provided the exhaust air is still far from dew point, and can accomplish drying. In any event, fresh makeup air is needed to account for the moisture pickup; otherwise the drying medium will quickly become saturated. The dryer along with its ducting must be well insulated to obtain good thermal efficiency.

15.2.2 DESIGN PARAMETERS

The design of impingement dryers is both simplified and aggravated by the excessive number of design variables or parameters that can be specified or chosen arbitrarily; at least a dozen of these are important and have been studied systematically in recent literature. The designer's task is simplified by the fact that almost any arbitrary design can be made to work by adjusting one or several of the other operating parameters (e.g., jet velocity or temperature). The difficult task for the design is to choose a system that will give optimal system design (in terms of energy consumption and throughput and product quality) without undue adjustments during operation. It is also desirable to be able to control the dryer readily and be able to predict its performance when one or several of the operating variables are changed randomly.

Thus, it is apparent that the design problems may be posed in an infinite number of ways. The following is just one approach suggested by the author; no pretense is made that it is the most desirable approach. Most industrial designs are based on more arbitrary and trial-and-error procedures. However, field experience can be used to advantage in improving the design:

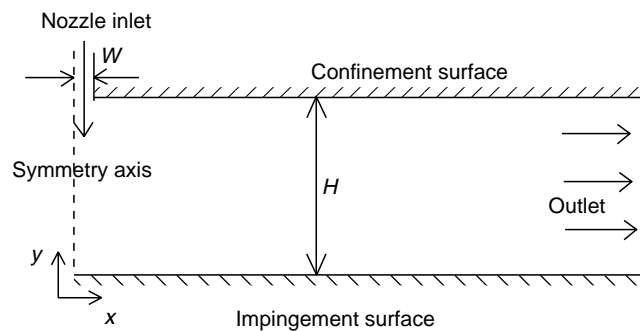


FIGURE 15.1 Configuration of flow field.

1. Select nozzle configuration (e.g., multiple slot and round jet arrays or exhaust port location) (see Figure 15.1 and Figure 15.2).
2. Select nozzle geometry (at least 50 different designs are in use, most of them chosen arbitrarily).
3. Select jet velocity, temperature, and nozzle-target spacing. (These are interrelated. A minimum geometric spacing is needed for fabrication purposes, and ease of access, for example. Jet temperature and velocity are dictated by product thermal sensitivity: the higher the temperature, the better the thermal efficiency will be in general. Jet Reynolds number must exceed 1000, probably 2000, to achieve fully turbulent regime.)
4. Calculate drying rates (constant rate only) using pertinent empirical correlations and applying various corrections for surface motion, high transfer rates, large temperature differences between the air jet and the web surface, and so on. Compute product surface temperature. Make a parametric study to determine quantitatively the influence of various parameters.
5. Determine air recycle ratio (by mass and enthalpy balance).
6. Redo steps 4 and 5, accounting for changes in jet temperature and humidity due to recycle. (Note: The recirculated exhaust may be heated in some instances.)

15.2.3 NOZZLE CONFIGURATION

Although not popularly recognized until recently, the selection of the nozzle geometry and multinozzle configuration have important bearing on the initial capital cost and operating costs, as well as the product quality (e.g., nonuniform moisture distribution). It appears that most of the old designs utilize nozzle shapes and configurations chosen arbitrarily. Once selected for ease of fabrication and economy of

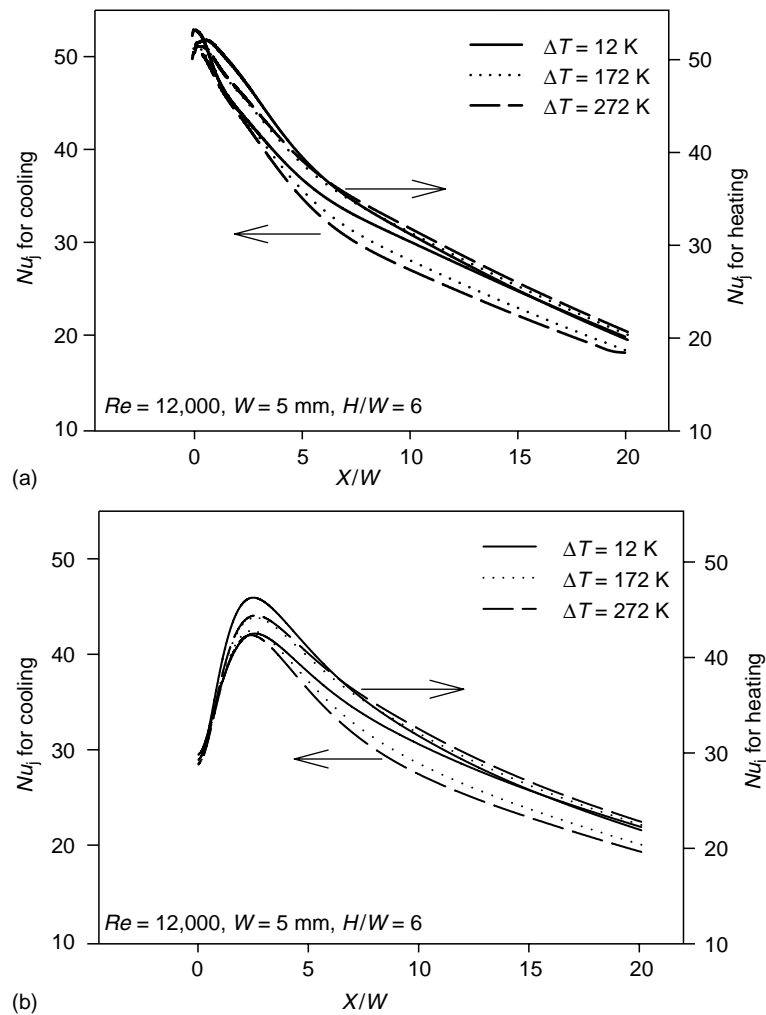


FIGURE 15.2 Effect of large temperature differences between the jet and impingement surface on jet Nusslet number for cooling and heating. (a) Standard $k-\epsilon$ model; (b) RSM model.

scale, the same nozzle box is retained in all subsequent designs. The dryer is typically constructed of a series of similar nozzle boxes arranged in a modular fashion with strategically located exhaust ports. The air extraction systems must be chosen to minimize any nonuniformities in the heat transfer distribution in the cross-machine direction. The seriousness of this factor depends on the nozzle configuration used; for example, it is more important for multiple slot arrangement than for round jet arrays. If the target surface moves rapidly with respect to the stationary nozzle boxes (i.e., moves at a speed of at least 10% of the jet velocity at impact), the nonuniformities in heat-mass transfer rate in machine direction tend to even out. On the other hand, such smoothing does not occur in the case of cross-machine direction nonuniformities if they persist throughout the dryer length.

15.2.4 SELECTION OF NOZZLE CONFIGURATION

The two basic multiple jet configurations for nozzles are arrays of round jets (issuing out of orifices or perforated plates) and rows of slot jets. In both configurations, suitable exhaust ports must be located to prevent the deleterious effects of crossflow due to spent flow from upstream jets. In general, the exhaust ports are located to remove the exhaust flow before it affects the heat transfer performance seriously. For jet arrays, it is common to provide a circular port or a rectangular slot for every five rows of staggered jets. For in-line arrays (not popularly used), the crossflow interference is expectedly much more significant. This is also true for slot jets since all of the spent flow must cross the downstream jet unless alternate paths of exhaust are provided. Care must be exercised in the design of extraction ports since these can lead to

lateral nonuniformity in heat or mass transfer rate distributions although the jet flow itself is uniform.

Slot nozzles are convenient and easy to fabricate for smaller width dryers (1–2 m). For larger widths and for dryers employing elevated temperature jets, it is difficult to maintain a uniform jet width without developing excessive thermal stresses. The fabrication costs are high and are not justifiable. Most large-scale units (e.g., Yankee dryer for tissue) therefore use perforated plates to generate arrays of jets. They are easier and cheaper to fabricate.

Obot et al. [1] have compared the heat transfer performance of jet arrays and multiple slot jets. The difference between these two configurations is too small to allow a logical choice to be made solely on thermal grounds. In general, either one may be chosen if adequate provision is made to extract the exhaust gas properly.

15.2.5 SELECTION OF NOZZLE GEOMETRY

Nozzle geometry effects can be profound, and it is recognized recently (e.g., see Refs. [2,3]). Contoured nozzles yield a uniform velocity profile and the exit plane but with a lower turbulence level. Although the pressure drop is reduced at the given flow rate (discharge coefficient in excess of 0.90) for contoured nozzles, the heat transferred per unit pumping power may not necessarily be higher. Indeed, the cost of fabrication of contoured nozzles (round or slot) precludes their industrial application.

Obot [2] has compared heat transfer and discharge coefficients for jets issuing from round nozzles and tubes. More recently, Hardisty [3] has presented extensive data for discharge coefficients and heat transfer under eight differently shaped slot jets. Regardless of nozzle shape, he found that narrower slots give higher heat transfer coefficients. He carried out tests for single slot jets to determine: (a) the effect of changing the shape of the nozzle holding the slot width constant and (b) the effect of changing the nozzle width while maintaining the same nozzle shape. These data are valid for single slots, but the relative trends may be expected to hold even for multiple jets. He presented correlations for the average heat transfer coefficient using the effective nozzle width w' (which is less than the geometric nozzle width w), which is related to the geometric nozzle width w by

$$w' = wC_D$$

where C_D is the discharge coefficient of the nozzle defined as $C_D = (\text{actual flow rate})/(\rho w L V_j)$. In this

equation, w is the nozzle width, L is the length of nozzle, ρ is the air density, V_j is the jet centerline velocity. Note that C_D can only be measured empirically. Florscheutz et al. [4] have shown that C_D for punched plates depends not only on the orifice geometry and Reynolds number but also on the array pattern and presence of crossflow. For small H , C_D may also depend on H .

It is impossible to give definitive recommendations for nozzle design. Simplicity of fabrication suggests the use of punched orifices or slots despite their low discharge coefficient values. More complex nozzle shapes are found in practice although their value in augmenting heat transfer rates is debatable. Often patent considerations dictate selection of the nozzle geometry and configuration. Regardless of the choice of the nozzle, it is important to provide proper exhaust of the spent flow to ensure that it does not cross the neighboring jets.

15.2.6 PRODUCT QUALITY CONSIDERATIONS

Depending on the type and quality of the product, the dryer designer will be limited in the choice of one or more of the design variables, namely, the type of nozzle (round or slot), jet temperature, and jet velocity. If the surface dried is sensitive to mechanical stresses (e.g., coated or printed sheet), the velocity at impact is limited to a value determined experimentally in laboratory tests. If cross-machine drying uniformity is critical (e.g., coated papers), slot nozzles are preferred over arrays of round holes although the latter may yield higher drying rates at a given open area; indeed, according to Pinter and Greimel [5], the heat transfer rates for round hole arrays may be 50–100% higher than those for slot arrays of the same open area (1–4%).

The energy consumption for an impingement dryer consists of two components: (a) electrical power for fan drives and (b) heat for the air jets. The electrical power needed to achieve a given heat transfer coefficient depends on the design of the nozzle, plenum chamber, jet velocity, recycle ratio, and pressure losses in the ducting. In general, an optimum value for the internozzle spacing exists for a given slot width. (See Ref. [5] for data on different slot widths, open area ratios, Reynolds numbers, and so on.) With good control of supply and exhaust air and good insulation, a well-designed dryer should yield a thermal efficiency of the order of 75–90%. Specific power consumption of 3100 kJ/kg of water evaporated has been reported for air-floater dryers. When jets are used for pneumatic conveyance of the web or sheet

to be dried, additional aerodynamic considerations need to be taken into account.

15.3 HEAT TRANSFER CORRELATIONS

15.3.1 RECOMMENDED CORRELATIONS

Obot et al. [1], Obot [2], Saad [6], Das [7], Martin [8,9], Dyban and Mazur [10], and several others have tabulated the numerous empirical correlations for stagnation and space-averaged heat transfer under single or multiple, round or slot turbulent jets impinging on stationary, impermeable, plane surfaces. It is not the intention of this chapter to list and review all these correlations. The reader is referred to the literature cited for details.

Since single slot or round jets are of no practical interest, correlation for this case will not be listed here. Only arrays of round and slot nozzles will be considered. Martin [9] has recommended the following correlations for these two configurations in the *Heat Exchanger Design Handbook*. Appropriate caution must be exercised by the user to account for various extraneous effects already noted and those discussed in subsequent chapters.

Figure 15.2 shows the spatial arrangement of round and slot nozzles in arrays and the averaging area appropriate to each. Let f is the nozzle exit area/area of square or hexagon attached to it. The following table defines f for the triangular or square pitch round nozzles and parallel slot arrays:

ARN_Δ:

$$f = \frac{\pi}{2\sqrt{3}} \left(\frac{D}{L} \right)^3$$

ARN:

$$\frac{\pi}{4} \left(\frac{D}{L} \right)^2$$

ASN:

$$\frac{w}{L}$$

where ARN_Δ is the array of round nozzles, triangular pitch, ARN is the array of round nozzles, square pitch, ASN is the array of slot nozzles, and L is the spacing between nozzles.

Martin gives the following correlations for these major configurations. For arrays of round nozzles (ARN),

$$\begin{aligned} \left(\frac{\overline{Sh}}{Sc^{0.42}} \right)_{\text{ARN}} &= \frac{Nu}{Pr^{0.42}} \\ &= \left(1 + \left(\frac{H/d}{0.6\sqrt{f}} \right)^6 \right)^{-0.05} \\ &\quad \sqrt{f} \left(\frac{1 - 2.2\sqrt{f}}{1 + 0.2(H/d - 6)\sqrt{f}} \right) Re^{2/3} \end{aligned}$$

The range of application is $2000 \leq Re \leq 100,000$; $0.004 \leq f \leq 0.04$; and $2 \leq H/D \leq 12$; the accuracy is $\pm 15\%$.

For arrays of slot nozzles (ASN),

$$\begin{aligned} \left(\frac{\overline{Sh}}{Sc^{0.42}} \right) &= \frac{Nu}{Pr^{0.42}} = \frac{2}{6} f_0^{3/4} \left(\frac{4Re}{f/f_0 + f_0/f} \right) \\ \text{where } f_0 &= (60 + 4(H/2w - 2)^2)^{-1/2} \end{aligned}$$

valid in the range $1500 \leq Re \leq 40,000$; $0.008 \leq f \leq 2.5 f_0$; $1 \leq H/S \leq 40$; the accuracy is $\pm 15\%$.

Martin [9] has also provided a detailed discussion on optimal spatial arrangements of nozzles. Here the term *optimal* means a combination of geometric variables that results in maximal average \overline{Nu} (or \overline{Sh}) for a given blower rating per unit area of the heat transfer or drying surface. Within engineering approximation, his optimization analysis yields the following results for both round and slot nozzles: optimal spacing for slots and optimal pitch for ARN are both $\approx 1.40H$, with $S_{\text{opt}} = 0.2H$ and $D_{\text{opt}} = 0.18H$. These optima should be taken only roughly. Other effects may overshadow these and shift the optima significantly.

15.3.2 EFFECT OF VARIOUS PARAMETERS

Following is only a brief outline of the various extraneous effects that have been investigated to varying degrees. For details, the reader is referred to the literature cited. The effects of crossflow, movement of the impingement target, large temperature differences, high drying rate, and artificial turbulence are among the most important to be considered in the design of impingement dryers. With large temperature differences between the jet and the surface, some radiant transfer may also be present. All dryers must be enclosed: the effect of confinement must be accounted for if the correlations used are for unconfined impingement. The effect of entrainment of low-temperature ambient air is always to reduce the temperature driving force and thus worsen the thermal performance. Infiltration of ambient air should be scrupulously avoided in impingement dryers, as in any other types.

15.4.3.1 Effect of Crossflow

Spent flow from individual jets or clusters of jets must be exhausted properly or the performance of adjoining jets can fall even by a factor of 2. For slot jets and rows or arrays of round jets, experimental studies have measured the effect of induced crossflow (i.e., crossflow due to upstream rows of jets made to cross the downstream jets to exhaust), as well as imposed crossflow (i.e., a flow parallel to the impingement surface that is not due to upstream jets). For ASN, the effect of crossflow is severe [6]. No more than two jet rows should be allowed prior to exhaust. For staggered ARN arrangements, the crossflow effect is less severe; an exhaust port may be provided for every three to five rows. After five rows, the crossflow effect drops the average Nu rapidly in the downstream direction [11].

If the drying performance of an existing dryer is to be improved by adding extra nozzles, care must be taken that the extra jets do not impair the performance of existing jets.

Another undesirable effect of crossflow is that, depending upon how it is directed with respect to the motion of the surface being dried, one may encounter a nonuniform moisture profile.

Recently, Galant and Martinez [12] have provided an empirical correlation for heat transfer under ARN subjected to known crossflow. As a rule, the designer should avoid the development of strong crossflows in the dryer by providing well-located exhaust ports. The work of Nagpal [13] and Kercher and Tabakoff [14] is also relevant when evaluating the effect of spent flow on ARN impingement.

15.4.3.2 Effect of Semiconfinement

Saad et al. [11], Obot et al. [15], Folayan [16], and others have observed that, because of the favorable pressure gradient induced by confining walls (parallel to the impingement surface), the heat transfer rates are reduced somewhat (10–20%) compared with free jet impingement. The reduction is more significant at lower spacings, whereas at larger spacings the effect may be neglected.

Most of the recent work, as opposed to the earlier work, is concerned with semiconfined impinging jets. The results are then applicable without correction for confinement.

In the absence of confinement, ambient air entrainment effects may cause unpredictable effects; for hot air impingement the heat transfer performance may worsen by 20–50% depending upon the flow rate, difference between the jet and the ambient temperature, and nozzle-surface spacing.

15.4.3.3 Effect of Mass Transfer

Large mass fluxes normal to the surface are caused by high evaporation rates; this is especially true of the high-velocity, hot air jet impingement employed in paper drying. The presence of normal flux causes thickening of the thermal-concentration boundary layer and hence a reduction in the heat and mass transfer rates. Since most correlations used in practice are for no net mass flux, corrections must be applied to the empirically determined \overline{Nu} or \overline{Sh} . Keey [17] has discussed the correction factors developed on the basis of film theory and boundary layer theory. For spray drying, the reduction due to high evaporation flux in the nozzle zone may be as high as 15%. Similar corrections (~10%) may apply in an intense impingement drying of paper.

More recently, Kast [18] has presented modified correction factors for high mass fluxes in laminar and turbulent flows. Although more complex in form, their effect on the calculated drying rates for superheated steam drying of paper was found by Loo and Mujumdar [19] to be negligible. The simpler forms given by Keey are therefore recommended for practical calculations.

15.4.3.4 Effect of Large Temperature Differences

Very little work exists on the effect of large temperature differences between the jet and the target surface. Das [7] has presented empirical correlations of the following form for a single slot jet; the correlation may be applied to ASN for large internozzle spacings and for exhaust slots located midway between adjoining slots:

$$\overline{Nu}_j = K Re_j^a \left(\frac{H}{w}\right)^b \left(\frac{T_j}{T_s}\right)^c Pr_j^{1/3}$$

His data and correlation apply over $5000 < Re < 20,000$, $8 < H/w/12$, and $1.18 \leq T_j/T_s \leq 2.06$. The values of a , b , and c are given in Ref. [7]. Because of higher viscosities of gases at higher temperatures, impingement dryers often operate with low jet Reynolds numbers (100–2000) for which no correlation exists for large temperature differences.

No correlations exist for other configurations. Correction factors recommended for tube flows may be applied for impinging jets in the absence of suitable correlations.

15.4.3.5 Effect of Suction

It is observed that a synergistic effect occurs when impingement drying is combined with through-drying,

as in combined drying of permeable paper or textile. Application of suction increases the impingement heat transfer rate. However, it is not clear if the added capital and operating costs of through-drying are justified in a general application. For design purposes, a 10–15% increase in the average impingement heat or mass transfer rate may account for the suction effect within engineering accuracy. Very limited experimental or analytic information is currently available to permit prediction of the suction effect.

It is worth noting that the product permeability in general will change with moisture content as drying proceeds. Thus, for a fixed applied pressure drop across the permeable target, the suction velocity will in general increase in the downstream direction.

15.4.3.6 Effect of Surface Motion

Van Heiningen et al. [20] have reported local Nu values for a confined single turbulent slot jet impinging on a rotating drum using a specially designed rig. Earlier, Fechner [21] made measurements of average Nu on rotating drums subjected to single and multiple slot jet impingement. On the basis of the limited data at hand, it appears it is safe to neglect the effect of surface motion if the surface linear velocity is less than 20% of the jet velocity at impact. The local Nu is altered significantly, but the average value is little affected. Work currently underway at McGill University will shed more light on the effects of surface motion with or without suction and oblique impingement [22,23].

15.4.3.7 Effect of Oblique Impingement

For rapidly moving impingement surfaces, it is likely that an optimum angle of impingement (other than normal) may exist. Korger and Krizek [24] found that, with oblique impingement, although the local Nu changed, the average value changed very little compared to normal impingement for a stationary surface. According to Baines and Keffer [25], the surface-averaged shear stress for a single turbulent slot jet impinging obliquely on a rotating drum was maximum for an angle of 60° between the target and the jet axis. A similar effect may be present for heat or mass transfer, although no analogy can be made between momentum and heat or mass transfer in the impingement region.

15.4.3.8 Miscellaneous

The effects of surface roughness, curvature of impingement surface, and artificially induced turbulence have been studied primarily with regard to their application in turbine vane cooling rather than drying (see Refs. [26–28]).

To tailor heat and mass transfer rates and their distribution on the impingement surface, the designer has numerous possible choices. One that is not commonly exploited is the use of nonuniform arrays (e.g., varying D with fixed or variable pitch for ARN or variable S or W for ASN), which can be tailored to meet the requirements of varying drying rates. This is particularly useful for drying in the falling rate period of heat-sensitive products.

The effect of swirl in the jet flow has been studied by Ward and Mahmood [29]. Since the \overline{Nu} decreases with increasing swirl number, it is clearly undesirable to introduce swirl in the jet flow.

On the other hand, an artificial increase in jet turbulence in the region of the target surface helps to augment the average heat or mass transfer rates significantly. Ali Khan and coworkers [30,31] have studied, for round jets, the effect of placing a perforated plate just upstream of the impingement surface. It is thus possible to manipulate the magnitude and distribution of Nu by varying few parameters. They used punched plates (diameter of holes = 3–10 mm, pitch = 4–36 mm with open area 0.63–51%). For a punched plate hole diameter is 0.06 times the jet diameter, they found that maximal augmentation occurred for the punched plate placed three times the punched plate hole diameter. Although optimal, this spacing is too close for practical utility. Further work is needed to exploit this technique of augmentation of impingement heat and mass transfer.

The use of turbulence generation placed in the nozzle or upstream is wasteful as the jet turbulence decays rapidly as it approaches the impingement surface. Thus the added pumping power leads to little or no increase in the impingement transfer rate.

15.4 RECENT DEVELOPMENTS AND CLOSING REMARKS

Impinging jet heat transfer is a very active area of research. Numerous papers have appeared in the recent literature, which deal with computational fluid dynamic simulations. The turbulent model predictions are still not very reliable or accurate but they are still useful to estimate trends for design purposes. Some new results dealing with single and multiple impinging jets issuing out of noncircular nozzles (e.g., square, elliptic, etc.) may be of future interest although currently round nozzles will continue to be the preferred choice in view of the ease of their fabrication. Since impinging jets of gas-particle jet can yield significant enhancement over single-phase gas jets, this would be an area of potential interest in special applications.

Very little information exists for the impingement heat transfer coefficient with high temperature jets. Das et al. [39] also presented data on the effect of large temperature differences on the local and average heat transfer rates under a confined single slot jet by experiments over a range of temperature differences, from 50 to 300°C [6]. More recently, Milosavljevic [40] presented an overview of an experimental investigation of impingement heat transfer rate at high air impingement temperatures from 100 to 700°C under arrays of round jets. The effects of impingement air temperature as well as velocity on the average heat transfer coefficients are reported. They also compared the measured heat transfer coefficients with those predicted by the correlation of Martin [8]. The experimental results indicate that the heat transfer coefficient drops over the air jet temperature interval, 100 to 400°C much faster than what Martin's correlation predicts. It reaches a plateau between 400 and 600°C

and again increases slightly over the interval 600 to 700°C. The comparison between the measured heat transfer coefficients and those predicted by the correlation of Martin shows that for jet temperature of 100°C the measured coefficients are relatively close to those predicted by Martin's correlations; however, the measured heat transfer coefficients tend to be somewhat lower than those predicted by Martin's correlations over the temperature range of 100 to 400°C. Over the temperature range from 400 to 600°C, the deviation starts to decrease. Note that no account was made for the effect of radiation. Refs [32–38] provide extensive reviews of impinging jet heat transfer literature.

Shi et al. [41] have conducted an extensive numerical study of the effect of large temperature differences ranging from 12 to 272°C, between the jet and the impingement surface on the local Nusselt number under a single turbulent slot impinging jet. [Figure 15.1](#)

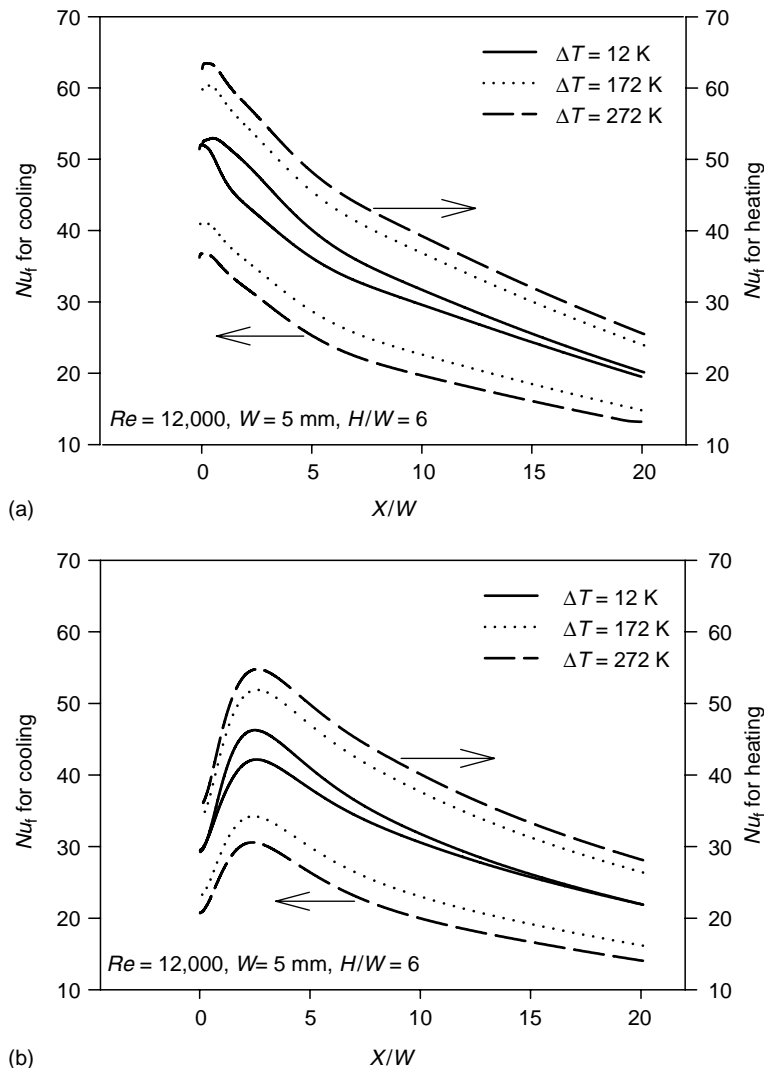


FIGURE 15.3 Effect of large temperature differences between the jet and impingement surface on film Nusselt number for cooling and heating. (a) Standard k - ϵ model; (b) RSM model.

shows a schematic diagram of the impinging slot jet flow configuration used. They distinguish heating and cooling applications since the thermophysical properties of the fluid in the vicinity of the target surface vary in different directions for the two cases. A comparative analysis in the turbulent flow regimes is made of the standard k - ϵ and Reynolds stress turbulence models for a constant target surface temperature. The local Nusselt numbers calculated using thermal conductivity values at the jet, film, or wall temperatures are labeled Nu_j , Nu_f , and Nu_w , respectively. Nusselt number distributions with different definitions of Nusselt number were compared. Their results show that large temperature differences between the jet and the impingement surface lead to significant differences in the numerical values of the heat transfer coefficients. At large temperature differences (over 100°C), significant differences occur between the three possible definitions of Nusselt numbers. This is true for both heating and cooling with impinging jets; only the former is of interest in drying. It is shown that use of the jet temperature as the reference temperature for the calculation of the Nusselt numbers shows the least spread, with ΔT (see Figure 15.2 and Figure 15.3). This result allows the designer to use previously published correlations, which were obtained at small temperature differences between the jet and the impingement surface. This conclusion is valid regardless of the jet Reynolds number and nozzle-to-surface spacing. Also, the same conclusion is drawn from the use of k - ϵ as well as the Reynolds stress models of turbulence. Although the results are for a single slot jet it may be assumed in the absence of additional results that a similar result will apply for arrays of jets.

NOMENCLATURE

H	nozzle-to-plate spacing
Nu	Nusselt number, hW/k
W	nozzle width
λ	thermal conductivity

SUBSCRIPT

j	inlet
f	film
w	impingement surface

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16 Pneumatic and Flash Drying

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16.1 INTRODUCTION

Drying is a separation process that converts a wet solid, semisolid, or liquid feedstock into a solid product by evaporation of the liquid into a vapor phase with the application of heat. Essential features of the drying process are phase change and production of a solid.

Thermal drying is one of the most important unit operations in most industrial sectors. Indeed, it is hard to find a product in daily use that has not undergone drying as a stage of its manufacture.

Drying is an essential operation in the chemical, agricultural, biotechnology, food, polymer, ceramic, pharmaceutical, pulp and paper, and wood processing industries. Drying is extremely energy-intensive and in many cases has important implications as the thermal energy needed for drying is obtained by combustion of fossil fuels, leading to emission of carbon dioxide. Well-designed modern drying equipment with high thermal efficiencies is becoming increasingly important.

One of the most widely used drying systems is flash drying and is also known as pneumatic drying. Flash dryers are most commonly direct drying units

and are also known as convective dryers. Pneumatic or flash dryers may be classified as gas–solid transport systems that are characterized by continuous convective heat and mass transfer processes. Hot air produced by indirect heating or direct firing is the most common drying medium in these systems. In direct flash dryers, the gas stream transports the solid particles through the system, and makes direct contact with the material to be dried. This gas stream (drying medium) also supplies the heat required for drying and carries away the evaporated moisture. Superheated steam can also be used as drying medium yielding sometimes to higher efficiencies and often to higher product quality.

The large surface area for heat and mass transfer and the high convective heat and mass transfer coefficients, which take place at these units, result in high drying rates and as a result, high drying capacity. The size of particulates to be dried is usually in the range of 10–500 μm . One of the features of these types of dryers is the relatively short contact time between the hot air and the particulate materials (0.5–10 s) at the drying section. Because of this the material temperature stays always low in the drying process.

16.2 BASIC OPERATION PRINCIPLE AND APPLICATIONS OF FLASH DRYERS

Figure 16.1 shows a simple pneumatic flash drying system in which particulate solids are dried during transport in a hot gas stream (usually air or combustion gases). The simple flash drying system includes six basic components: the gas heater, the wet material feeder, the drying duct, the separator, exhaust fan, and a dried product collector. The wet particles are fed into the hot gas stream sometimes with special mixing devices. The stream flows up the drying tube. The gas velocity must be greater than the free fall velocity of the largest particle to be dried. The gas velocity in relation to the particle velocity is high. Thermal contact between the conveying air and the solids as mentioned above is usually very short and therefore flash dryers are most suitable for removal of external moisture (surface moisture) and are less suitable for removal of internal moisture. At the end of the drying process a dust separation arrangement is installed. It must comply with the regulations for pollution control. For this purpose cyclone dust separators, fabric filters, electrostatic precipitators, wet scrubbers, and fabric filters are used.

High rates of evaporation in flash dryers are leading to low temperatures of the dried material and indicate that flash dryers are particularly useful for drying granular, crystalline, pasty, and powdery products, etc. Flash dryers are used in various branches

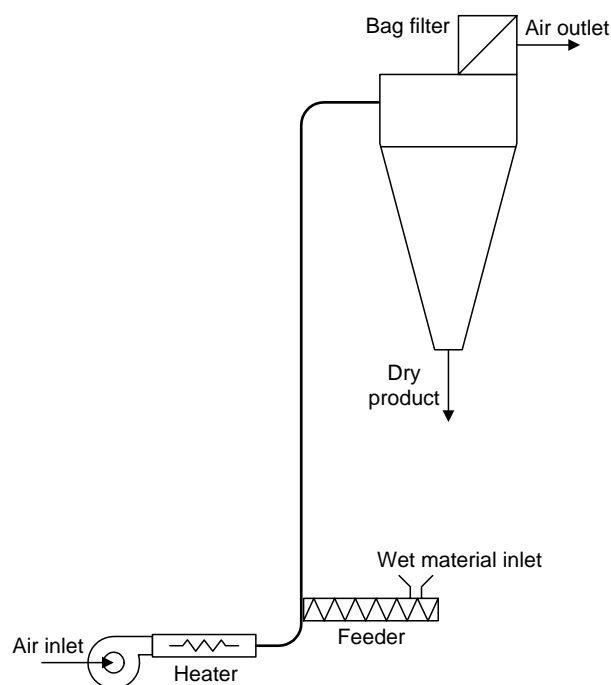


FIGURE 16.1 Simple flash drying system.

of the chemical, pharmaceutical, ceramic, gypsum, wood, and mining industries. Flash dryers are simple in construction and have low capital cost and they are almost trouble free.

Vertical type of construction, which facilitates installation in existing buildings, is advantageous for the flash drying systems. The tube of most flash dryers is of circular and uniform cross section. In some cases, the tube may diverge and converge and may have sudden expansions and contractions. The tube may be heated through the wall to keep up the temperature of the gas.

In order to shorten the drying time, recirculation of the material is used. In this case the number of cycles of different particles may be different and as a result the residence time of the particles will not be the same. In conclusion the advantages of flash dryers are the following:

- Short contact time and parallel flow make possible to dry thermolabile materials.
- The dryer needs only a very small area and can be installed outside a building. However, the gas cleaning system should be located inside the building in order to avoid moisture and dust deposition.
- The dryer is easy to control. The low material content in the dryer enables equilibrium conditions to be reached very quickly.
- Due to small number of moving parts the maintenance cost is low.
- The capital costs are low in comparison with other types of dryers.
- Simultaneous drying and transportation is useful for materials handling process.

The disadvantages of flash dryers are as follows:

- High efficiency of gas cleaning system is required.
- Because of powder emission, the dryer cannot be used for toxic materials.
- In some cases this disadvantage can be avoided using superheated steam as a drying agent.
- For lumped materials difficult to disperse, drying is impossible to carry out in this apparatus.
- There is a risk of fire and explosion, so care must be taken to avoid flammability limits in the dryer.
- In general, especially when recirculation is applied, not all material particles have the same residence time in the dryer [1].

In order to achieve efficient pneumatic drying process, the air velocity should be as low as possible to achieve materials transport, the mass flow rate of the gas should be the minimum necessary to achieve the

specified drying rate, the temperature of the hot gas should be as high as possible without exceeding limits imposed by the thermal sensitivity of the solids or safety considerations, and the construction of the dryer should allow to achieve thermal equilibrium between the gas and the solid [2].

16.3 DESIGN OF FLASH DRYERS

The materials dried in flash dryers have different properties and each product requires specific design solutions. It depends on the initial and final required moisture, temperature sensitivity, size and shape of the particles, etc. Finally, each product to be dried requires an optimum solution of the problems involved (efficiency and product quality).

Design procedure of dryers has to find:

- Dimensions of flash dryer
- Choose the type and amount of drying agent
- Requirement of energy supply
- Inlet and outlet parameters for the drying agent and material to be dried (temperature, moisture content, velocity, etc.)

Basically, the design consists of execution of the following steps:

- Heat balance
- Mass balance
- Momentum balance
- Determination of heat and mass transfer coefficients

By design of the whole drying system, care must be taken of the gas-heating unit, the material feed section, the particle separation section, and the product collection system.

The feed system has to be carefully chosen and designed in order to supply the wet material into the dryer at the required rate. Typical feed systems are shown in Figure 16.2 [3]. Metering and feed elements, sometimes with mixing devices arranged upstream lead the wet product into the flash dryer. For free-flowing powdery solids, a screw feeder or a rotary valve may be used effectively. Pasty or sticky materials need to be preconditioned by blending them with dried product using single- or twin-shaft paddle blender and then dispersed mechanically using a kicker mill or one of the several other designs of rotating disperser [4].

The selection of gas–solid separators is based mainly on the material characteristics, required degree of separation, concentration of solids, moisture content of solids, environmental regulations, and cost. In utilization are mainly: gravity separators; different types of cyclones with different efficiencies; fabric filters, and wet scrubbers. Usually a combination of separation units is applied.

There are different possibilities of modifications of simple flash dryers. For instance in a simple flash dryer a rapid decrease of drying force along the tube takes place. The design of a dryer with internal pipe through which hot drying agent is flowing leads to an increase of the driving force. Another possibility to increase the drying force is to heat up the tube of the dryer through the wall.

In order to decrease the dryer height and increase the drying time, the material is recirculated. In Figure 16.3 a two-stage system is presented [3]. The solid particles, after passing through the first stage, which has the form of a vertical tube, are separated in the upper part of the dryer, fall down, and are directed to the second stage. Wet material is supplied to the first stage by a special feeding system. Each drying stage in the system is equipped with its own heat generator. The drying agent from the second drying stage is fed

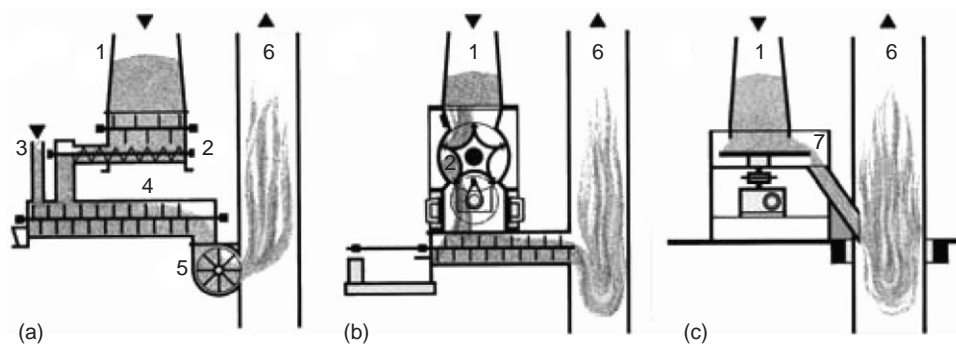


FIGURE 16.2 Typical feed systems for pneumatic flash dryers. 1, Wet product bin; 2, metering; 3, recirculated product; 4, mixer; 5, sling; 6, flash dryer tube; 7, disc feeder. (From *Flash Dryer*, Deutsche Babcock, Babcock-BSH GMBH, 1998. With permission.)

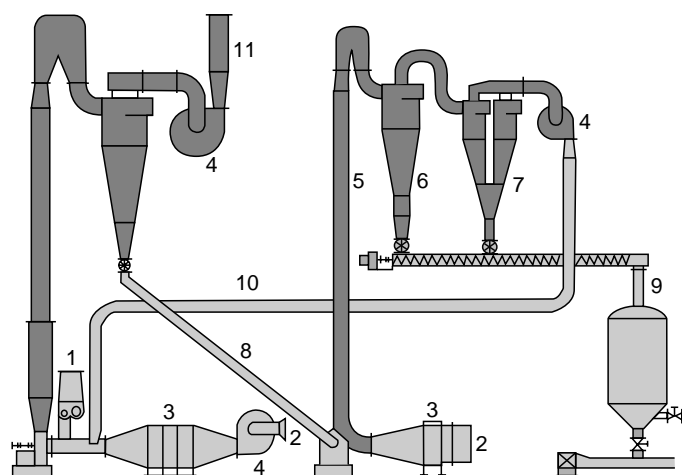


FIGURE 16.3 Two-stage pneumatic dryer with vapor utilization. 1, Wet product; 2, primary air inlet; 3, heat generator; 4, fan; 5, dryer tube; 6, cyclones; 7, cyclone separator; 8, predried product discharge; 9, dried product discharge; 10, vapor return line; 11, waste gas stack. (From *Flash Dryer*, Deutsche Babcock, Babcock-BSH GMBH, 1998. With permission.)

back to the first drying stage. This system is particularly efficient if the drying agent is superheated steam. Separation is by means of cyclone separators.

A second drying stage can also be used as a cooling stage. Two-stage flash dryers can be used for products difficult to dry, e.g., for methylcellulose. The number of cycles of different particles in two-stage systems may be different and as a result the resistance time of the particles will not be the same.

For longer resistance times the duct can be formed into a continuous loop (ring dryers). In these systems

the material is recirculated until it is dried to the required humidity. For instance high-temperature short-time ring dryers are used in the food industry to expand the starch cell structure in potatoes or carrots to give a rigid porous structure, which enhances conventional drying rates [5].

Figure 16.4 presents a spin-flash dryer that can be utilized for some special applications. As described in Ref. [4] the spin-flash dryer is basically a mechanically agitated fluidized bed device for very short residence times. Hence it is targeted for surface moisture

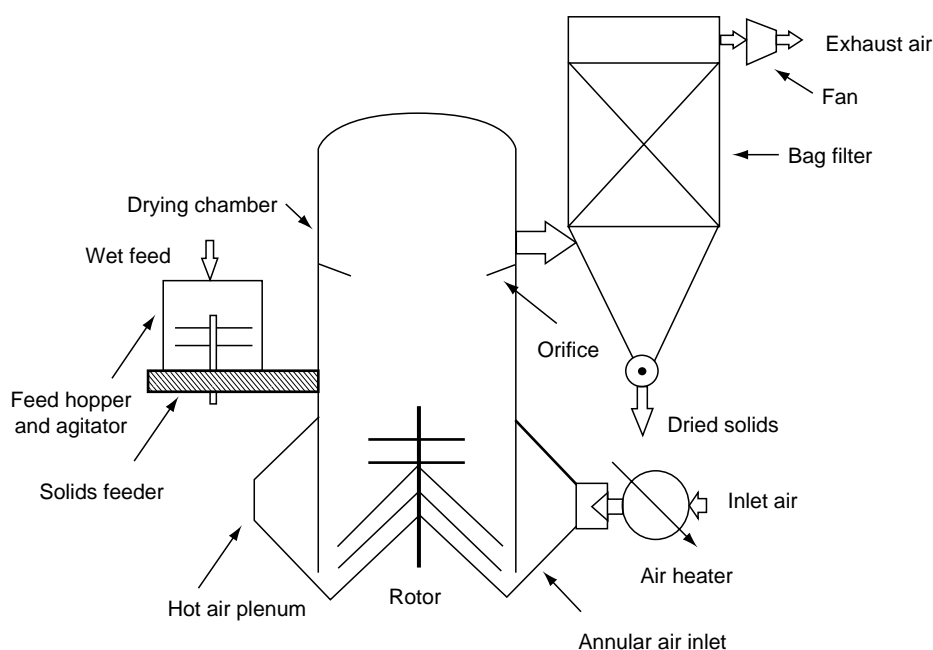


FIGURE 16.4 Spin-flash dryer. (From Devahastin, S. (Ed.), *Mujumdar's Practical Guide To Industrial Drying—Principles, Equipment And New Developments*, Exergex Corporation, Montreal, Canada, 2000. With permission.)

removal. As can be seen in Figure 16.4, a rotor, placed at the bottom of the chamber, is used to disperse the feed, which falls into the agitated fluidized bed by gravity. Hot drying air enters the chamber tangentially and spirals upward, carrying and drying the dispersed particles. The exhaust air containing the dried powder is entering into a separation device, which separates the powder from the exhaust air. Heavier wet particles remain within the drying chamber for a longer time and are broken up by the rotor. Thus only dried fine powder can escape to the gas separation system. This type of dryer can be a replacement for the more expensive spray dryer (which needs more thermal energy because the feed is wetter due to the pumpability requirements and also expensive because of the need for an atomizer). It is suited for drying sludge, pulps, pastes, filter cakes, high viscosity liquids, without the use of an atomizer. Numerous materials have been dried successfully in such units at capacities up to 10 tons/h. The spin-flash dryer units are more expensive than the conventional flash or fluidized bed dryers. Care must be taken to ensure that there is no danger of product accumulation on the walls due to stickiness.

Flash dryers using superheated steam as drying medium instead of air have some advantages such as no fire or explosion risk and higher efficiency (if exhaust steam is utilized elsewhere in the process). It is well known that in air-drying units the latent heat in exhaust gases is difficult to recover. Sometimes the quality of the dried product in superheated steam dryers is superior in comparison to air-drying units. The limitations of using superheated steam as drying medium are that the system itself and the operation of the system are more complex. Leaks are prohibited as noncondensables cause problems with energy recovery by condensation or compression of the exhaust steam. The feeding and discharge process must not allow infiltration of air and start-up and shutdown processes are more complex than for air dryers.

As mentioned by Devahastin [4] more recently flash dryers consisting of inert media have been employed at pilot scales to dry slurries and suspensions, which are sprayed onto them. The particles are coated thinly by the slurry and dried rapidly as a thin film. Due to particle-particle interactions, particle collisions, and shrinkage of the film in the drying process, the powder from the slurry is produced.

Flash dryers can be used as a drying stage in more complex systems for instance as a predrying stage to a fluidized bed dryer (batch or continuous) or fluidized bed cooler, spray dryer, drum dryer, etc. In the food industry flash dryers are often used after spray drying to produce foods that have a lower moisture content

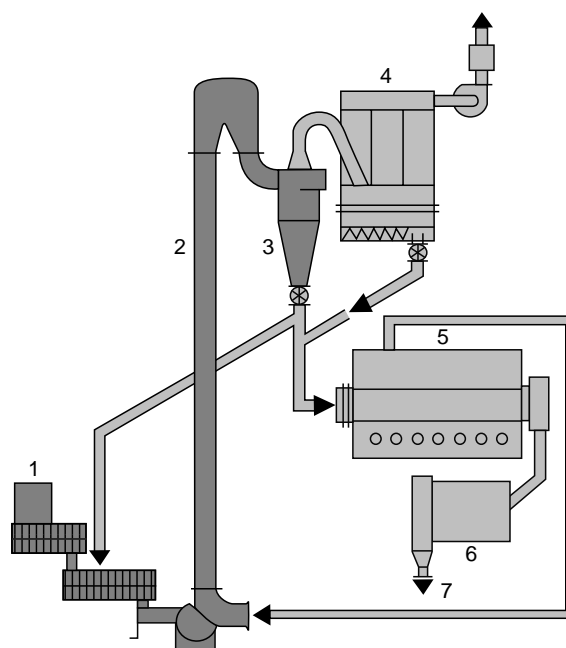


FIGURE 16.5 Flash dryer as a drying stage. 1, Wet product feed; 2, flash dryer; 3, cyclones; 4, fabric filter; 5, indirectly heated rotary calciner; 6, cooler, final product discharge. (From *Flash Dryer*, Deutsche Babcock, Babcock-BSH GMBH, 1998. With permission.)

than normal like special milk or egg powders and potato granules [5].

In Figure 16.5 a flash dryer is used as a drying stage in a complex system with a rotary calcining unit [3]. The drying agent in the flash dryer is indirectly heated in the calcining unit. Such arrangement provides sufficient heat utilization and is used by Babcock BSH for instance, for catalyst compounds and other products.

16.4 MATERIALS DRIED IN FLASH DRYERS

Flash dryers are suitable for drying granular, powdery, pasty, and crystalline products. As mentioned above the residence time of the particles in the dryer is very short that leads to the fact that only products containing surface moisture can be dried in flash dryers. Drying of heat-sensitive products in this type of dryers is very useful. Flash dryers are successfully used in the chemical, food, pharmaceutical, mining, ceramic, and wood industries. Some of the materials dried in flash dryers as described by Kisakürek [6] are:

Magnesium sulfate, magnesium carbonate, copper sulfate, dicalcium phosphate, ammonium sulfate and phosphate, calcium carbonate and phosphate, and boric and adipic acids are common examples of chemicals and by-products. Antibiotics, salt, blood clot,

bonemeal, bread crumbs, cornstarch, corn gluten, casein, gravy powder, soup powder, vegetable protein, spent tea, wheat starch, soybean protein, meat residue, and flour are examples of food products. Cement, aniline dyes, blowing agents, chlorinated rubber, coal dust, copper oxide, gypsum, iron oxide, and silica gel catalyst are typical by-products and minerals that can be dried in a very efficient way in flash dryers. Flash dryers are widely used in the plastic and polymer industries.

16.5 MODELING AND SIMULATIONS OF PNEUMATIC AND FLASH DRYERS

Mathematical modeling is a very important aspect in drying technology, allowing the engineer to choose suitable operating conditions for the chosen method of drying and if necessary apply scale-up procedures [7]. It should be kept in mind that the developed mathematical model should be experimentally validated in order to use it as a design tool. Reliable mathematical modeling for conveying of various powders in a dilute phase pneumatic conveying systems was developed and validated during the last three decades [8–11]. In a dilute phase flow, commonly referred as a suspension flow, the transport velocity is sufficient to ensure that the majority of the particles are suspended in the conveying gas. Since the particles in pneumatic and flash dryers are conveying in a suspension mode of flow, the various models, which were developed for pneumatic transport systems, were extended to model the flow in pneumatic and flash dryers by including heat and mass transfer between the particles and the conveying gas.

In general, two approaches can be used for modeling the flow through pneumatic and flash dryers. The first approach is based on empirical correlations for specific dryer and dried products. In this approach, a variety of semiempirical correlations [12–14] for estimating the pressure drop have been proposed for gas–solids flow in pipes. Frequently these models consider the total pressure drop as the sum of gas and solids pressure drop components:

$$\Delta p = \Delta p_g + \Delta p_s \quad (16.1)$$

This type of relationship is usually employed in the analysis of experimental data, where the total pressure drop is measured and the gas pressure drop component is evaluated by assuming that only gas is flowing in the pipe. A correlation may then be derived for the solids pressure drop component. Examples of this type of approach are the work of Muschelknautz and Wojahn [9], Pan and Wypych [12], and Mason

et al. [13]. Pan and Wypych [12] employed a modified version of Equation 16.1 by expressing the solids pressure drop as a function of the gas pressure drop multiplied by the solid loading ratio and a corrected friction factor as follows:

$$\Delta p = (1 + \alpha)\Delta p_g \quad (16.2)$$

$$\Delta p_g = 4f \frac{L}{D} \frac{1}{2} \rho_g U_g^2 \quad (16.3)$$

$$\alpha = \frac{\lambda_s}{4f} \frac{\dot{m}_s}{\dot{m}_g} \quad (16.4)$$

A similar approach was adopted by Mason et al. [13] and Bradley et al. [15] for estimating the pressure drop caused by bends in the pneumatic transport system.

In order to estimate the moisture content of the particle at the dryer outlet, two additional common assumptions are needed, namely isothermal flow and particle's exit temperature similar to the gas temperature. Based on these assumptions, various macroscopic mass and energy balance equations can be solved [16,17].

The second approach is based on theoretical and mathematical modeling for gas–particle flows. Three types of theoretical approaches can be used for modeling the gas–particle flows in the pneumatic dryer, namely two-fluid theory [18], Eulerian granular [19], and discrete element method [20,21]. Both the two-fluid theory and the Eulerian granular theory are based on macroscopic balance equations of mass, momentum, and energy for both the gas and the solid phases. It is assumed that both phases are occupying any point (x, y, z) of the computational domain with its own volume fraction. The solid phase is considered as a pseudofluid. The main difference between these theories is that the Eulerian granular method employs the kinetic theory of rarify gases to model the granular phase properties, such as pressure, temperature, viscosity, etc., whereas the two-fluid theory uses macroscopic correlations to model similar properties for the solid phase. It should be noted that traditionally, the two-fluid theory was widely used to model dilute phase flow whereas the Eulerian granular was used to simulate both dense and dilute phase flows. Unlike these theories, the discrete element method is an Eulerian–Lagrangian approach, in which the gas phase is assumed as the continuous phase, which occupies every point in the computational domain, and the solid particles are occupying discrete points in the computational domain. As a

consequence, mass, momentum, and energy balance equation should be solved for each particle within the computational domain. This method is able to take into account various types of particle–particle and wall–particle interactions from the basic dynamic approach and gas–particle interactions from the basic hydrodynamic models for the flow of a single particle through a conveying phase. Hence, there is no need to develop or to use macroscopic modeling for the transport of heat and mass from the solid phase to the conveying gas. This modeling needs large amount of memory and CPU time in order to solve real problems, which include millions of particles. As a result, no full-scale three-dimensional problem has been solved yet by using the discrete element method.

16.5.1 HYDRODYNAMIC MODELS

Many researches adopted one of the above-mentioned approaches and modified it to include various aspects of the pneumatic drying process. Andrieu and Bressat [16] presented a simple model for pneumatic drying of PVC particles. Their model was based on elementary momentum, heat, and mass transfer between the fluid and the particles. In order to simplify their model, they assumed that the flow is unidirectional, the relative velocity is a function of the buoyancy and drag forces, solid temperature is uniform and equal to the evaporation temperature and that evaporation of free water occurs in a constant rate period. Based on their simplifying assumptions, six balance equations were written for six unknowns, namely relative velocity, air humidity, solid moisture content, equilibrium humidity, and both solid and fluid temperatures. The model was then solved numerically and satisfactory agreement with their experimental results was obtained. Similar model was presented by Tanthapanichakoon and Srivotanai [22]. Their model was solved numerically and compared to their experimental data. Their comparison between the experimental data and their model predictions showed large scattering for the gas temperature and absolute humidity. However, their comparisons for the solid temperature and the water content were failed.

Mindziul and Kmiec [23–25] investigated the aerodynamics of the gas–solid flow in a pneumatic flash dryer. Their mathematical model was based on the continuity equation for both the gas and the solid phase and momentum equations for the solid phase and the solid–gas mixture. Heat and mass transfer were neglected. Although the drying apparatus was composed of three elements with varying cross-sectional area, one-dimensional model was solved. The effect of various empirical correlations for solid–wall friction factor has been investigated. The results

of the numerical calculations were partially compared with experimental data and the influence of the friction factor on the pressure, gas and particle velocities, voidage and residence time of particle along the axis of the apparatus have been presented.

Blasco and Alvarez [26] and Alvarez and Blasco [27] considered the application of flash drying to moisture removal of fish and soya meals. Heat, momentum, and mass balance equations were formulated. The model was solved numerically with appropriate coefficients of convective heat and mass transfer. Dilute phase transport of homogeneous radial monosize particle distribution was considered. The conveying superheated steam was assumed to be an ideal gas. The initial period for heating the particles, during which condensation takes place, was neglected. Using the film theory [28], the effect of the mass transfer on the heat transfer coefficient was considered. A variable diffusivity model was utilized for the prediction of the drying rate during the postcritical drying period. Using a pulse technique under isothermal conditions, the empirical parameters of the variable diffusivity model were experimentally determined. The predictions of the model were then compared with their experimental data and good agreement was presented.

Kemp et al. [29] presented a theoretical model for particle motion, heat and mass transfer, and drying rates in vertical tubular pneumatic conveying dryers. The model was one dimensional and it took into account particle–wall interaction, feed agglomeration effects, and the effect of particle shape on the drag factor. The flow pattern at the inlet, near the feed point, was neglected, i.e., fully developed flow. Kemp and Oakley [17] extended this model and employed it for simulating co- and countercurrent dispersion-type dryers. Equations for particle motion, heat and mass transfer, heat and mass balance, and local gas conditions were solved simultaneously over a small one-dimensional increment along the dryer. Using the Ranz–Marshall and modified Weber heat transfer correlations, the moisture content of the particles was underestimated. Similar observations were obtained by Baeyens et al. [30] and Levy and Borde [31]. Since the heat transfer correlations were obtained for a single particle, it is not therefore surprising that the proximity of other particles in the conveying system reduces the heat and mass transfer rates. In order to overcome this problem, Kemp and Oakley [17] applied a fitting mode procedure to achieve good agreements between their numerical simulations and the experimental data.

Silva and Correa [32] used DryPak for simulating the drying of sand in a pneumatic dryer. Their predictions were compared with the experimental results

and the two models of Rocha [33]. The main differences between the DryPak and Rocha mathematical models and their assumptions are given below.

16.5.1.1 Rocha [33] Models

The basic difference between both models is related to conservation equation of momentum. In the first model (Model a), the momentum conservation equation for the fluid as a mixture of fluid and particles was solved, whereas in the second model (Model b), conservation of momentum for each phase was solved. For both models, the following assumptions were considered: steady-state one-dimensional flow; nonhygroscopic spherical particles; no shrinkage during drying; plug flow for both phases; uniform properties and geometry at a pipe cross section; work done between the phases was neglected. Based on the above-mentioned assumptions, mass, momentum, and energy balance equations were formulated for the mixture and the solid phase. The correlation of Ranz and Marshall was used for calculating the heat and the mass transfer coefficients. Although Rocha [33] introduced a heat transfer term from the fluid phase to the ambient, no specific model was presented.

16.5.1.2 DryPak Model [34]

All the assumptions, which were considered by Rocha [33], were also considered in DryPak balance equations, with the exception that in DryPak adiabatic flow conditions were assumed. Other differences were in the way of calculating the area for heat and mass transfer and modification of the heat and mass transfer coefficients. DryPak used the Frossling equation for calculating the Nusselt number. Different types of heat and mass transfer analogies were presented and Ackermann correction was used to include the influence of mass transfer on the heat transfer coefficient. It should be noted that unlike Rocha [33] models, DryPak can take into account particles shrinkage; internal resistance to heat and mass transfer; and moisture content profile inside the particle could be obtained, although it was not used in the study of Silva and Correa [32]. Silva and Correa [32] concluded that predictions of DryPak produced better agreements with experimental data than the numerical results of Rocha [33].

Levy and Borde [35] adopted the two-fluid theory for modeling the flow of particulate materials through pneumatic dryer. The model was solved for a one-dimensional steady-state condition and was applied to the drying process of wet PVC particles in a large-scale pneumatic dryer and to the drying process of wet sand in a laboratory-scale pneumatic dryer.

A two-stage drying process was implemented. In the first drying stage, heat transfer controls evaporation from the saturated outer surface of the particle to the surrounding gas. At the second stage, the particles were assumed to have a wet core and a dry outer crust; the evaporation process of the liquid from a particle assumed to be governed by diffusion through the particle crust and by convection into the gas medium. As evaporation proceeds, the wet core shrinks whereas the particle dries. The drying process is assumed to stop when the moisture content of a particle falls to a predefined value or when the particle reaches the exit of the pneumatic dryer. The developed model was solved numerically and two operating conditions, adiabatic and given pneumatic dryer wall temperature, were simulated. Comparison between the prediction of the numerical models of Rocha and DryPak [34], which were presented by Silva and Correa [32], with the prediction of our numerical simulation revealed better agreements with DryPak than with the models of Rocha. The results of the developed model were also compared with experimental results of Baeyens et al. [30] and Rocha [33].

Rocha and Paixão [36] presented a pseudo two-dimensional mathematical model for a vertical pneumatic dryer. Their model was based on the two-fluid approach. Axial and radial profiles were considered for gas and solid velocity, water content, porosity, temperatures, and pressure. The balance equations were solved numerically using a finite difference method and the distributions of the flow field characteristics were presented. This model was not validated with experimental results.

Silva and Nerba [37] also used the two-fluid approach and presented a mathematical model of drying in cyclone. Slip condition of particles on the wall, particle-wall heat transfer, and particles shrinkage were considered. The mathematical model considered a steady state, incompressible, two-dimensional, axisymmetric, turbulent gas-solids flow. The gravity force effect on the particles was neglected. The particles were assumed to be spherical and distributed in a layer of uniform concentration on the cyclone wall and a very small concentration in the central flow. The discretized balance equations were solved by the SIMPLE algorithm [38]. Silva and Nerba [37] compared the predictions of their numerical simulations with experimental results and claimed that the most influencing parameters on the predictions are the particle slip conditions and the material shrinkage during the drying process.

Unlike the above-mentioned models, Fyhr and Rasmuson [39,40] and Cartaxo and Rocha [41] used an Eulerian-Lagrangian approach, in which the gas

phase is assumed as the continuous phase and the solid particles are occupying discrete points in the computational domain. As a consequence, mass, momentum, and energy balance equations were solved for each particle within the computational domain.

Fyhr and Rasmuson [39,40] presented a two-dimensional model for superheated steam drying of wood chips in a pneumatic conveying dryer. One-dimensional plug flow was assumed. Steady-state single particle flow and drying models were solved interactively. Particle-particle interactions were neglected. The irregular movement and the nonsphericity shape of the wood chips were accounted by measuring drag and heat transfer coefficients. The predictions of the temperature and the pressure profiles as well as the final moisture content of bark chips agreed well with experimental results. Based on the model validation, a parametric study was conducted. The calculation showed that the drying rate varies in a very complex manner through the dryer. The internal resistance to mass transfer becomes a dominant factor in the drying of less permeable wood chips. As the particle size was increased, the heat transfer rate decreases and the residence time increases. Hence, they concluded that less permeable wood species or larger chips size leads to longer dryer in order to obtain the desired final moisture content.

Another two-dimensional, discrete element model was presented by Cartaxo and Rocha [41]. In this work, only the dynamic phenomenon was investigated (i.e., heat and mass transfer between the phases were not considered). Thus the influence of the momentum coupling between the discrete particles and the conveying air on the air radial velocity and the mass concentration profiles were presented. An object-oriented numerical model was developed to simulate the conveying of large spherical particles (3 mm) through 9.14 m vertical tube with 7.62 cm bore size.

16.5.2 TWO-FLUID MODEL BALANCE EQUATIONS

In the following, the Eulerian governing equations for the pneumatic drying process are presented in their three-dimensional form. These equations are based on the two-fluid approach [18].

16.5.2.1 The Continuity Equations

The continuity equation for the k -phase is given by

$$\frac{\partial}{\partial t}(\varepsilon_k \rho_k) + \nabla \cdot (\varepsilon_k \rho_k \mathbf{V}_k) = S_k \quad (16.5)$$

where k -phase can be the gas or the solid phase, ε_k , ρ_k , and \mathbf{V}_k are the volume fraction, the density, and the

velocity vector of the k -phase. The mass source term of the k -phase is S_k and to maintain the conservation of mass $S_g = -S_s$.

16.5.2.2 The Momentum Equations

The momentum equation for the k -phase is given by

$$\begin{aligned} \frac{\partial}{\partial t}(\varepsilon_k \rho_k \mathbf{V}_k) + \nabla \cdot [\varepsilon_k \rho_k \mathbf{V}_k \mathbf{V}_k] \\ = -\nabla \cdot [\varepsilon_k \tau_k] - \nabla(\varepsilon_k P_k) + \varepsilon_k \rho_k \mathbf{g} + \mathbf{M}_{kj} + S_k \mathbf{V}_s \end{aligned} \quad (16.6)$$

Generally, the variation of the solid's density, which composes the wet particle, is less than few percent. As a consequence, constant solid density may be assumed to simplify the model. Thus by using the mixture theory the density of the dispersed phase can be expressed as

$$\frac{1}{\rho_s} = \frac{\xi}{\rho_w} + \frac{1-\xi}{\rho_{si}} \quad (16.7)$$

where ξ is the liquid mass ratio in the particle and ρ_w and ρ_{si} are the densities of the liquid and the solid, which compose the particle. Another common assumption is that the conveying air behaves as an ideal gas. Thus the gas pressure-density relation is described by

$$P_g = \rho_g \mathcal{R} T_g \quad (16.8)$$

The effective normal stress of the solids phase may be written as a sum of the sheared gas pressure and the solids contact stress. Thus the effective normal stress of the solid phase is described by

$$P_s = \rho_g \mathcal{R} T_g + \sigma_{n0}(\varepsilon_s/\varepsilon_{s0})^{1/\beta} \quad (16.9)$$

where σ_{n0} is a particular value of the solids contact stress for solid volume fraction ε_{s0} and β is a constant coefficient over a given range of contact pressure [42,43].

The interphase momentum transfer is represented by

$$\mathbf{M}_{kj} = K(\mathbf{V}_k - \mathbf{V}_j) + P_k \nabla \varepsilon_k \quad (16.10)$$

The interphase momentum transfer term can be derived from correlation developed to model fluidization processes, since the range of solids concentrations experienced in pneumatic transport systems is similar. This form has been employed by Patel and Cross [44] for modeling gas-solid fluidized

beds. For solids concentrations greater than 0.2, the interphase friction coefficient, K , may be computed by using the Ergun [45] equation

$$K = 150 \frac{\varepsilon_s^2}{\varepsilon_g} \frac{\mu}{d_s^2} + 1.75 \varepsilon_s \frac{1}{d_s} \rho_g |\mathbf{V}_g - \mathbf{V}_s| \quad (16.11)$$

For solids concentrations less than 0.2, the interphase friction coefficient is usually based upon the aerodynamic force on particle as follows:

$$K = (C_D \varepsilon_g^{-2.65}) \left(\frac{3\varepsilon_s}{2d_s} \right) \frac{1}{2} \varepsilon_g \rho_g |\mathbf{V}_g - \mathbf{V}_s| \quad (16.12)$$

where the single particle drag coefficient, C_D , is given by [46]

$$C_D = \max \left\{ \frac{24}{Re} (1 + 0.15 Re^{0.687}), 0.44 \right\} \quad (16.13)$$

and is modified to take account of multiparticle effects using the method of Richardson and Zaki [47]. The particle Reynolds number is given by

$$Re = \frac{\rho_g d_s (\varepsilon_g |\mathbf{V}_g - \mathbf{V}_s|)}{\mu_g} \quad (16.14)$$

The turbulent stresses, τ_k , in the momentum equations for the k -phase might be calculated by using the Boussinesq turbulent-viscosity model [8] for both phases or by applying a model of a Newtonian fluid for the gas phase and a granular shear stress for the solid phase [19].

The friction forces between each phase and the pipe wall can be modeled by adding a source term to the phase momentum equation for those control volumes adjacent to the pipe wall [11,23,39].

16.5.2.3 The Energy Equations

The conservation of energy in multiphase application can be written as an enthalpy equation for each phase:

$$\begin{aligned} \frac{\partial}{\partial t} (\varepsilon_k \rho_k h_k) + \nabla \cdot [\varepsilon_k \rho_k \mathbf{V}_k h_k] \\ = \varepsilon_k \frac{\partial p_k}{\partial t} + \tau_k: \nabla \mathbf{V}_k - \nabla \cdot \mathbf{q}_k + Q_k + Q_{kj} + S_k h_{kj} \end{aligned} \quad (16.15)$$

In this equation, h_k is the specific enthalpy of the k -phase, q_k is the heat flux, Q_k is a heat source term (due to chemical reaction or radiation), Q_{kj} is the interphase heat exchange between the phases, and h_{kj} is the

interphase enthalpy (i.e., the enthalpy of the vapor at the temperature of the solid particles).

16.5.2.4 Heat and Mass Transfer

The rate of energy transfer between the phases is usually expressed as a function of the temperature difference between that of the conveying gas and that of the particle surface (i.e., $T_g - T_{ss}$). Hence, the interphase heat exchange between the phases can be calculated by

$$Q_{gs} = \frac{6\varepsilon_s}{d_s} h_{gs} (T_g - T_{ss}) \quad (16.16)$$

The convective heat transfer coefficient, h_{gs} , is calculated from the Nusselt number, Nu , which is defined as

$$Nu = \frac{h_{gs} d_s}{k_g} = F(Re, Pr) \quad (16.17)$$

and is often expressed as a function of the Reynolds number (Re) and Prandtl number (Pr), which are defined as follows:

$$Re = \frac{\rho_g |u_r| d_s}{\mu_g}; \quad Pr = \frac{\mu_g c_{pg}}{k_g} \quad (16.18)$$

Note that k_g , μ_g , and c_{pg} are the thermal conductivity, the viscosity, and heat capacity of the gas phase, respectively. Table 16.1 presents common empirical correlations that have been used in the literature to calculate the heat transfer coefficient in gas-particle flows.

The mass transfer source term per unit volume can be obtained by multiplying the evaporation rate from a single particle, \dot{m}_s , by the total number of particles in the control volume:

$$S_g = \frac{6\varepsilon}{\pi d_s^3} \dot{m}_s \quad (16.19)$$

The drying model for a single wet particle and slurry droplet is based upon a two-stage drying process [48,49]. In the first drying period, the gas phase resistance controls the evaporation rate. Similar to heat transfer, this resistance is between the gas and the wet envelope of the particle. This may be expressed by

$$\dot{m}_s = h_m \pi d_s^2 \left(\frac{M_w p_{vo}}{\mathcal{R} T_{ss}} - \frac{M_w p_{vg}}{\mathcal{R} T_g} \right) \quad (16.20)$$

where h_m is the convective mass transfer coefficient, M_w is the molecular weight of the water, \mathcal{R} is the

TABLE 16.1**Empirical Correlations for Heat Transfer Coefficient in Gas-Particle Flows**

Modified Ranz–Marshall correlation [48]	$Nu = \frac{2 + 0.6Re^{0.5}Pr^{0.333}}{(1 + B)^{0.7}}$ $B = \frac{c_{pv}(T_g - T_d)}{H_{fg}}$	Developed for a single wet particle, taking into account the resistance of the liquid vapors around the particle to the heat transfer by Spalding number, B . c_{pv} denotes the heat capacity of the liquid vapors in the gas phase and H_{fg} is the latent heat of evaporation for the fluid
Modified Ranz–Marshall correlation [29]	$Nu = 2 + (0.5 Re^{0.5} + 0.06 Re^{0.8})Pr^{0.333}$	Takes into account turbulent boundary layer around the particle
Gamson correlation [30]	$Nu = 1.06Re^{0.59}Pr^{0.33}$	Developed for a fluidized bed dryer
De Brandt correlation [30]	$Nu = 0.16Re^{1.3}Pr^{0.67}$	Developed for a pneumatic dryer
Baeyens et al. [30] correlation	$Nu = 0.15Re$	Developed for a large-scale pneumatic dryer

universal gas constant, and p_{vo} and p_{vg} are the partial pressures of the water vapor at the particle crust and the gas phase, respectively.

The second drying period starts at a critical solid-to-liquid mass ratio, ξ_{cr} , which is obtained from a minimum void fraction, i.e., the porosity of the particles, ε (typically varied between 0.05 and 0.25). During the second period of the drying process, a dry crust starts to form, which causes a second resistance to mass and heat transfer. Thus, the wet particle consists of a dry crust surrounding a wet core. This resistance is governed by a diffusion process, which occurs between the outside diameter of the particle, d_{so} , and the diameter of the wet core, d_{si} . Assuming that the particle is not shrinking during the second drying period, the outside diameter of the particle remains constant and the diameter of the wet core decreases. The equation for the evaporation rate from a single particle is expressed as a Stephan-type diffusion rule [50]

$$\dot{m}_s = -\frac{d_{si} - d_{so}}{d_{so}d_{si}} \frac{2\pi\varepsilon D_v p}{\mathcal{R}T_{ave}} \ell n \left(\frac{p - p_{sat}}{p - \frac{RT_{ss}}{h_m \pi d_{so}^2 M_w} \dot{m}_s - \frac{p_{vg} T_{ss}}{T_g}} \right) \quad (16.21)$$

where D_v is the diffusion coefficient, p_{sat} is the saturation pressure inside the wet core, and T_{ave} is the average temperature of the particle.

In analogy to the heat transfer coefficient, the mass transfer coefficient h_m is calculated from the Sherwood number, Sh , which is equivalent to Nusselt number, Nu

$$Sh = \frac{h_m d_s}{D_v} = F(Re, Sc) \quad (16.22)$$

and is often expressed as a function of the Reynolds number, Re , and the Schmidt number, Sc , which is equivalent to Prandtl number, Pr , and is defined by

$$Sc = \frac{\mu_g}{\rho_g D_v} \quad (16.23)$$

The correlations for the Nusselt number (see Table 16.1) can be used to calculate the Sherwood number, Sh , and the mass transfer coefficient h_m by replacing the Prandtl number, Pr , with the Schmidt number, Sc .

During the first drying stage the diameter of the particle, d_s , shrinks due to evaporation from the outer surface to the surrounding gas. Thus the diameter of the wet particle can be calculated by

$$\frac{d}{dx} d_s = \frac{2}{\rho_w u_s \pi d_s^2} \dot{m}_s \quad (16.24)$$

At the second drying stage, the evaporation process of liquid from a particle is assumed to be governed by diffusion through the particle crust and convection into the gas medium. As evaporation proceeds, the wet core shrinks as the particle dries. In general, both the particle outer diameter and the wet core diameter can be shrunked, which may deform the particle's shape and size. In order to simplify the model, it was assumed that the particle's outer diameter remains constant during the second drying period. Thus, only the change of the wet core diameter, d_{si} , was considered:

$$\frac{d}{dx} d_{si} = \frac{2}{\varepsilon \rho_w u_s \pi d_{si}^2} \dot{m}_s \quad (16.25)$$

16.5.3 CASE STUDY

The two-fluid model has been used for modeling the flow of particulate materials through pneumatic dryer. The model was solved numerically for a one-dimensional steady-state condition and was applied to the drying process of wet sand in a pneumatic dryer. A two-stage drying process was implemented.

The predictions of the numerical simulations were compared with the experimental results of Rocha [33] (presented by Silva and Correa [32]) that were obtained in a 4-m high pneumatic dryer with diameter of 5.25 cm. In this study, 380- μm sand particles having density of 2622 kg/m^3 and mass flow rate of $4.74 \times 10^{-3} \text{ kg/s}$ were dried with $3.947 \times 10^{-2} \text{ kg/s}$ air mass flow rate. The comparison between the predictions of the numerical simulations and the experimental data for changes of gas temperature, solid temperature, gas humidity, and particle's moisture content with length under adiabatic and known wall temperature operating conditions is presented in Figure 16.6a–d, respectively. When known wall temperature operating conditions were simulated, it was assumed that in average the pipe wall temperature is just about the outlet air temperature, and it is falling linearly from 360 K at the inlet to 354 K at the outlet. In these figures the circle symbols represent the experimental data that were published by Silva and Correa [32] and the two solid lines represent the predictions of the numerical simulations for the adiabatic and known wall temperature operating conditions.

It is clearly seen that the numerical model predicted the gas and the solid temperature profiles (Figure 16.6a and b) very well when known wall temperature operating conditions were simulated. The maximum

relative error was 0.35 and 0.03%, respectively. When adiabatic flow condition was simulated, the gas temperature was overestimated and the maximum relative error was 5%. The predictions of the numerical simulations for the gas humidity (Figure 16.6c) were very good for both simulation conditions, i.e., adiabatic and known wall temperature. The maximum relative errors were 1.2 and 0.70%, respectively. The predictions of the numerical simulations for the particle moisture content (Figure 16.6d) were also very good for both simulation conditions, i.e., adiabatic and known wall temperature, although only two experimental data were given. The maximum relative errors were about 20% at the pipe outlet (i.e., when the particles moisture contents were approximately zero). A comparison between the prediction of the numerical models of Rocha and DryPak [34], which were presented by Silva and Correa [32], with the prediction of the numerical simulation revealed better agreements with DryPak than with the models of Rocha.

It should be pointed out that the two-fluid approach, as described and demonstrated in the previous sections, was widely used and validated for various types of pneumatic conveying systems and pneumatic flash dryers. Nevertheless, it is only one of various approaches that can be adopted. Section

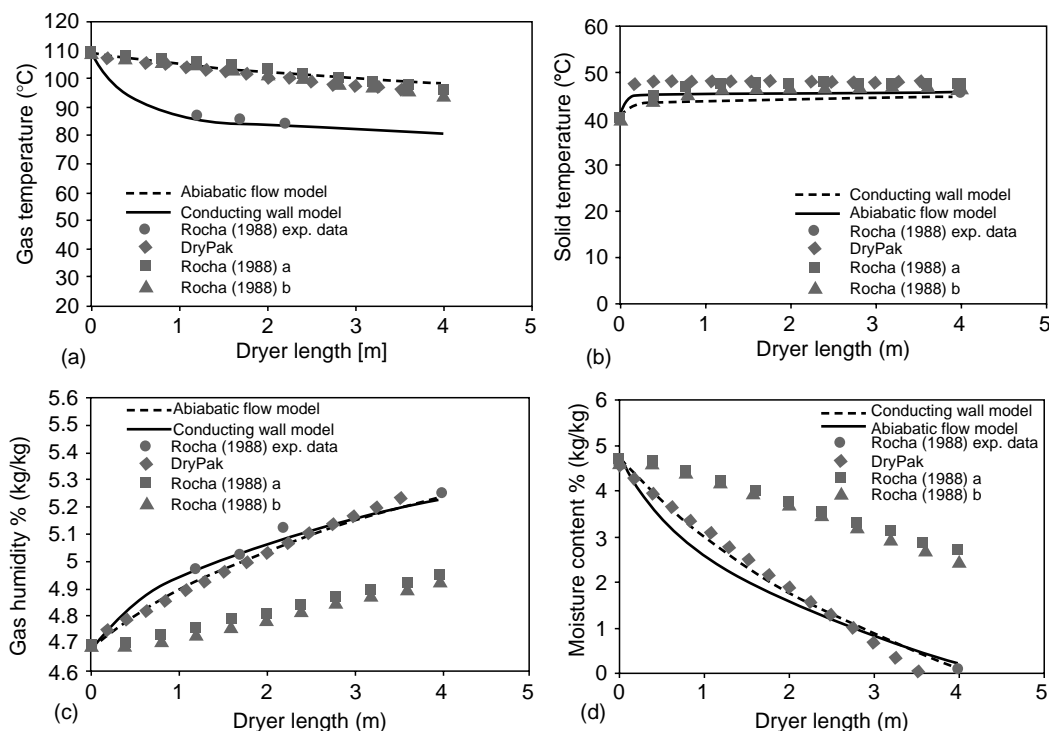


FIGURE 16.6 Comparison between the predictions of the pneumatic drying model, DryPak model [35], Rocha [33] models and the experimental data for changes of (a) gas temperature, (b) solid temperature, (c) gas humidity, and (d) particle's moisture content with length under adiabatic and known wall temperature operating conditions.

16.5.1 presents a summary of the various studies that were conducted on pneumatic flash dryers and the various mathematical models that have been adopted by them.

16.6 EXPECTED NEW DEVELOPMENTS IN FLASH DRYERS

These developments are expected in the fields of:

- Enhancement of product quality
- Increased efficiency by heat recovery
- Applications of heat pumps in drying systems
- Efficient combined processes
- Development of models for overall systems that would include mechanical dewatering, drying, heat recovery, powder collection, cooling stages, etc. in order to study interactions between various parts of the system and develop improved design procedures

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17 Conveyor Dryers

Dan Poirier

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17.1 INTRODUCTION

The conveyor dryer is conceptually very simple. Product is carried through the dryer on conveyors and hot air is forced through the bed of product. It is often described as simply a conveyor in a box with hot air. The reality, however, is that the conveyor dryer is one of the most versatile dryers available. Few drying technologies can match the conveyor dryer's ability to handle such a wide range of products. Products as varied in composition, shape, and size as coated breakfast cereals, nuts, animal feed, charcoal briquettes, and rubber can be dried in a conveyor dryer. Although it is simple in concept, an improper understanding of the heat and mass transfer processes in the conveyor dryer will surely lead to poor product

handling, wasted energy, and nonuniform product quality. This chapter describes the various conveyor dryer configurations that are available, typical products that can be dried in the conveyor dryer, and how to properly size and operate a conveyor dryer. This information will help anyone involved in conveyor dryer design, operation, or evaluation.

17.2 CONVEYOR DRYER CONFIGURATIONS

In general, the conveyor dryer is best suited for drying particulate material in the 1- to 50-mm diameter range. Drying in the conveyor dryer involves forcing air through the bed of product; therefore impesmiabie sheeted materials or slurries are not well suited to

conveyor drying. Ideally, the product can be stacked on the conveyor without excessive sticking or clumping. Stacking the product results in a much smaller dryer for a given retention time. Typical moisture ranges are as high as 95% for some fruits entering the dryer, down to less than 1% moisture on a wet weight basis for some synthetic rubbers at the dryer discharge. Although very light products are difficult to handle, they can be dried in a conveyor dryer if the dryer is properly designed to handle such products. For example, designing to have all down-flowing air through the beds will help to contain light products on the conveyor bed. Finally, typical drying times are 5 to 240 min. Drying times significantly longer than this typically result in such a large conveyor dryer, that the purchase price is difficult to justify.

There are numerous conveyor dryer configurations available in terms of conveyor and airflow arrangement. The following sections will describe some typical dryer configurations and list the products that are most often dried on the various configurations.

17.2.1 SINGLE PASS/SINGLE-STAGE DRYERS

The simplest conveyor arrangement is to have a single conveyor to carry the product through the dryer. Figure 17.1 shows a typical arrangement for a single pass dryer. Product is distributed onto the conveyor at the feed end and conveyed through the dryer. Once

the product is inside the dryer, air is forced through the bed by fans, which are typically mounted on the dryer enclosure. The airflow direction through the bed often alternates from up flow to down flow, in order to minimize moisture variations through the bed. Many products, such as packaged dry foods, must be cooled before packaging. A cooling section can be added to the end of the dryer to accomplish this cooling. If cooling to a temperature above ambient is sufficient, ambient air can be forced through the bed to cool the product. If cooling to a temperature below ambient is required (for example to set a coating by getting it below its glass transition temperature), chilled air is used as the cooling medium.

Figure 17.2 shows a cross section of a typical single pass dryer with a single plenum airflow arrangement. The plenums are the chambers located beside the conveyor bed, which collect the process air and channel it into and out of the conveyor bed area. In some configurations, there are two plenums rather than a single plenum. The dual plenum arrangement allows for greater flexibility in terms of airflow direction and also allows for more uniform drying.

The single pass arrangement offers a number of advantages. Typically, the dryer is separated into independent zones, which have their own heat source and circulating fans. The air temperature and air velocity can be controlled as the product progresses through the dryer. Another advantage of this arrangement is the ability to return the conveyor

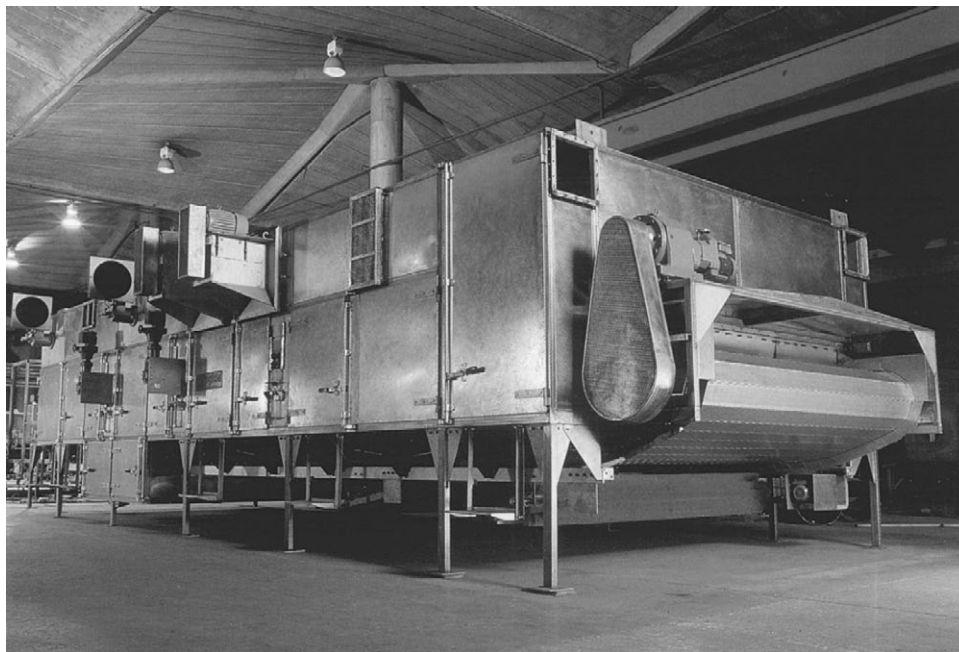


FIGURE 17.1 Typical single pass dryer.

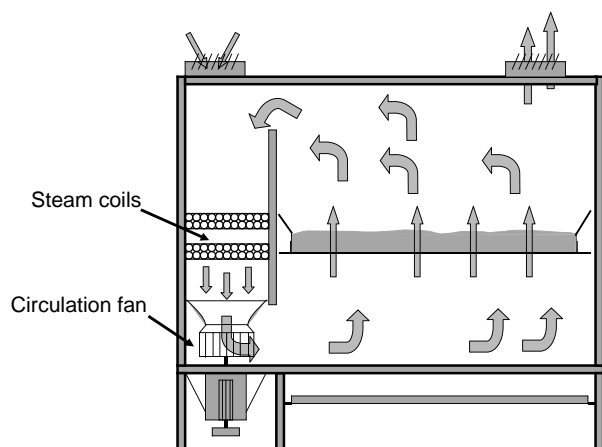


FIGURE 17.2 Typical cross section of a steam heated, single plenum, single pass dryer.

bed under the dryer, allowing easy access to the bed for on-line bed brushes or other bed cleaning accessories.

The single pass conveyor dryer is best suited to products that do not clump together or stick to the conveyor bed. This arrangement is also well suited for fragile products since the product does not get transferred from one bed to another as it progresses through the dryer. The disadvantage of the single pass/single-stage configuration is that the same bed depth must be used throughout the complete drying process.

Some examples of products that are often dried on single pass/single-stage dryers include expanded, non-coated snacks, baked snacks (for final moisture removal after the oven), nuts, charcoal, some synthetic rubbers, super absorbent polymers, etc.

17.2.2 SINGLE PASS/MULTIPLE-STAGE DRYERS

The multiple-stage dryer configuration overcomes the disadvantage of the single pass/ single-stage configuration, in terms of bed depth. Figure 17.3 shows a three-stage single pass dryer. Many products can be stacked much deeper after they have been partially dried than when they initially enter the dryer. The multiple-stage dryer configuration allows the operator to run the various conveyor beds at different

speeds. By running each stage slower than the preceding stage, the bed can be increased in depth as the product progresses through the dryer. This means that a given product retention time can be achieved in a smaller dryer compared to a single pass/single-stage dryer configuration. The turnover of the product from one stage to the other offers the added benefit of breaking-up clumps of products and reorienting the product for more uniform drying of products that tend to clump together.

The single pass/multiple-stage dryer configuration is the most versatile dryer configuration available. Not only can air temperature and air velocity through the bed of product be controlled as the product progresses through the dryer, but bed depth can also be controlled. The only drawback of this arrangement is the higher cost and relatively large floor space requirement.

Examples of some products typically dried on single pass/multiple-stage dryers are: french fries, coated cereals, nuts, fruits and vegetables, some synthetic rubbers, etc.

17.2.3 MULTIPLE PASS DRYERS

The multiple pass dryer offers many of the same benefits as the single pass/multiple-stage dryer, but in a much smaller footprint. In this arrangement, the conveyor beds are arranged one above the other, running in opposite directions. Figure 17.4 shows a three-pass dryer. The product enters the dryer on the top bed and cascades down to the lower beds. Two or three passes are typical, but it is not uncommon to have more than three passes in such dryers. Although the multiple pass dryer offers the same flexibility as the multiple-stage dryer in terms of bed depth control, it becomes difficult to zone the dryer such that air temperature and airflow can be controlled as the product progresses through the dryer. Such zoning is possible; however, complex airflow arrangements in the plenums and heat source areas are required. This arrangement also limits access to the beds for cleaning, compared to the single pass/multiple-stage dryer arrangement. In spite of these drawbacks, the multiple pass dryer is the most popular conveyor

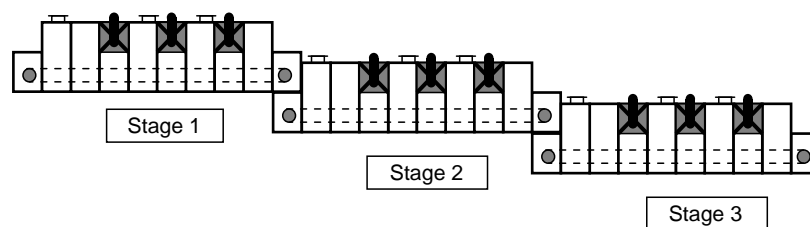


FIGURE 17.3 Three-stage, single pass dryer.



FIGURE 17.4 Three pass gas heated dryer.

dryer arrangement found in many industries. The relatively low cost and small footprint of this dryer arrangement and the ability to vary bed depths make this the ideal dryer for many processes. Some examples of products typically dried on multiple pass dryers are extruded feeds and cereals, some coated cereals and pastas.

17.3 CONVEYOR DRYER COMPONENTS

This section describes some of the key components of the conveyor dryer.

17.3.1 CONVEYOR ASSEMBLY

The conveyor assembly carries the product through the dryer. It must be strong enough to carry the load of the product and be able to run continuously for years with minimal maintenance. It must also be open enough to allow air to pass through the product, but at the same time the openings must be small enough to contain the product on the bed.

The conveyor bed itself is typically made of either a mesh belting or perforated sheet metal bedplates. The mesh belting is typically supported by slider bars or by a traveling support frame, which takes the load of the product. Bedplates typically include integral stiffeners and are designed to support the weight of the product without slider bars or any other supporting structure. [Figure 17.5](#) shows a typical hinged bedplate. The bedplates are typically 100–300 mm

(4–12 in.) wide in the direction of travel and 1–4 m (3.3–13 ft) wide across the machine. The openings in the bedplates are either round holes or slots. The slots will typically have a lesser tendency to plug than the round holes, except on some flat products that can get wedged into the slots. Choosing the correct perforation size and percent open area or mesh size is critical to the proper operation of the dryer. The openings must be small enough to contain the smallest product. Unfortunately, the smaller the opening, the higher the tendency for the openings to get plugged with either product, fines, or coating material. A plugged bed will not allow air to flow through, therefore drying rate and uniformity will suffer.

The conveyor bed can be driven by either driving the conveyor bed directly, or by driving bed chains attached to both sides of the conveyor. The arrangement with attached conveyor chains is the most common.

The materials of construction of the conveyor bed depend on the product to be dried. Products that are destined for human consumption or that are corrosive typically require stainless steel construction. AISI 300 series stainless steels such as 304 or 316 are the most common types of stainless steel; however, some AISI 400 series stainless steels such as 409 are also used. Other products are typically dried on carbon steel conveyor beds. As long as condensation on the bedplates is avoided when the dryer is not operated, a carbon steel conveyor bed can give the same service life as a stainless steel conveyor bed on noncorrosive products.

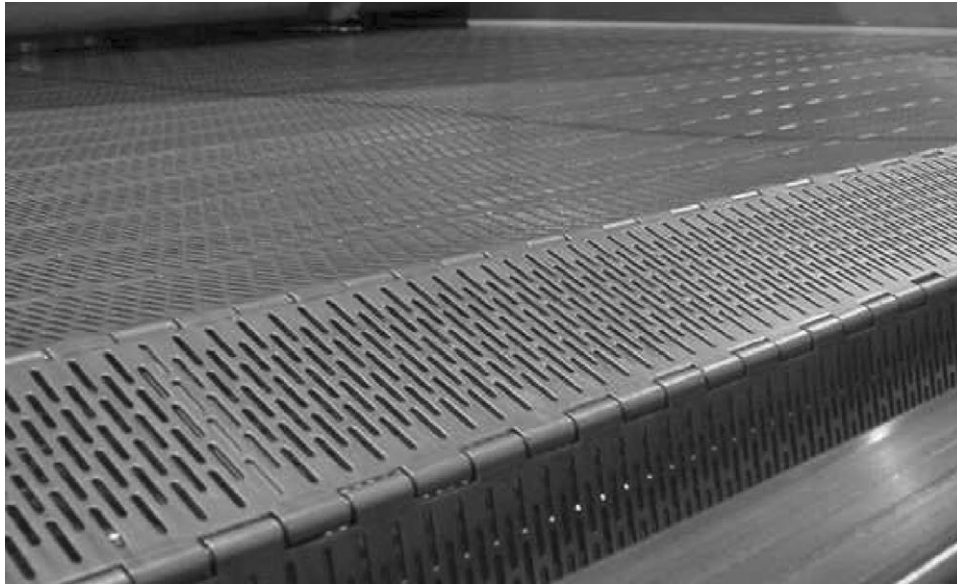


FIGURE 17.5 Typical hinged bedplate.

Since the product is typically stacked anywhere from 25 mm (1 in.) to 1 m (3.3 ft) deep on the conveyor bed, some means of containing the bed of product along the sides of the conveyor bed must be incorporated into the design. This is typically accomplished by side guides. Typically, traveling side guides attached to the bedplates contain the lower part of the bed and stationary side guides attached to the dryer frame contain the remainder of the bed. The gap between the traveling and stationary side guides must be tight enough to contain the product and to minimize bypassing of process air. In some cases, the stationary side guide can be made to seal directly to the moving conveyor without the use of traveling side guides.

17.3.2 DRYER ENCLOSURE

The dryer enclosure is typically constructed of a steel frame with insulated panels and doors. The enclosure must contain the hot air and resist corrosion. Dryers that will be used for human food drying to dry corrosive products or which will be “washed down” with water typically have a stainless steel enclosure. A properly designed dryer enclosure will allow thorough and easy access into the enclosure for cleaning, inspection, and maintenance.

17.3.3 AIR FANS

The dryer fans can be either integral to the dryer or externally mounted. Typically, the fans that circulate

the air through the bed are separate from the fan or fans that exhaust air from the dryer. The air circulation fans are typically “plug fans,” which plug into the dryer enclosure. The fan wheel of the plug fan is not in a fan scroll, but rather is an open wheel, which pressurizes the plenum or chamber in which it is mounted. The exhaust fans can also be plug fans, but are more often, separate scroll fans.

Proper fan design and fan specifications are critical to proper conveyor dryer operation. Most conveyor dryers will operate with a wide range of products and production rates. The fans must be properly selected to be as insensitive to these variations as possible. The most common type of fan used is the backward inclined fan. This fan is chosen due to its nonoverloading characteristics, high efficiency, low noise, and relatively low cost. [Figure 17.6](#) shows a typical backward inclined fan wheel.

17.3.4 HEAT SOURCE

Heat input to the conveyor dryer is typically by combustion of natural gas or by indirect steam heat through steam coils. However, a number of other heat sources are also commonly used. Some other common heat sources are LP gas, fuel oil, thermal oil, or electric heaters. The heat source is typically installed directly in-line with the circulated air. In some cases, however, such as when fuel oil is used on a human food product or when large amounts of dust are entrained in the airstream, the heat is transferred to the circulated process air indirectly through an air-to-air heat exchanger.

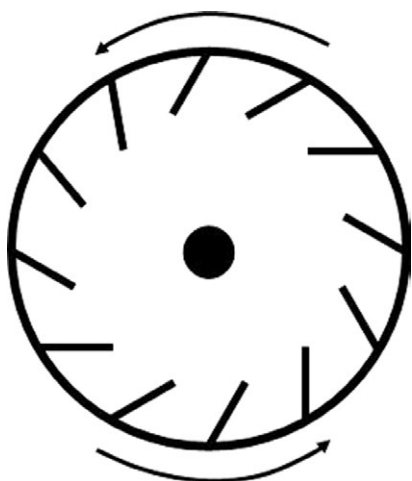


FIGURE 17.6 Typical backward inclined fan wheel.

In other cases, heated make-up air can be used as the source of heat rather than a heat source in the dryer. This is done when a stream of low humidity hot air is available from another process, or when direct combustion in the circulated process airstream may be hazardous.

17.3.5 PRODUCT FEEDER/SPREADER

One of the most critical components on the conveyor dryer is the product feeder or spreader. As in any convective drying system, it is imperative that airflow through the product be uniform. Convective heat and mass transfer is proportional to airflow velocity past the product.

Figure 17.7 illustrates the importance of uniform bed loading in a conveyor dryer. In a nonuniform bed, air will flow preferentially through the shallower areas

of the bed which present less resistance to airflow. These shallow areas will therefore be overdried compared to the rest of the bed of product.

Product can be loaded onto the conveyor in a number of ways. Oscillating feeders can consist of either an inclined spout or a conveyor (belted or vibratory), which oscillates from side to side across the bed (see Figure 17.8). Bias cut vibratory feeders consist of a vibratory conveyor, which becomes narrower as the product is conveyed across the bed. The product drops off the narrowing edge of the feeder onto the conveyor effectively, feeding the product across the bed. In some cases, it is preferred to simply feed the product onto the middle of the conveyor and spread the product afterwards. This can be accomplished by rotating paddle spreaders, which consist of a series of inclined paddles on a rotating shaft, or reciprocating-type spreaders, which “rake” the product across the conveyor by traveling back and forth across the bed. Products that do not clump together or compact, such as nuts, can be fed onto the bed using a simple hopper with an adjustable opening at the bottom.

Whichever type of feeder or spreader is used, its importance cannot be overemphasized. The characteristics of the product must be taken into account when selecting the type of feeder. If the feeding or spreading is not right, the dryer will not be able to dry the product uniformly. The end result will be poor product quality, wasted energy, and reduced capacity.

17.3.6 ACCESSORIES

Many products, such as coated cereals, will tend to clump together or stick to the conveyor bed as they dry in a conveyor dryer. Clumps of product are a

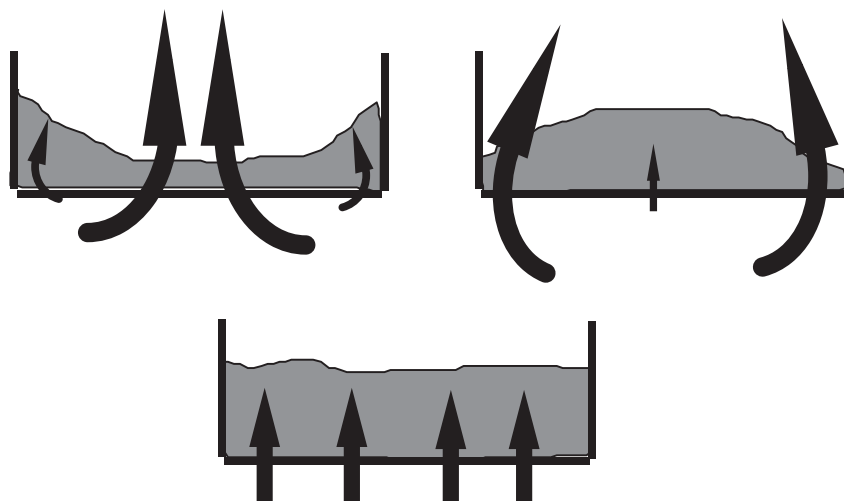


FIGURE 17.7 Effect of bed loading on airflow.



FIGURE 17.8 Typical oscillating spout feeder.

problem in the conveyor drying process since air will tend to flow around clumps rather than through them. This results in wet product inside the clumps. Product sticking to the conveyor bed is also a problem if the product does not properly transfer to the subsequent bed or discharge. A number of accessories are available to help address these issues. Lump breakers or pickers typically consist of rotating shafts with protruding fingers or tines. These fingers pass through the product and break the clumps or help to remove the product from the bed. These devices must be strategically placed. If they are placed too soon in the drying process, the clumps will re-form. If they are too late in the drying process, product damage can occur. Bed brushes (see [Figure 17.9](#)) can be used to clean the beds for those materials that tend to stick to the conveyor bed or plug the bed perforations.

Other accessories that can be added to a conveyor dryer include Clean-In-Place (CIP) systems, fire suppression systems, heat recovery systems, and automated control systems. Heat recovery systems, control strategies, and sensor options will be described further in a later section.

17.4 CONVEYOR DRYER SELECTION AND SIZING

The first step in selecting a conveyor dryer is to choose the proper configuration for the product that is produced. The various configurations available were described earlier in this chapter. In some cases, the dryer configuration is dictated not only by the product, but also by cost and available space. The single pass/multiple-stage dryer is the most versatile dryer configuration available; however, it also requires the most capital investment and the most floor space.

Once the dryer configuration is established, the size of the dryer can be determined. Sizing the dryer can be based on theoretical models, laboratory testing, or previous production experience with the product that is dried.

Although it would be nice to develop a generic model of conveyor drying that covers all products, theoretical modeling will never replace testing or previous experience. Many of the effects experienced on conveyor dryers are very difficult to predict. For example, the permeability of a bed of product will often

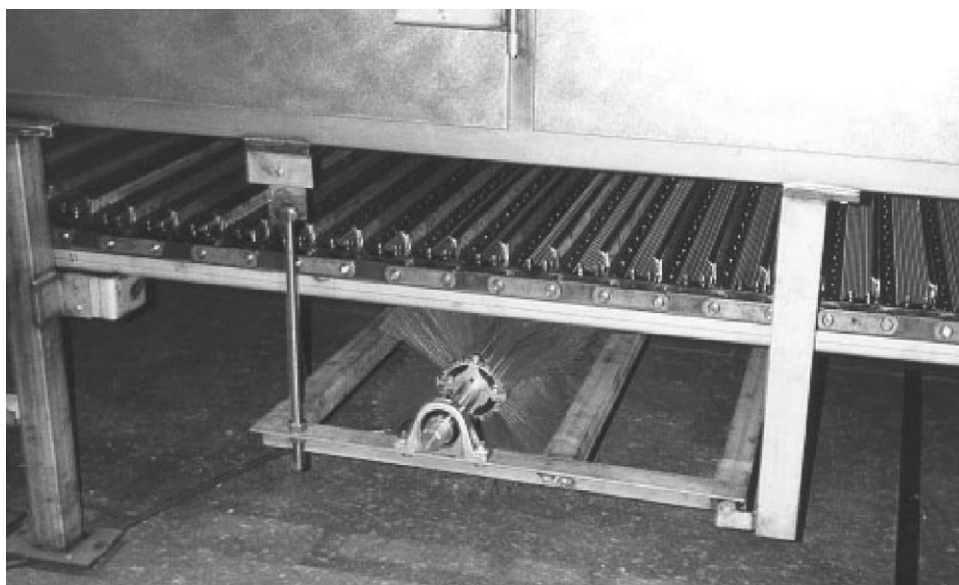


FIGURE 17.9 Typical bed brush.

not change linearly with increasing bed depth. With some moist products, the permeability of the bed will suddenly drop as the particles compress together and begin to blind over the bed. Other effects such as case hardening, product clumping, and product quality are also difficult to predict with a theoretical model. Often, laboratory testing is required to determine limiting temperatures for acceptable product quality. The taste of many food products, for example, will degrade if the temperature is too high. Most conveyor dryer manufacturers are only making limited use of theoretical models for dryer sizing. Typically, simple empirical models based on field or laboratory data are used.

Laboratory testing of small samples of the product can be very useful in determining the required size of a conveyor dryer. The continuous, conveyor dryer process can be simulated in the laboratory by a small, static batch tray dryer. The temperature and airflow direction can be changed with time to simulate the progression of the product from one zone to the other in the continuous conveyor dryer. The turnovers (transfer of the product from one conveyor to the other), however, are difficult to simulate in the static tray dryer. For this reason, some products are test run in a continuous laboratory conveyor dryer. This requires significantly more product than for the tray dryer testing, however the simulation is closer to the real-life production situation.

When performing laboratory tests or taking field data, it is useful to characterize the product in terms of a volumetric heat transfer coefficient. Since the product in a conveyor dryer is stacked, it is very

difficult to determine the amount of product surface area exposed to the drying air. This surface area is required to express the heat transfer coefficient in the typical form of

$$Q = h \cdot A \cdot \Delta T$$

where Q is the heat transfer (kW), h is the heat transfer coefficient ($\text{kW}/\text{m}^2 \cdot ^\circ\text{C}$), A is the exposed surface area (m^2), and ΔT is the temperature difference between the air and the product.

Since the exposed surface area is difficult to define in a stacked bed of product, it is much easier to define the heat transfer in the form

$$Q = h_v \cdot V \cdot \Delta T$$

where h_v is the volumetric heat transfer coefficient ($\text{kW}/\text{m}^3 \cdot ^\circ\text{C}$) and V is the volume of the product bed (m^3).

The volume of the bed is much easier to measure than the exposed surface area. Once the value of h_v has been determined for a given product and airflow, it can be used to predict the performance of the dryer for other air temperatures and operating conditions. When using this approach, it is critical to define how the term ΔT is calculated. The air temperature can change dramatically as it passes through the bed of product. It is best to define ΔT in terms of an average air temperature through the bed and the product temperature.

The most common way to size a conveyor dryer is to base the sizing on previous experience. Most

TABLE 17.1
Typical Retention Times for Conveyor-Dried Products

Product	Inlet Product Moisture (% wwb)	Discharge Product Moisture (% wwb)	Retention Time (min)	Air Temperature (°C)	Airflow Through Bed ((m ³ /h)/m ²)
Extruded feed	15–35	2–10	15–60	100–175	1800–3600
Fruits and vegetables	85–95	5–15	60–240	50–120	3600–7200
Coated cereal	7–16	1–5	5–20	100–150	1500–3000
Synthetic rubber	8–15	0–1	20–60	80–120	2500–4500
Charcoal briquettes	25–30	4–8	120–200	130–170	3500–6500

conveyor dryer manufacturers and many producers who use conveyor dryers have a large database of field test data, which can be used to size conveyor dryers. Typically, the sizing parameter is product retention time in the dryer, based on a given set of operating conditions. Table 17.1 shows some typical retention times for conveyor-dried products.

Once the dryer size and the configuration are determined, the size of all the components can be calculated. This is done by calculating a heat and mass balance on the drying process. This involves doing a heat balance, a mass balance on the water, a mass balance on the dry ingredients, and a mass

balance on any other volatile components. Table 17.2 shows a typical result from a heat balance calculation on a four heat zone, two-stage dryer. Heat zones 1 and 2 are in the first stage and heat zones 3 and 4 are in the second stage. The product is an extruded grain-based product.

The four heat load terms can be calculated in simplified terms as follows.

17.4.1 HEAT TO PRODUCT

The heat required to heat the product can be calculated based on the specific heat of the product and the

TABLE 17.2
Typical Heat Balance Calculation Result

	Zone 1	Zone 2	Zone 3	Zone 4	Total
Product infeed (kg/h)	6,786	6,009	5,614	5,230	—
Product infeed moisture (% wwb)	30	21	15.4	9.2	—
Product discharge (kg/h)	6,009	5,614	5,230	5,000	—
Product discharge moisture (% wwb)	21	15.4	9.2	5.0	—
Evaporation (kg/h)	777	395	384	230	1,786
Product infeed temperature (°C)	25	67	72	77	—
Product discharge temperature (°C)	67	72	77	80	—
Bed depth (mm)	100	100	250	250	—
Bed area (m ²)	15	15	15	15	60
Retention time (min)	8.6	8.6	21.6	21.6	60.4
Air temperature into bed (°C)	140	130	120	110	—
Air temperature out of bed (°C)	67	100	91	92	—
Airflow (normal m ³ /h)	25,000	25,000	25,000	25,000	100,000
Exhaust airflow (normal m ³ /h)	8,100	4,900	4,600	2,900	20,500
Heat to product (kW)	111	13	13	8	145
Heat to evaporated water (kW)	549	279	271	163	1,263
Heat to exhaust air ^a (kW)	108	116	97	63	383
Losses (kW)	10	10	10	10	40
Total heat consumption (kW)	778	418	391	244	1,831

^aThis term refers to the heat required to raise the incoming make-up air temperature to the exhaust air temperature. Remember that all the air in the conveyor dryer enters at the make-up air temperature and ultimately leaves at the exhaust temperature after having been recirculated through the bed numerous times.

change in temperature. The total heat transfer going to warm-up the product for the above example is

$$Q_{\text{Product}} = m_{\text{Product}} C_{p\text{Product}} \Delta T_{\text{Product}}$$

$$Q_{\text{Product}} = (5000 \text{ kg/h}) \times (1 \text{ h}/3600 \text{ s})$$

$$\times (1.9 \text{ kJ/kg}^\circ\text{C}) \times (80 - 25^\circ\text{C}) = 145 \text{ kW}$$

where Q_{Product} is the heat transfer to warm-up the product (kW), m_{Product} is the mass flow rate of product out of the dryer (kg/h), $C_{p\text{Product}}$ is the specific heat of the product out of the dryer (kJ/kg $^\circ\text{C}$), and $\Delta T_{\text{Product}}$ is the difference in temperature of the product out of the dryer compared to the product entering the dryer ($^\circ\text{C}$).

17.4.2 HEAT TO EVAPORATED WATER

The heat required to evaporate the water and to heat the evaporated water to the exhaust temperature can be calculated based on the change in enthalpy of the water. In the above example, the evaporated water enters the dryer as liquid water at a temperature of 25°C (enthalpy = 105 kJ/kg) and leaves as water vapor at an average temperature of 83°C (enthalpy = 2650 kJ/kg). The total heat transfer going to the evaporated water in the above example is

$$Q_{\text{Evaporation}} = m_{\text{Evaporation}}(h_{\text{Vapor}} - h_{\text{Water}})$$

$$Q_{\text{Evaporation}} = (1786 \text{ kg/h}) \times (1 \text{ h}/3600 \text{ s})$$

$$\times (2650 - 105 \text{ kJ/kg}) = 1263 \text{ kW}$$

where $Q_{\text{Evaporation}}$ is the heat transferred to the evaporated water (kW), $m_{\text{Evaporation}}$ is the mass flow rate of evaporated water out of the dryer (kg/h), h_{Water} is the enthalpy of the water in the product entering the dryer (kJ/kg), and h_{Vapor} is the enthalpy of the vapor exhausted from the dryer (kJ/kg).

17.4.3 HEAT TO EXHAUST AIR

The make-up air enters the dryer at 25°C in the above example and leaves the dryer at an average temperature of 83°C . The total heat transfer required to heat this air is

$$Q_{\text{Air}} = m_{\text{Air}} C_{p\text{Air}} \Delta T_{\text{Air}}$$

$$Q_{\text{Air}} = (20,500 \text{ nm}^3/\text{h}) \times (1.29 \text{ kg/m}^3)$$

$$\times (1 \text{ h}/3600 \text{ s}) \times (0.9 \text{ kJ/kg}^\circ\text{C})$$

$$\times (83 - 25^\circ\text{C}) = 383 \text{ kW}$$

where Q_{Air} is the heat transfer to warm-up the make-up/exhaust air (kW), m_{Air} is the mass flow rate of air

into the dryer (kg/h), $C_{p\text{Air}}$ is the specific heat of the air entering the dryer (kJ/kg $^\circ\text{C}$), and ΔT_{Air} is the difference in temperature of the air exhausting from the dryer compared to the air entering the dryer ($^\circ\text{C}$).

17.4.4 LOSSES

All dryers will have some thermal losses. The level of these losses depends on the amount of exposed surfaces, the amount of insulation, and the temperature differential between the process air and the ambient air. Calculating these losses is very much involved; however, they typically represent less than 7% of the total heat load in the dryer.

17.5 DRYER OPERATION AND CONTROL

In order to properly design a dryer operation and control scheme, it is imperative that the designer and operator have a good understanding of drying fundamentals and of how the conveyor dryer operates. This section describes the key variables in the conveyor dryer process and how to control them.

17.5.1 PROCESS AIR TEMPERATURE

The temperature of the air in the dryer is usually the primary variable used to control the drying rate. If the air temperature is increased, the increased heat transfer to the product will result in increased evaporation. Lowering the air temperature will decrease the drying rate. If the product is coming out of the dryer too wet, increasing the air temperature in the dryer will bring the discharge product moisture content back down to the proper value.

Many products have either maximum temperature limitations or minimum temperature requirements. Although the temperature of the product in the dryer is determined to a great extent by the air temperature, it is important to understand that the product in the dryer is typically at a much lower temperature than the air temperature due to evaporative cooling. If, however, the product is dried to a very low moisture content, near or approaching bone dry, the evaporation rate is reduced and the product may approach the air temperature.

In some processes, it is possible to automate the process air temperature adjustment. If on-line moisture meters are available, they can be used in a control scheme to automatically adjust the process air temperature. Because the retention time is typically quite long in a conveyor dryer, this type of moisture control typically works best if the moisture of the product entering the dryer can be measured and a feed-forward

control loop used to set air temperatures. Discharge moisture measurements can also be used with a feedback control loop, but if an upset occurs upstream of the dryer, in some processes, it can be hours before it makes its way through the dryer.

Other methods of moisture control employ temperature sensors coupled with models of the conveyor drying process. The temperature drop through the bed of product can be used to calculate the amount of heat transfer occurring as the air passes through the product. If, for example, the product moisture into the dryer suddenly drops, the heat transfer will be reduced. A temperature sensor on the return-air side of the bed will sense this drop in heat transfer as a rise in temperature and the control loop can take corrective action.

17.5.2 RETENTION TIME/BED DEPTH

The retention time, of the product in the dryer, is also a key variable used to control drying. The retention time is determined by the conveyor bed speed. If the beds are slowed down, the product will have more time in the dryer to dry, but the bed depth will be increased. To maximize production at the lowest possible air temperature, it is necessary to run the conveyor beds as slowly as possible (long retention time) without allowing the product to get excessively deep on the conveyor beds. If the product gets too deep, it may become difficult to force air through the bed of product. This loss in airflow can more than offset any gains in drying capacity due to longer retention time. With other products, excessive bed depths can cause product clumping or product damage. There is an optimal bed depth for each product and for each conveyor bed. Typically, the first bed in a multiple pass or multiple-stage conveyor dryer is run shallower than subsequent beds. Once the product surface is partially dry, the product can typically be stacked deeper without clumping or deformation problems.

17.5.3 PROCESS AIRFLOW

Airflow through the beds is often set during the design phase and is not a controllable variable. In some cases, where a conveyor dryer is designed to dry products with a wide range of bulk densities, some means of adjusting airflow through the bed is incorporated into the design. This could be a damper on the fan inlet or variable speed drives on the fans. When such airflow adjustment is available, the operator can reduce airflow for light products to minimize the chances of entraining product or fine particles in the airstream.

17.5.4 PROCESS AIR HUMIDITY

The humidity level of the air in the conveyor dryer is a critical control variable. Excessively high humidity levels will result in lost drying capacity and possibly dryer corrosion. If the humidity level in the dryer is too low, energy consumption may be excessive and with some products, case hardening may be a problem.

Due to the importance of process air humidity in terms of energy consumption and product quality, some conveyor dryer users have installed humidity sensors in the various heat zones of their conveyor dryers. The output from these humidity sensors can be tied to zone exhaust damper actuators or to a variable speed exhaust fan to adjust the amount of exhaust based on the humidity in the dryer.

17.6 HEAT RECOVERY

Section 17.4 summarized the energy flow in a typical conveyor dryer drying process. Although the majority of the energy goes to evaporate water, a significant percentage of energy is lost through the exhaust air. The simplest way to reduce this loss is to use preheated make-up air. This minimizes the amount of heat used in the dryer to heat the make-up air. In many cases, a conveyor dryer is installed in-line with a cooler. If the cooler uses forced convection with ambient air as the cooling medium, the spent cooling air can be used as make-up air to the dryer.

If no source of preheated make-up air (such as spent cooling air) exists, the enthalpy of the exhaust air can be used to preheat the make-up air through a heat exchanger. Typically, an air-to-air heat exchanger is used; however, in some cases it makes sense to use air to water heat exchangers and to preheat the make-up air using water coils. Although such heat recovery systems can often reduce total energy consumption by over 10%, the benefit must be weighed against the potential lost production if the dryer needs to be periodically shut down to clean plugged heat exchangers.

17.7 SANITATION

Sanitary design is critical to any conveyor dryer that will process food products. Sanitary design essentially refers to minimizing the possibility of microbiological growth in the dryer. For example, any crevice in the dryer which can collect fines from a food product, and which cannot be easily accessed for cleaning is a potential location for microbiological growth. This potential is increased in situations where the dryer is washed down with water. If the moist fines remain in

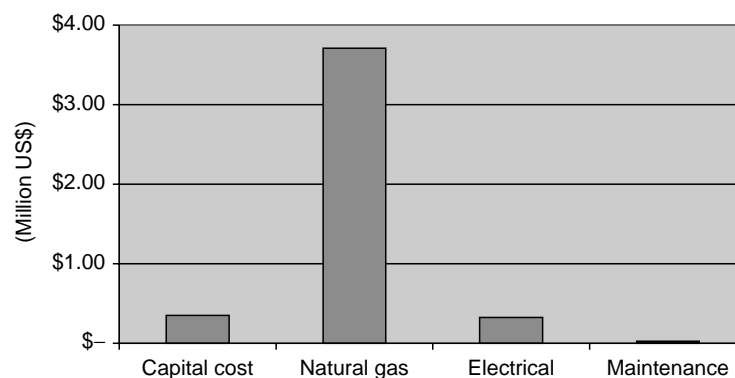


FIGURE 17.10 Typical lifetime cost of a 10 ton/h extruded petfood dryer based on 20-year life, expressed in year 2001 (US dollars).

the crevice without heating above the temperature necessary to kill the growth, a dangerous situation can occur. All food processors are aware of this potential and continually fight to minimize the risk.

Sanitary conveyor dryer design involves three basic steps. First, the product should be well contained on the conveyor bed. This means not only eliminating areas where product could leak off the bed, but also properly controlling the airflow in the dryer to minimize the possibility of fine particles being entrained in the process airstream. Second, the dryer design must minimize as many crevices, ledges, and dead airflow areas as possible. These areas will tend to collect the fine particles, which are entrained in the process airstream. Finally, the design of the dryer must allow easy access to all areas inside the dryer for cleaning. This may be the most important design criteria for achieving a successful sanitary dryer design. If the cleaning crew cannot easily access an area of the dryer, it will not be cleaned.

17.8 CONVEYOR DRYER ECONOMICS

As with any industrial dryer, the operating cost of a conveyor dryer is typically a significant contributor to the overall production cost of the finished product. A common mistake when purchasing a conveyor dryer is to underestimate the value of incremental improvements in dryer design. A small improvement

in drying uniformity can often mean substantial savings over the life of the dryer.

Figure 17.10 shows typical costs for a 10 ton/h petfood conveyor dryer. These values are based on typical extruded petfood (over 5 million tons of petfood are dried on conveyor dryers in the United States alone, every year).

Clearly, the capital cost of the dryer represents a very small percentage (typically less than 10%) of the overall cost of a dryer. Although the fuel cost for the dryer represents the highest cost of operating the dryer, the largest savings can typically be realized by reducing losses in production. This dryer will produce approximately US\$1 billion worth of product over its lifetime. Even a 0.5% loss in production due, for example, to poor moisture uniformity represents a larger cost item than the fuel cost. If the product (petfood in this case) is overdried to compensate for nonuniform moisture distribution in the product discharging from the dryer, this represents a loss in production since drying removes weight from a product that is typically sold by weight. A poorly operated or poorly designed dryer can also cause lost production due to off spec. product losses or other product losses in the dryer (e.g., product spilling onto the floor).

The bottom line is, do not be shortsighted when purchasing your conveyor dryer. The choices made at the procurement stage will affect you for years to come. A small saving at the procurement stage may result in large losses over the life of the dryer.

18 Infrared Drying

Cristina Ratti and Arun S. Mujumdar

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18.1 INTRODUCTION

Drying is the most common and most energy-consuming industrial operation. With literally hundreds of variants actually used in drying of particulate solids, pastes, continuous sheets, slurries or solutions, it provides the most diversity among chemical engineering unit operations.

One of the increasingly popular, but not yet common, methods of supplying heat to the product for drying is infrared (IR) radiation. Although this type of heat transmission was used incidentally in the past accompanying other types of heat transfer during dehydration, IR dryers are now designed to utilize radiant heat as the primary source (Williams-Gardner, 1971).

The most common current applications of IR drying are in dehydration of coated films and webs and to correct moisture profiles in drying of paper and board. Theoretical work and laboratory-scale experimental results on IR drying of paints, coatings, adhesives, ink, paper, board, textiles, etc. can be found in the literature (e.g., Navarri et al., 1992; Kuang et al., 1992; Therien et al., 1991; Cote et al., 1990). On the other hand,

reports on IR drying applied to other products like foodstuffs, wood or sand are not very common as yet. Most published data on IR drying of foods comes from (the former) USSR, the United States, and the East European countries (Hallström et al., 1988). Ginzburg (1969) described IR drying of grains, flour, vegetables, pasta, meat, fish, etc. and showed that IR drying can be successfully applied to foodstuffs. There are many current industrial applications of drying agricultural produce by IR. Sandu (1986) pointed out as advantages of IR drying in foods, the versatility of IR heating, simplicity of the required equipment, easy accommodation of the IR heating with convective, conductive, and microwave heating, fast transient response, and also significant energy savings. Experimental and theoretical works on IR drying, of opaque and semitransparent materials (silica sand, brick, brown coal, graphite suspensions and slurry of surplus activated sludge) have been performed by Hasatani et al. (1983, 1988).

The purpose of this chapter is to give a general review of IR drying with special reference to industrial applications. A detailed description of this process as applied to paper drying, can be found elsewhere

in this handbook (chapter by K.T. Ojala and M.J. Lampinen). Note that many direct as well as indirect dryers can be modified to accommodate IR heaters. Indeed, combined convective and IR dryers have been shown to be very attractive. Also, IR heating can be coupled effectively with vacuum operation to permit removal of evaporated moisture. IR heating may be applied continuously or intermittently (in space or time) to save energy and often to improve product quality.

18.2 BASIC PRINCIPLES

18.2.1 THEORY

Transmission of electromagnetic radiation does not need a medium for its propagation. The wavelength spectrum of the radiation depends on the nature and temperature of the heat source. Every body emits radiation due to its temperature level, which is called “thermal radiation” because it generates heat. The wavelength range of thermal radiation is 0.1–100 μm within the spectrum. IR radiation falls in this category and is conventionally classified as (Sandu, 1986): near IR (0.75–3.00 μm), medium IR (3.00–25 μm), and far IR (25–100 μm).

Thermal radiation incident upon a body may be absorbed and its energy converted into heat, reflected from the surface or transmitted through the material following the balance:

$$\rho + \alpha + \tau = 1 \quad (18.1)$$

where ρ is the reflectivity, α the absorptivity, and τ the transmissivity. For monochromatic incident radiation, these properties are called “spectral” and when that radiation is polychromatic they are defined as “total” (Sandu, 1986). Materials may be classified based on their transmissivity, depending on the physical state of the body where the radiation impinges. A body that does not allow the radiation to be transmitted through it is called “opaque” and is characterized by $\tau = 0$. Examples of these are most solids. On the other hand, liquids and some solids like rock salt or glass have a defined transmissivity so they are “transparent” to radiation.

The reflection may be “regular” (also termed specular) or “diffuse,” which depends on the surface finish of the material. In the former case, the angle of incidence of the radiation is equal to the angle of reflection due to highly polished surface or a smooth surface. When the surface has roughnesses larger than the wavelength, radiation is reflected diffusely in all directions.

Generally solid bodies absorb all of the radiation in a very narrow layer near the surface. This is a very important consideration in modeling the heat transfer process, since mathematically this concept

transforms a term within the energy balance into a boundary condition. An ideal body that absorbs all of the incident energy without reflecting or transmitting is called a “black body,” for which $\alpha = 1$.

The total amount of radiation emitted by a body per unit area and time is called “total emissive power E ” (Kreith, 1965) and depends on the temperature and the surface characteristics of the body. This energy is emitted from a surface in all directions and at all wavelengths. A black body is also defined as the one that emits the maximum radiation per unit area. The emissive power of a black body, E_b , depends only on its temperature. The emissivity of a body, ε , is then defined as the ratio of its total emissive power to that of a black body at the same temperature, $\varepsilon = E/E_b$.

As was pointed out earlier the total emissive power has energy from all the wavelengths in the spectrum of the radiation. On the other hand, the monochromatic emissive power, E_λ , is the radiant energy contained between wavelengths λ and $\lambda + d\lambda$ (Welyt et al., 1984). For a black body, this power is expressed by (Planck’s law of radiation)

$$E_{b,\lambda} = \frac{2\pi c^2 h \lambda^{-5}}{\exp\left(\frac{ch}{\lambda T}\right) - 1} \quad (18.2)$$

so the monochromatic emissivity of a body is defined as $\varepsilon_\lambda = E_\lambda/E_{b\lambda}$. Kirchhoff’s law states that under thermodynamic equilibrium (which requires all surfaces be at the same temperature), the monochromatic absorptivity and emissivity of a body are equal.

Equation 18.2 has a maximum that is related to the temperature by the following expression (Wien’s displacement law):

$$\lambda_{\max} T = 2897.6 \mu\text{K} \quad (18.3)$$

Equation 18.2 may be integrated over all wavelengths to obtain the total emissive power for a black body (Stefan–Boltzmann law):

$$E_b = \int_0^\infty E_{b,\lambda} d\lambda = \sigma T^4 \quad (18.4)$$

where σ is the Stefan–Boltzmann constant.

A gray body is defined as one which has the same emissivity over the entire wavelength spectrum. Thus, Kirchhoff’s law may be applied to gray bodies independently of their temperature.

Heat exchange by radiation between two black bodies at different temperatures may be obtained using the Stefan–Boltzmann law and is expressed by

$$Q_r = A_i F_{ij} \sigma (T_i^4 - T_j^4) \quad (18.5)$$

where F_{ij} is the shape or view factor between surfaces i and j . By definition, this geometrical factor takes into account the part of the total radiation emitted by the surface i that is intercepted by surface j . To calculate theoretically this factor is rather complicated but for most common geometries there are charts and formulas available in the literature (Welty et al., 1984; Kreith, 1965). A useful equation known, as the “reciprocity theorem,” relates the shape factors and the areas of both surfaces through the following equation:

$$A_i F_{ij} = A_j F_{ji} \quad (18.6)$$

If the exchange of energy by radiation is between N bodies, the shape factors must follow the relation:

$$\sum_{j=1}^N F_{ij} = 1 \quad (18.7)$$

In fact, very few bodies behave as black bodies so a more realistic assumption would be to treat those as gray bodies. The net radiation between two gray bodies is then given by the following equation:

$$Q_r = \frac{\sigma(T_i^4 - T_j^4)}{\left(\frac{\rho_i}{\varepsilon_i A_i} + \frac{1}{A_i F_{ij}} + \frac{\rho_j}{\varepsilon_j A_j}\right)} \quad (18.8)$$

Also, it must be noted that sometimes the electromagnetic radiation that impinge on a body may be attenuated inside the body by scattering along with absorption. Scattering takes into account that electromagnetic radiation may undergo a change in direction, which can result in a partial loss or gain of energy (Siegel and Howell, 1972, p. 420). Suppose that I_λ represents a spectral radiation impinging normally a layer of material where it is absorbed and scattered, so the intensity of monochromatic radiation is attenuated following the relationship (called Bouguer’s law, Siegel and Howell, 1972, p. 413):

$$I_\lambda(z) = I_\lambda(0) \exp \left[- \int_0^z K_\lambda(z^*) dz^* \right] \quad (18.9)$$

where K_λ is the extinction coefficient, z the distance, and $I_\lambda(0)$ is the radiation at the surface of the body. The extinction coefficient depends on temperature, pressure, composition, and the wavelength of the incident radiation. It may pointed out that Equation 18.9 is also termed as “Lambert’s law” or “Bouguer–Lambert law,” and “Beer’s law” when the extinction coefficient is put in mass terms, but it must not be confused with the “Lambert’s cosine law.”

18.2.2 RADIATION PROPERTIES OF MATERIALS

The design and modeling of any process always requires a profound knowledge of the materials. Specifically for IR drying, this fact may be the clue to accomplish a safe and efficient process, because radiation properties of both the radiator and the material to be dried must be matched in order to obtain most efficient results.

Emissivity, absorptivity, reflectivity, and transmissivity are the key radiation properties. The relative magnitudes of α , ρ , and τ depend not only on the material, its thickness, and its surface finish, but also on the wavelength of the radiation (Kreith, 1965). Nevertheless the emission of electromagnetic waves is a property of the material only.

Electrical conductors (e.g., metals) generally show an increase in emissivity ε with an increase of wavelength of the radiation. On the other hand, nonconductors such as asbestos, cork, wood, concrete, etc. show the opposite trend. The emissivity of many bodies also show directional properties but as the data available are scarce, a good approximation is to suppose an average value for $\varepsilon/\varepsilon_n = 1.2$ for polished metallic surfaces and 0.96 for nonmetallic surfaces (Kreith, 1965).

For practical purposes, only a mean value of the emissivity or absorptivity over the direction is required. Sieber (1941) obtained experimental data on total emissivity of opaque materials depending on the temperature of the source. Many authors (Ginzburg, 1969; Kreith, 1965) have reproduced these results graphically (Figure 18.1). The behaviour of electrical conductors and nonconductors with temperature of the radiator can be approximately interpreted from the dependency of the monochromatic emissivity on wavelength and the relationship between temperature of the radiator and the wavelength.

At radiation temperatures in the range from 227 to 620°C, the total reflectivity of polished pure silver is between 0.98 and 0.968 and for polished pure gold from 0.982 to 0.965. For polished aluminum, the reflectivity varies from 0.961 to 0.943 in a temperature range from 223 to 577°C (Welty et al., 1984). The high reflectivity of these materials is the reason why reflectors of radiation lamps are made of a thin layer of silver, and polished aluminum is used as a facing material for internal partitions in equipment for IR radiation (Ginzburg, 1969). It may be noted that for the construction of the equipment for IR drying and in selecting the reflectors for radiator lamps, opaque materials with high reflectivity are required.

The material to be dried by IR requires a low reflectivity in order to minimize the power required to heat it, and depending on the specific drying process, a high or medium absorptivity. When drying

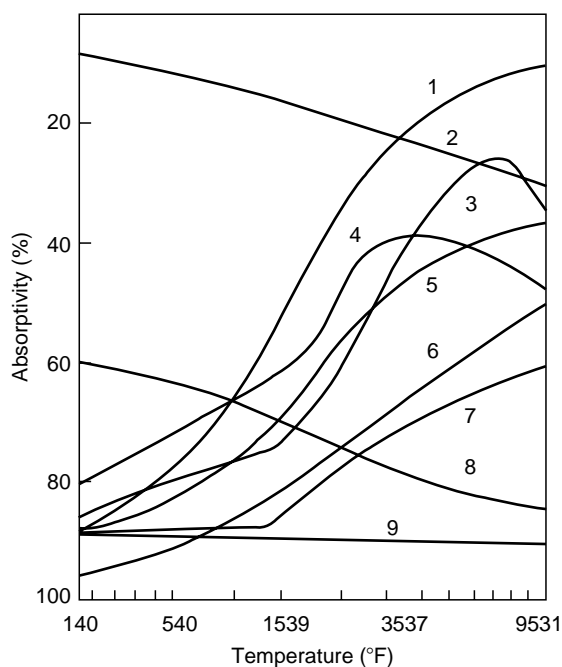


FIGURE 18.1 Absorptivity of some materials as a function of temperature (1: fireproof clay, white; 2: aluminium; 3: wood; 4: cork; 5: asbestos; 6: porcelain; 7: concrete; 8: graphite; 9: roofing silvers).

paints or coatings, a high absorptivity of the material is usually better, but in drying thick moist materials such as foodstuffs, it is preferable to use a material with high transmissivity to avoid extremely intensive heating and thermal damage of the surface. It is important to point out that if the absorptivity of a material is low, its transmissivity is high, and vice versa.

Properties like absorptivity and transmissivity of moist materials are not frequently encountered in the literature. In addition to the dependency with wavelength and thickness, they also depend on the water content. One of the most extensive reports on experimental data of these properties can be found in Ginzburg (1969). Mohsenin (1984) also presents a good compilation of data on radiation properties of agricultural and food products.

The variation of absorptivity or transmissivity of moist materials with wavelength is difficult to estimate without experimental data. For many materials, transmissivity is higher at lower wavelengths (Ginzburg, 1969). Foodstuffs, as an example, are complex mixtures of different large biochemical molecules and polymers, inorganic salts, and water (Sandu, 1986) and the absorption bands of each of these constituents are not the same. As an example, Figure 18.2 shows the IR absorption spectrum of liquid water. Generally, many fully wet materials have their minimum absorptivity at those wavelengths where water has its maximum transmissivity pointing out the important role that water plays in radiation absorption.

As drying proceeds, the material that is dried suffers a change in its radiation properties, increasing its reflectivity, and consequently lowering its absorptivity at low water contents. It is then possible to change adequately the temperature of the emitter in order to improve the absorption of radiation during drying.

The transmissivity decreases with an increase in layer thickness, whereas absorptivity increases. An approximate way of representing experimental transmissivity data as a function of thickness is presented by Ginzburg (1969). Table 18.1 and Table 18.2 show a compilation of experimental data from the literature (Kreith, 1965; Ginzburg, 1969; Sandu, 1986) about the variation of transmissivity of foodstuffs and other materials commonly dried, with thickness, water content, and wavelength.

In order to show an example of how both properties of the emitter and the product to be dried should be matched during IR drying, Figure 18.3 shows the radiant energy peaks for quartz tungsten filament at 2500 and 1925 K together with the spectral absorptivity of potato with 74.5% of water content (Sandu, 1986) and 10-mm thickness. To avoid overheating of the surface and to allow the radiation to penetrate into the product it would be better to choose the heat source at 2500 K because its maximum is located in the wavelength where the absorptivity of the foodstuff is not very high. On the other hand, if the lower

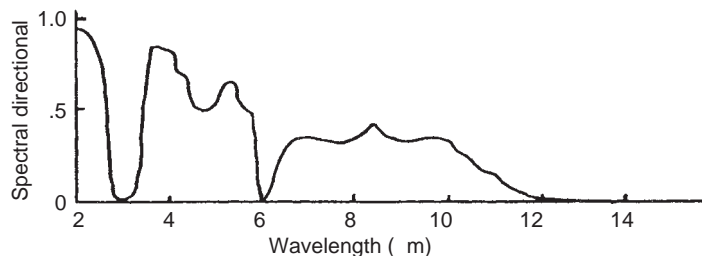


FIGURE 18.2 Absorption spectrum of water.

TABLE 18.1
Transmissivity of Selected Foodstuffs

Product	Spectral Peak, λ (μm)	T_r ($^{\circ}\text{C}$)	Thickness (mm)	W (%)	τ or τ_λ (%)
Bread made of wheat flour	—	400	2.00	—	1.04
			2.50		0.97
			3.50		0.46
			5.00		0.00
Dough made of wheat flour	—	Mirror lamp	1.00	44.0	5.70
			2.75		1.76
			5.00		0.39
			9.00		0.03
Tomato paste	1.075	—	0.50	60.0	95.0
	1.075		0.50	70.0	91.4
	1.075		0.50	85.0	55.8
	1.190		0.50	85.0	46.7
	1.350		0.50	85.0	38.5
	3.400		0.50	85.0	30.8
Potatoes	1.100	—	2.00	80.5	50.0
	1.100		8.00	80.5	16.0
Potato starch	1.100	—	1.00	11.8	13.0
	1.100		2.00	11.8	5.00
	1.100		8.00	11.8	0.23
Beet	1.100	—	2.00	85.5	40.0
	1.100		6.00	85.5	18.0
Fruit gel (marmalade)	1.100	—	2.00	30.0	18.0
	1.100		7.00	30.0	2.50

heat source temperature is used, the surface may be damaged (scorched) due to intense surface heating. Mohsenin (1984) presents an interesting discussion on the approach to be followed in order to obtain a successful IR drying of foods and agricultural products.

18.3 STEADY INFRARED DRYING

18.3.1 MODELING THE PROCESS

Only the energy balance equation must be modified to include IR in the calculation procedure for drying

TABLE 18.2
Transmissivity of Selected Materials

Product	Spectral Peak or λ (μm)	Thickness (mm)	τ or τ_λ (%)
Cigarette paper (raw)	1.075	—	73.0
	3.220		48.0
	6.150		20.0
Paper, thick (moist wall paper)	1.190	—	18.0
	3.750		10.0
	6.150		7.0
Wool cloth, dry	1.190	—	14.3
	3.220		12.3
	7.750		2.0
Sand	1.190	3.00	4.8
	6.150	3.00	2.0
Wood	1.190	1.50	4.3
	6.150	1.50	0.7

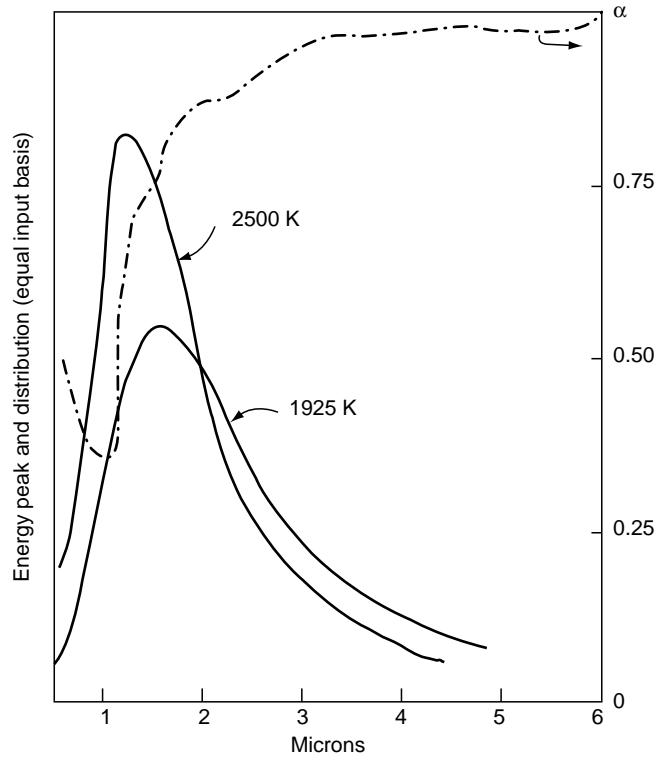


FIGURE 18.3 Energy peak and distribution (equal input basis) of quartz tungsten filaments together with the absorption bands of a slice (10 mm) of potato with 74.5% water content (---).

whereas the mass balance equation remains the same as that for conventional drying:

$$\frac{\partial W}{\partial t} = \frac{D_r}{Z^2} \frac{\partial^2 W}{\partial z^2} \quad (18.10)$$

The water flux is given in terms of (Crapiste et al., 1988)

$$n_w = -\frac{\rho_{s,o} X_o}{Z} D_r \frac{\partial W}{\partial z} \quad (18.11)$$

As the driving force for mass transfer depends on temperature, the high temperatures achieved by the drying surface with IR heating enhance the drying rate. In order to simplify the modeling of IR drying, it is convenient to start with IR drying of a flat partially “transparent” particle in which the heat conduction is one-dimensional. In the special case when the internal to external heat transfer resistance ratio is much greater than 0.1, the energy balance is given by

$$\rho_m C_{psh} \frac{\partial T}{\partial t} = k \frac{\partial^2 T}{\partial z^2} + \frac{\partial q_a}{\partial z} \quad (18.12)$$

with the following boundary and initial conditions.

$$\text{Initial condition: } t = 0, \forall z \quad T = T_o \quad (18.13)$$

$$\text{Boundary conditions: } t = t, z = 0 \text{ (center)}$$

$$\frac{\partial T}{\partial z} = 0 \quad (18.14)$$

$$t = t, z = Z \text{ (surface)}$$

$$kA \frac{\partial T}{\partial z} \Big|_{z=Z} = h_g A (T_g - T) \Big|_{z=Z} - n_w A \Delta H_s \quad (18.15)$$

The above model is commonly called “semitransparent” (Hasatani et al., 1983). The variable q may be obtained by averaging Equation 18.9 over all the wavelengths. If the average net radiation that impinges on a surface is Q_r (defined in Equation 18.8), q_a is given by

$$q_a = Q_r \exp \left[- \int_0^z \bar{K} dz^* \right] \quad (18.16)$$

If the particle is “opaque” to radiation, the energy equation becomes

$$\rho_m c_{psh} \frac{\partial T}{\partial t} = k \frac{\partial^2 T}{\partial z^2} \quad (18.17)$$

As was pointed out earlier in this chapter, an “opaque” solid absorbs radiation in a narrow zone near the surface rather than attenuating within its volume; so the second term on the right side of Equation 18.12 must appear in the boundary condition. Thus the boundary condition given by Equation 18.15 transforms to

$$kA \left. \frac{\partial T}{\partial z} \right|_{z=Z} = h_g A (T_g - T|_{z=Z}) - n_w A \Delta H_s + Q_r \quad (18.18)$$

whereas the other boundary condition and the initial condition remain the same. This model is called “opaque” (Hasatani et al., 1983).

In the case that only an average particle temperature is required or that the internal to external heat transfer resistance ratio is smaller than 0.1 (which means that the temperature profiles inside the particle can be considered to be almost uniform), the following simplification of Equation 18.17 is permissible:

$$\rho_m c_p \frac{\partial T}{\partial t} = k \frac{1}{\Delta z} \left(\frac{\partial T}{\partial z} \right) = \frac{1}{\Delta z} \left(k \frac{\partial T}{\partial z} \right) \Big|_{z=Z} \quad (18.19)$$

Inserting Equation 18.18 into the above equation, the energy equation becomes

$$m c_p \frac{dT}{dt} = h_g A (T_g - T) - n_w A \Delta H_s + Q_r \quad (18.20)$$

The model equations to represent the behavior of a dryer are an extension of the above explanation for IR drying of a single particle. In paper, paint, or textile drying, continuous dryers are commonly used. The equations describing continuous IR dryers can be found in the literature (Kuang et al., 1992; Cote et al., 1990). Nevertheless, a batch dryer may also be used for deep bed drying of foodstuffs because such materials have a solid structure from the beginning of the process (Heldman and Singh, 1981). The modeling equations for a batch dryer with circulation of air through the bed can be developed as follows.

Food particles experience shrinkage during drying, but since shrinkage is not appreciatively affected by air or particle temperature (Ratti, 1994) it may be assumed that IR drying does not affect shrinkage differently from conventional drying. The key assumptions employed in this model are

- (1) One-dimensional transport of heat and mass
- (2) Uniform velocity distribution in the dryer (plug flow of drying air)

- (3) Adiabatic system (well insulated)
- (4) Conduction heat transfer between particles in the bed and contact diffusion are negligible
- (5) Shrinking particles
- (6) The air is completely transparent to radiation.

Although in the drying process the air contains water vapor that absorbs IR radiation, this may be neglected because the amount of water vapor is small relative to the air. As an example, at a dry bulb temperature of 60°C and 40% of relative humidity, the absolute humidity is 0.053 kg water/kg dry air.

To avoid the problem of shrinkage, a moving coordinate system that follows the movement of the shrinking particles may be used, taking as basis a differential control volume that contains always the same amount of dry mass as in the initial time. This coordinate system is expressed as (Ratti, 1991)

$$d(z/L_o) = \frac{\rho_{s,o}(1 - \varepsilon_{lo})}{\rho_s(1 - \varepsilon_l)} d\Lambda \quad (18.21)$$

Then the resulting equations that represent batch fixed-bed IR drying are

Mass balance in the gas phase:

$$\left[\frac{\partial Y}{\partial t} \right]_{\Lambda} = \frac{n_w a_v (1 - \varepsilon_l)}{\rho_a \varepsilon_l} - \frac{1}{SL_o} \frac{G_s}{\rho_a \varepsilon_l} \frac{\rho_s (1 - \varepsilon_l)}{\rho_{s,o} (1 - \varepsilon_{lo})} \frac{\partial Y}{\partial \Lambda} \quad (18.22)$$

Mass balance in the solid:

$$\left[\frac{\partial X}{\partial t} \right]_{\Lambda} = - \frac{n_w a_v}{\rho_s} \quad (18.23)$$

Energy balance in the solid:

$$\left[\frac{\partial T_s}{\partial t} \right]_{\Lambda} = \frac{a_v}{\rho_s (1 + X) C_{psh}} [h_g (T_g - T_s) - n_w \Delta H_s + Q'_r] \quad (18.24)$$

Energy balance in the gas phase:

$$\left[\frac{\partial T_g}{\partial t} \right]_{\Lambda} = - \frac{h_g a_v (1 - \varepsilon_l)}{\rho_a \varepsilon_l C_{pah}} (T_g - T_s) - \frac{1}{SL_o} \frac{G_s}{\rho_a \varepsilon_l} \frac{\rho_s (1 - \varepsilon_l)}{\rho_{s,o} (1 - \varepsilon_{lo})} \frac{\partial T_g}{\partial \Lambda} \quad (18.25)$$

where Q'_r is defined from Equation 18.8 as

$$Q'_r = \frac{\sigma (T_e^4 - T_s^4)}{\frac{(1 - \varepsilon_s)}{\varepsilon_s} + \frac{1}{F_{se}} + \frac{(1 - \varepsilon_e)}{\varepsilon_e (A_e/A_s)}} \quad (18.26)$$

The initial profile of the four variables in the fixed-bed is stated as

$$\lambda = 0 \begin{cases} X = X_o \\ T_s = T_{so} \\ Y = Y_{go} \\ T_g = T_{go} \end{cases} \quad \lambda \neq 0 \begin{cases} X = X_o \\ T_s = T_{so} \\ Y = Y_{sat}(T_{so}) \\ T_g = T_{so} \end{cases} \quad (18.27)$$

and the values of T_g and Y_g at $\lambda = 0$ are set in T_{go} and Y_{go} for constant inlet conditions.

Some advances in the application of engineering to the modeling and design of dryers under radiation are made in the recent years. As an example, a complete model for a multiple-zone process for drying polymer-solvent coatings has been developed by Cairncross et al. (1995). Parrouffe (1992) has demonstrated on the basis of extensive experimental data that one may, within engineering accuracy, use analogy between heat and mass transfer to estimate the convective heat or mass transfer coefficients even in the presence of intense radiative heat flux on the evaporating surface. Appropriate corrections must be employed, however, for the high evaporative mass flux at the surface.

18.3.2 ADVANTAGES AND LIMITATIONS

Many authors have pointed out the advantages and disadvantages of using IR drying (van't Land, 1991, p. 251; Hallström et al., 1988, p. 218; Nonhebel and Moss, 1971, p. 286; Dostie et al., 1989). In fact IR drying has many positive attributes, the main one being the reduction in drying time. Figure 18.4 shows the effect of IR drying compared to conventional convective drying of

acoustic tiles (Dostie et al., 1989). Also, IR drying offers solution to problems that seemed to be unsolvable in the past such as those associated with the carrying of volatile organic compounds from solvent-based paints by the exhaust hot air in conventional convective dryers. A summary of the advantages of IR drying follows:

1. High efficiency to convert electrical energy into heat for electrical IR.
2. Radiation penetrates directly into the product without heating the surroundings.
3. Uniform heating of the product.
4. Easy to program and manipulate the heating cycle for different products and to be adapted to changing conditions.
5. Leveling of the moisture profiles in the product and low product deterioration.
6. Ease of control.
7. IR sources are inexpensive compared to dielectric and microwave sources; have a long service life and low maintenance.
8. Directional characteristics that allow to dry selected parts of large objects.
9. Occupies little space and may easily be adapted to previously installed conventional dryers.
10. Low-cost technology.

On the other hand, the disadvantages are:

1. Scaling up of the heaters is not always straightforward.
2. Essentially surface dryers. Nevertheless, a great effort is done to improve this technology in

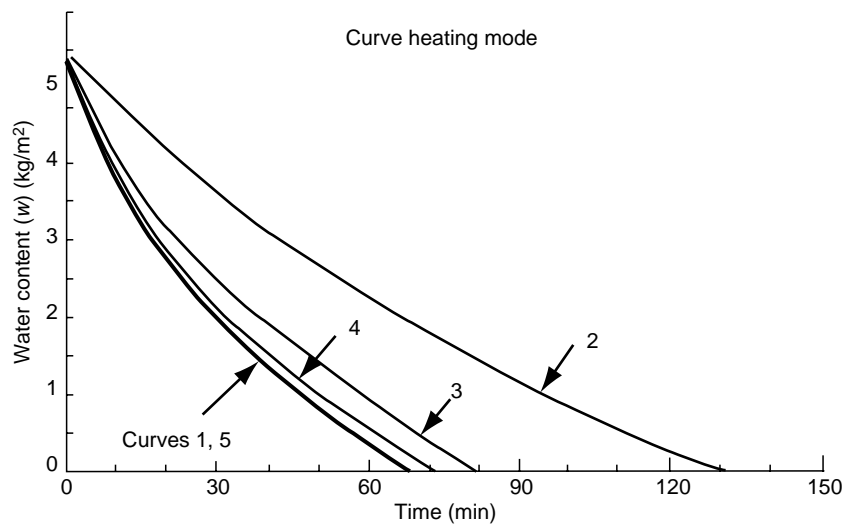


FIGURE 18.4 Drying times according to the heating process employed. (From Dostie, M., Séguin, J.-N., Maure, D., Ton-That, Q.-A., and Châtigny, R., in *Drying'89*, A.S. Mujumdar and M.A. Roques (Eds.), Hemisphere, New York, 1989. With permission.)

order to adapt it to drying thick materials with successful results.

3. The testing of the equipment must be carried out in the plant to assure a successful design.
4. Potential fire hazards must be considered in design and operation.

18.4 INDUSTRIAL INFRARED DRYING APPLICATIONS

18.4.1 APPLICATIONS OF INFRARED RADIATION IN INDUSTRY

IR heating is widely used in industry for surface drying or dehydration of thin sheets such as textiles, paper, films, paints, stove enamels, etc. Specifically in the automotive industry, the IR baking for paint-on-metal application is the most successful. Another sector where IR plays an important role is in pulp and paper industry. As an example, in Sweden this sector imposes higher demands on energy consumption than any other, and a relative new method that not only improves paper quality but also achieves energy-efficient drying is to use electrically operated IR radiation heating (Hannervall et al., 1992).

The disposal of hazardous waste is a less general IR application but, for instance, the IR drying of metal hydroxide sludge minimizes the cost impact of increasingly stringent disposal restrictions and when the metal content of sludge is of interest to recycle, it may be one way of changing a costly by-product into a profitable one (Davis and Wachter, 1992).

Although IR drying of thick porous materials has not yet been fully developed, some researchers showed, mainly by means of experimental results, that one of the applications where long wave IR heating is most efficient is in dehydration of foods (Ginzburg, 1969; Kimura et al., 1992). Recently, IR radiation has been investigated for drying of cashew kernels (Umesh-Hebbbar and Rastogi, 2001), herbs (Paakkonen et al., 1999), barley (Afzal et al., 1999), potato (Afzal and Abe, 1999; 1998; Masamura, 1998), shrimp (Fu and Lien, 1998), rough rice (Abe and Afzal, 1997), onion (Itoh, 1995), persimmons (Kim, 1993), and red pepper (Koh et al., 1990). In general, the previous articles deal with the application of IR heating to drying of different foods and the analysis of the characteristics and performance of these particular IR dryers. Japanese food industry uses this type of heating for drying of sea weed, curry sauce, carrots, and pumpkins among other things (Kimura et al., 1992). Other important IR applications in the food area (not specifically in drying), are cooking soybeans, cereal grains, cocoa beans and nuts, precooking rice, bacon, and barley grains for “ready to eat”

products, braising meat and frying. An automatic IR system for selective heating of foods, which could have a big impact in the sterilization and pasteurisation processes, has been recently analyzed (Irudayaraj and Jun, 2000).

18.4.2 INDUSTRIAL INFRARED DRYERS

Although IR dryers may be batch or continuous type, the latter is the most common arrangement. IR ovens for dryers are usually designed and constructed from standard IR sections arranged and integrated to the conventional dryers in such a way that IR radiation is directly intercepted by the product to be dried. These sections are selected on the basis of the particular application. It is desirable to test the product on a laboratory-scale IR oven under simulated conditions and to design the large-scale unit on the basis of the experimental data obtained. To accomplish a reliable design, it is also necessary to know the efficiency of conversion from electric to IR energy of the radiators used in the plant (unless gas-fired IR heaters are used). The main data required are the intensity of radiation and the residence time (Nonhebel and Moss, 1971, p. 289) but although oven style and cross section are easily determined; on the other hand, the selection of the heat source, time-temperature cycle, and the power density requires oven design experience (Fostoria, Technical Bulletin, 1992).

Air flow is required in IR ovens for two primary purposes:

- (a) Air movement to cool and protect oven walls and terminals
- (b) Oven exhaust to remove smoke, moisture, solvents, hazardous vapours, etc.

The decision of using natural or forced convection and the amount of air flow rate to meet the appropriate cooling effect must be adjusted to the specific application.

18.4.2.1 INFRARED SECTIONS

IR sections basically consist of a heat source (called radiator or emitter), a reflector, source sockets, electrical connections, and a shell where the parts of each section are built together (Figure 18.5). The main component is the radiator, which depending on the mode of heating, may be classified as:

1. *Electrically heated radiators*: In these radiators, the IR radiation is obtained by passing an electric current through a resistance, which raises its

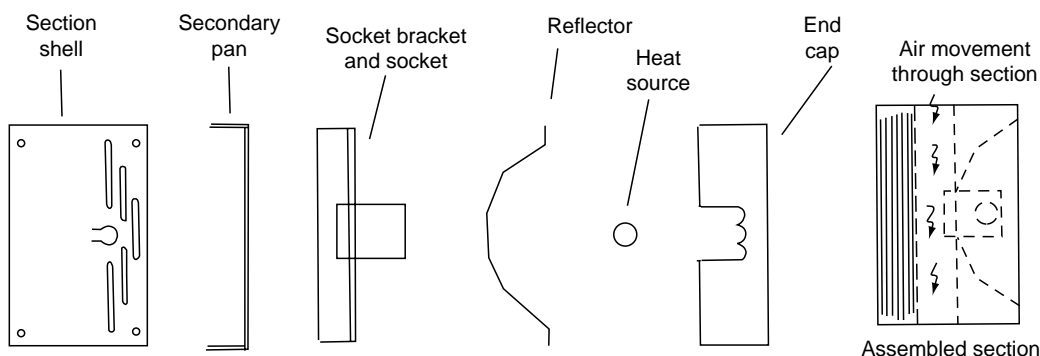


FIGURE 18.5 Parts of an IR section. (Courtesy of Fostoria Industries, Inc., Fostoria, OH. With permission.)

temperature (Hallström, 1988, p. 217). The most common are: metal sheath radiant rods, quartz tube, and quartz lamp. A typical cross section of a tube emitter is sketched in Figure 18.6a. One of the most important characteristics of such emitters is the radiant efficiency, which may be defined as the percentage of radiant output from a heat source referred to the energy input. There is a positive relationship between this efficiency and the temperature of the radiator. Also, as was pointed out previously, there exists an inverse relationship between this temperature and its

peak energy wavelength. The peak wavelength can be controlled by changing the temperature of the source so if different types of emitters operate at the same temperature they will all have the same peak wavelength as well as other characteristics like penetration and color sensitivity. Figure 18.7 presents the relationship between voltage and temperature of the radiator together with the efficiency curve and Figure 18.8 shows graphically the heat up and cool down rate of response of the more common emitters which can be an important criterion in the

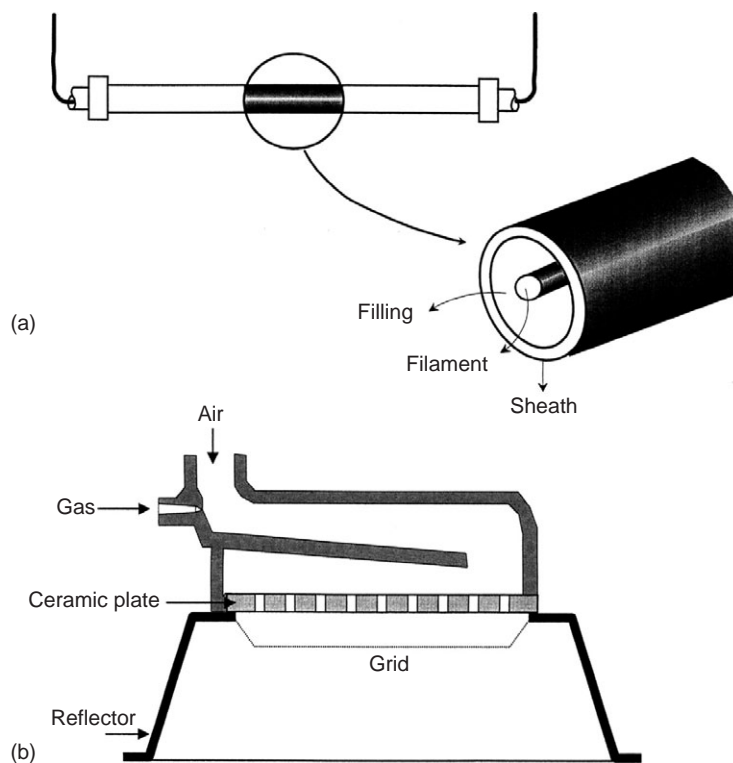


FIGURE 18.6 (a) Sketch of an electric IR source. (b) Sketch of a gas-fired IR source.

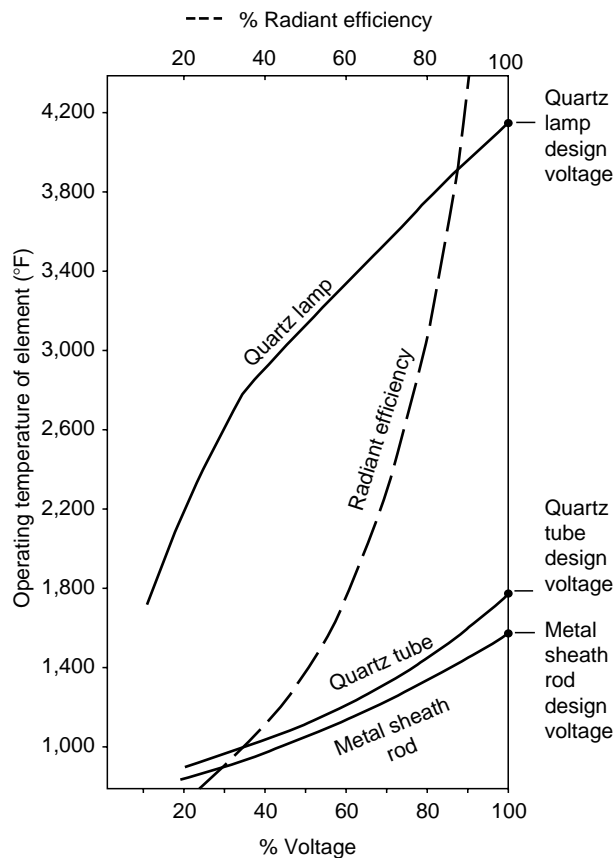


FIGURE 18.7 Radiant efficiency and relationship between voltage and temperature of various radiators. (Courtesy of Fostoria. With permission.)

selection of the proper source for a particular application. The main characteristics of the electric emitters are shown in Table 18.3.

2. *Gas-fired radiators:* These radiators consist of a perforated plate (metal or refractory), which is heated by gas flames in one of the surfaces so the plate raises its temperature and emits radiant energy. The porosity of the plate determines the temperature of the other surface so as to ensure a safe process. Figure 18.6b shows a sketch of this type of radiator (van't Land, 1991, p.250). The temperature of such a radiator is generally between 1500 to 1700°C with wavelengths from 2.7 to 2.3 μm (van't Land, 1991, p. 249). The radiant efficiency of such radiators is typically about 60%.

For practical purposes, choosing an emitter involves consideration of the following factors (Bischof, 1990):

1. Absorption characteristics of the material that is heated
2. Power density of the radiating area “seen” by the product
3. Ratio of convected heat to radiant heat
4. Nature of the installation
5. Type of control required

One of the most successful emitters is the quartz lamp because it ensures high power densities, maximum heat efficiency, flexible design parameters, and ease of controllability. Also, this type of emitter is fitted with a gold reflector to direct the radiation toward the product to be heated. Various reflector systems are also used (Hallström, 1988, p. 217):

- Individual metallic/gold reflectors
- Individual gilt twin quartz tube
- Flat metallic/ceramic cassette reflectors

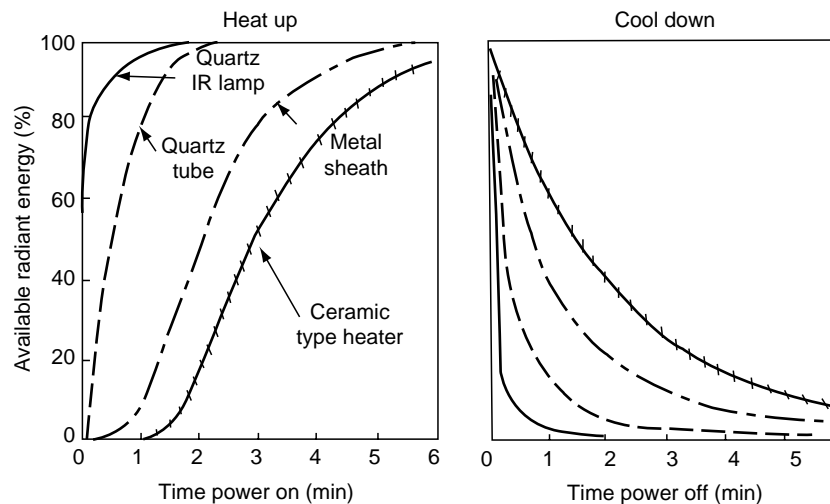


FIGURE 18.8 Heat up-cool down time cycles for IR sources. (Courtesy of Fostoria Industries, Inc., Fostoria, Ohio. With permission.)

TABLE 18.3
Properties of Electric Radiators

Type of Emitter	Metal Sheath	Quartz Tube	Quartz Lamp
Sheath material	Stainless steel	Translucent quartz	Clear quartz
Filament	Nickel-Chrome	Nickel-Chrome	Tungsten
Sheath diameter	3/8"	3/8", 5/8", 7/8"	3/8"
Filling material	Insulating powder	Air	Inert gas
Maximum filament temperature	1800°F	1800°F	4000°F
Peak wavelength	—	2.3 μm	1.15 μm
Voltage	240 V	up to 600 V	~ 600 V
Wattage	60 W per linear inch	30, 60 or 90 W p.l.i.	100 W p.l.i.
Radiant efficiency	50%	60%	86%

Figure 18.9 shows a sketch of such reflectors. The materials and the shape of the reflector determines its efficiency. Reflector materials must have high reflectivity, resist corrosion, heat and moisture, and be easily cleaned. They must also maintain the high reflectivity over a long period of time.

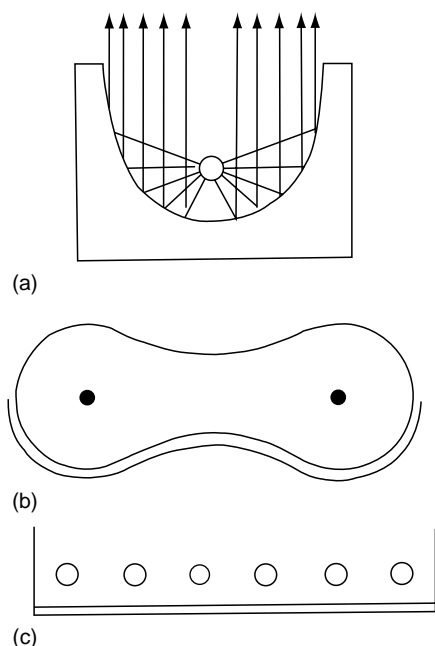


FIGURE 18.9 Different types of reflectors: (a) individual reflector; (b) individual gilt twin quartz tube; and (c) flat metallic-ceramic cassette reflector. (From Hallström, B., Skjöldebrand, C., and Trägårdh, C., *Heat Transfer and Food Products*, Elsevier Applied Science, London, 1988. With permission.)

18.4.2.2 RECENT DEVELOPMENTS IN DESIGN OF INFRARED DRYERS

Although an IR dryer can be built using an existing convective dryer by putting in the appropriate number of radiators over the product to be dried so as to direct the radiation on it, this technology is under improvement and new combinations of dryers have appeared in the market. As examples of these new trends and applications, Figure 18.10 and Figure 18.11 show two industrial IR dryers, one high velocity hot air impingement oven with IR electric heaters mounted between adjacent nozzles (Figure 18.10, Glenro, Technical Bulletin, 1992) specially for drying adhesives and inks on papers, foam and composite web substances, etc. and the other is a gas-heated IR dryer for metal hydroxide sludge volume reduction (Figure 18.11, JWI, Technical Bulletin, 1992). Another promising technology is the combination of intermittent IR radiation with continuous convection heating (Dostie et al., 1989) for drying thick porous materials such as panels made of wood and of acoustic tiles.

18.4.3 COSTS

The capital costs per kilowatt installed depending on the different heating modes are presented in Table 18.4 (Dostie, 1992). Specifically for IR drying, the radiators are generally the main cost of a dryer. Table 18.4 also presents an approximate relationship between the costs of different types of emitters. A lamp radiator has a life of 2,000 to 3,000 h whereas a sheathed element from 5,000 to 10,000 h (Nonhebel and Moss, 1971, p. 290). The replacement of the radiator elements is the main maintenance item. The figures in this table should be taken as guidelines rather than precise. With changes in technology,

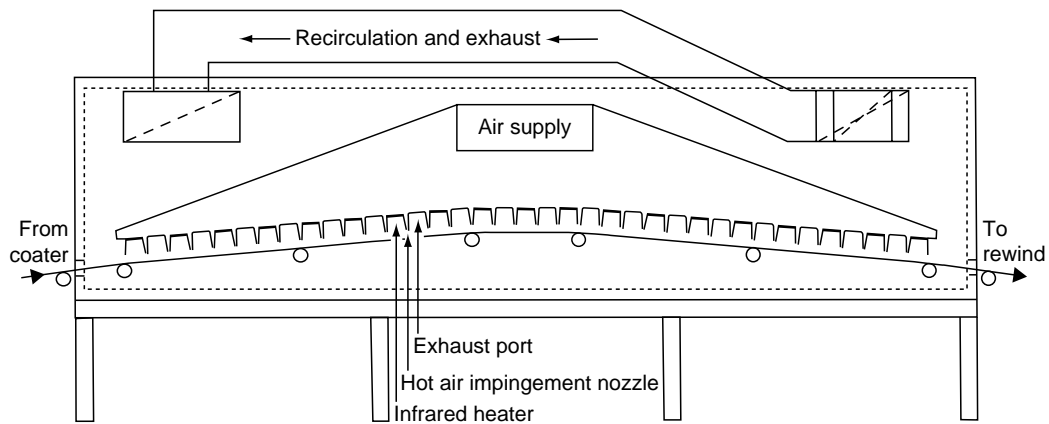


FIGURE 18.10 Impingement IR dryer. (Courtesy of Glenro, Inc., Paterson, New Jersey. With permission.)

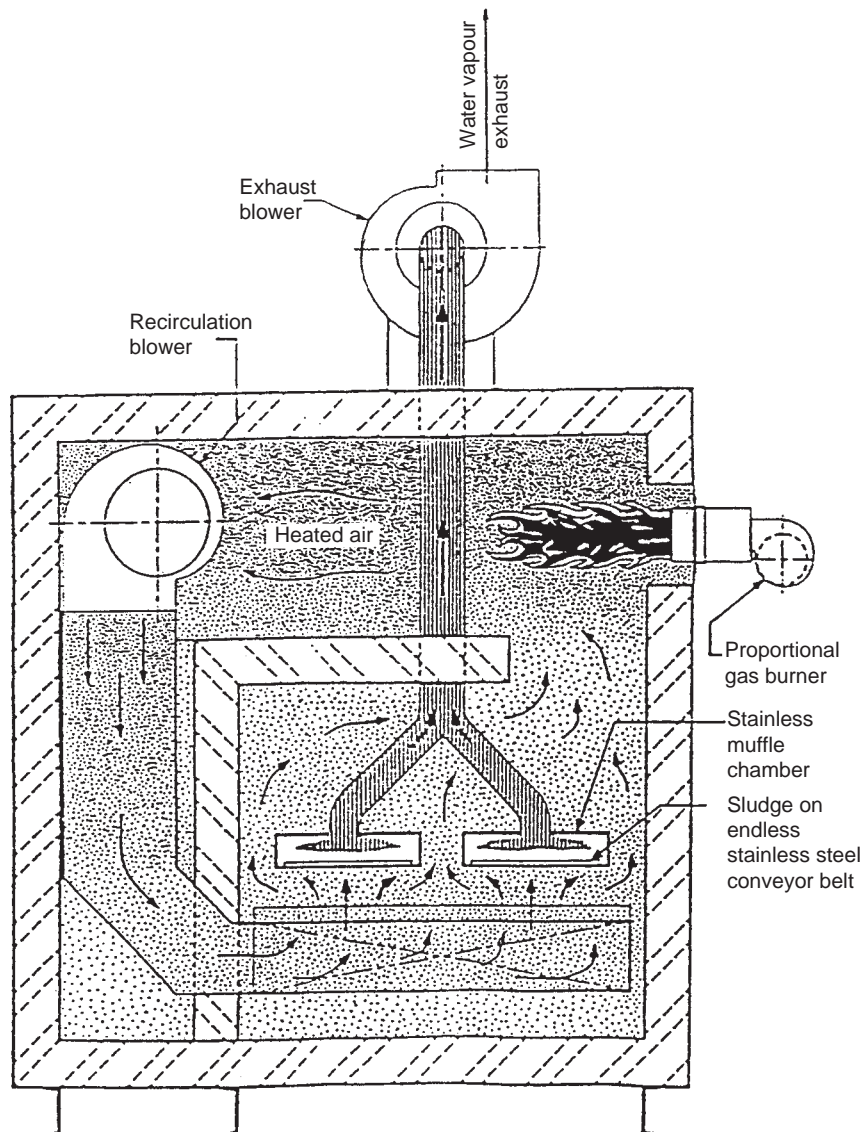


FIGURE 18.11 IR dryer for treatment of sludge. (Courtesy of JWI, Inc., Holland, Michigan. With permission.)

TABLE 18.4
Capital Costs for Different Modes of Heating and Types of Emitters (figures for Quebec, Canada, 1992)

Mode of Heating	Capital Cost (\$/kW)	Capital Cost (Equal Basis) \$/kW Referred to the Highest
Convection	300	—
Radio frequency	2000	—
IR	500	—
Electric lamp	—	0.49
Electric sheathed filament	—	1.00
Gas-fired emitter	—	0.93

these figures are likely to change with time. The costs are 1992 figures for Quebec.

18.5 CONCLUSIONS

In view of their several advantages, it is likely that IR drying in combination with convection or vacuum will become increasingly popular. Intermittent (spatial or time-wise) supply of IR heating has the potential merit of saving energy, reduce air consumption, and enhance the quality of heat-sensitive products. Dryers for continuous sheets or large surfaces (e.g., planks), utilizing a combination of impinging jets and radiant heaters spaced between jets, have already proven to be industrially viable. Also, combination of radiant heating under vacuum operation is a technically a sound process for drying certain products. Much fundamental and industrial R & D needs to be carried out to exploit fully the potential of IR drying technologies.

ACKNOWLEDGMENTS

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NOMENCLATURE

A	area
a_v	surface area/volume
C_{pah}	humid air specific heat
C_{psh}	humid solids specific heat
c	speed of the light
D_r	effective moisture diffusion coefficient
E	total emissive power
F	view factor
G_s	dry air flow rate
h	Planck constant

h_g	heat transfer coefficient
I	intensity of radiation
K	extinction coefficient
\bar{K}	average extinction coefficient
k	thermal conductivity
L_o	initial bed height
m	mass
n_w	water mass flux
Q_r	heat exchange between bodies
q_a	heat absorbed
S	dryer transversal section
T	temperature
t	time
W	dimensionless water content, X/X_o
X	solid water content (dry basis)
Y	air absolute humidity
z	distance
Z	particle half thickness

SUBSCRIPTS

b	black body
e	emitter
g	gas phase
i	surface
j	surface
n	in a direction normal to the surface
o	initial
r	radiator
s	solid
sat	at saturation
λ	monochromatic

GREEK SYMBOLS

α	absorptivity
ΔH_s	heat of sorption
ε	emissivity
ε_1	bed porosity
κ	Boltzmann constant
λ	wavelength

λ_{\max}	wavelength where E is maximum
ρ	reflectivity
ρ_a	air density
ρ_m	solid density
ρ_s	solid mass concentration
σ	Stefan–Boltzmann constant
τ	transmissivity

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19 Superheated Steam Drying

Arun S. Mujumdar

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19.1 INTRODUCTION

Although the concept was originally proposed over 100 years ago and the first industrial applications were reported some 60 years ago in Germany, superheated steam drying has emerged only in the past decade or so as a viable new technology with immense potential. Essentially, superheated steam drying (SSD) involves the use of superheated steam in a direct (convective) dryer in place of hot air, combustion, or flue gases as the drying medium to supply heat for drying and to carry off the evaporated moisture. Any direct or direct and indirect (e.g., combined convection and conduction) dryer can be operated as an SSD, in principle. The technology involved is more complex and hence this conversion is not simple. Additional criteria must be considered when selecting a dryer for SSD operation.

Currently there are fewer than ten major dryer manufacturers around the world that offer SSD technology on a commercial scale. Often, it is necessary to custom design an SSD for a new application, which may involve a new dryer type or an application to a new product. The author has presented an extensive review of principles, practice, industrial applications, potential new applications, market penetration

potential, and research and development (R&D) needs for SSD technologies [1]. Special consideration is given to potential applications for use of electricity and reuse of the exhaust steam from an SSD. Kumar and Mujumdar have provided an extensive bibliography and discussed the basic principles and applications of SSD [2].

One of the obvious advantages of SSD is that the dryer exhaust is also steam, albeit at lower specific enthalpy. In air drying, the latent heat in the exhausted steam is generally difficult and expensive to recover. Indeed, at current world prices of energy sources, it is often more expensive to recover energy in the exhaust steam than to waste it in the stack gases for most direct dryers with low-to-medium-temperature exhausts.

If air infiltration is avoided (or minimized to an acceptable level), it is possible to recover all of the latent heat supplied in the SSD from the exhaust by condensing the exhaust stream or by mechanical- or thermocompression to elevate its specific enthalpy for reuse in the dryer. As SSD will necessarily produce steam equal in amount to the water evaporated in the dryer, it is necessary to have a useful application for this excess steam in the process plant. If this steam is used elsewhere, the latent heat recovered is not charged to the SSD, leading to a net energy consumption figure of

the 1000–1500 kJ/kg water removed for SSD compared with 4000–6000 kJ/kg water removed in a corresponding hot-air dryer. Thus, reduced energy consumption is a clear advantage of SSD.

Other key advantages of SSD are as follows:

- No oxidation or combustion reactions are possible in SSD. This means no fire or explosion hazards and often also a better quality product.
- Higher drying rates are possible in both constant and falling rate periods, depending on the steam temperature. The higher thermal conductivity and heat capacity of superheated steam leads to higher drying rates for surface moisture above the so-called inversion temperature. Below the inversion temperature, drying in air is faster. In the falling rate, the higher product temperature in SSD (over 100°C at 1 bar) and lack of diffusional resistance to water vapor (no air) lead to faster drying rates. Also, it is known that many products that form “case-hardened skin” in rapid drying do not form such water-impermeable skins in SSD.
- For products containing toxic or expensive organic liquids that must be recovered, steam drying avoids the dangers of fire and explosion although allowing condensation of the off streams in relatively smaller condensers.
- SSD permits pasteurization, sterilization, and deodorization of food products.

Accompanying the advantages above are several limitations:

- The system is more complex. No leaks can be allowed. Feeding and discharge of SSDs must not allow infiltration of air. The product itself may bring in noncondensables. Start-up and shutdown are more complex operations for SSD than for an air dryer.
- Because feed enters at ambient temperature, there is inevitable condensation in the SSD before evaporation begins. This adds about 10–15% to the residence time in the dryer. At 1 bar operating pressure, the drying begins at a product temperature of 100°C in the constant rate period when surface water is removed.
- Products that may melt, undergo glass transitions, or be otherwise damaged at the saturation temperature of steam at the dryer operating pressure cannot clearly be dried in superheated steam even if they contain only surface moisture. Operation at reduced pressure, however, is a feasible option that may also enhance the drying rate.

- Products that may require oxidation reactions (e.g., browning of foods) to develop desired quality parameters cannot be dried in superheated steam. However, it may be possible to consider a two-stage drying process (e.g., steam drying followed by air drying). For drying of silk cocoons, for example, such a process appears to yield a higher quality product.
- If the steam produced in the dryer is not needed elsewhere in the process, the energy-related advantages of SSD do not exist. Also, steam cleaning may not always be a simple task. The chemical composition of the condensate must be carefully evaluated.
- Cost of the ancillaries (e.g., feeding systems, product-collection systems, exhaust steam recovery systems, etc.) is typically much more significant than the cost of the steam dryer alone. In most cases, SSD is a justifiable option only for very large tonnage, continuously operated systems because of the technoeconomics of the ancillary equipment needed.
- There is currently limited field experience with SSD for a smaller range of products. This database is expected to increase significantly in the coming decade. In the mean time, more pilot testing is recommended for an SSD application.

19.2 CLASSIFICATION AND SELECTION OF SUPERHEATED STEAM DRYERS

As noted above, any direct dryer, in principle, can be converted to superheated steam operation (e.g., flash, fluid bed, spray, impinging jet, conveyor dryers, etc.). Thermal efficiencies can be improved and the unit size reduced by supplying a part of the heat indirectly (e.g., by conduction or radiation). Note that the inversion temperature is lowered in the presence of indirect heat supply, which is a further benefit. For example, in the presence of appropriate radiant heating, the inversion temperature dropped from 250 to 170°C for a specific case reported in the literature.

However, it is not necessary to operate an SSD above the inversion temperature to benefit from the advantages of SSD. Aspects other than higher drying rates or lower energy consumption, such as quality or safe operation, may dominate the selection procedure in most cases. For relatively low-value products (e.g., sludges, coal, peat, hog fuel, etc.) that are also readily combustible in hot air and dried in large tonnages, the reduced net energy consumption in SSD is particularly advantageous as this also reduces the environmental emissions of greenhouse gases (e.g., CO₂) as well as toxic gases (NO_x, SO_x, etc.). However, the

excess steam produced must have a viable application in or near the process.

The following dryer types have been successfully tested at pilot scale and commercialized for at least some products:

- Flash dryers, with or without indirect heating of dryer walls with high-pressure steam
- Fluidized bed dryers, with or without immersed heat exchangers; operated at low, near-atmospheric, or high (up to 5 bar) pressures
- Spray dryers (operated at near-atmospheric pressures; for drying of whey; pilot scale only)
- Impinging jets (for newsprint, tissue paper, etc. at small scale; for textiles at commercial scale)
- Conveyor dryer, operated at near-atmospheric or high pressures
- Agitated bed dryers, operated at near-atmospheric pressure
- Packed bed and through circulation dryers
- Impinging stream (opposing jet concept) dryers
- Vibrated fluid bed dryers with immersed heat exchangers

Because of space limitations, this chapter discusses steam dryers only for a selected range of products. The references cited provide detailed information. Also, other chapters in this handbook include discussion of some steam-drying technologies (e.g., energy aspects, novel dryers, etc.).

19.2.1 DRYING OF SLUDGES

Sludges can be dried continuously in large tonnages in flash, fluid bed, or agitated trough-type dryers using superheated steam. Hirose and Hazama have reported on a sewage sludge treatment plant that used a steam-fluidized bed dryer [3]. The sludge, containing 400% water (dry basis), was mechanically dewatered and then evaporated to 75% before it was fed to the dryer to obtain a final moisture level of about 5%. The sludge had calorific values of 8.4–19 MJ/kg. A sludge with a calorific value of 12.6 MJ/kg required auxiliary energy of 420 kJ for per kg of sludge incinerated. The steam evaporated in the dryer was used as a heat source in the evaporator. The heat for the dryer was obtained by incinerating the dried sludge. The incineration process in a cyclone furnace vitrifies the sludge (by melting and solidification), thus reducing the volume of waste to be sent to a landfill. Further, the waste product is not leached into the soil.

An agitated multistage steam dryer with co-current flow of steam and the product has been developed successfully in Japan to dry 15 tons of

dewatered sludge per day. Steam at 360°C at flow rates up to 3600 kg/h enters the dryer and exits at 150°C, giving a volumetric heat transfer coefficient of up to 90 kcal/(m³ h °C). Unlike a fluid bed, the low-steam velocities cause little entrainment in the exhaust steam, which is cleaned in a cyclone. To avoid leakage of unpleasant odor, the dryer is held at a pressure of 10–100 mm water column below atmospheric pressure. For start-up, hot air is circulated and then water is injected into the hot air until the system is full of steam. It is important to note that sludges come in various chemical and biochemical compositions as well as physical characteristics. It is important to dewater (nonthermally) as much of the water as possible before the sludge is fed to a thermal dryer. Not all sludges can be combusted to provide a part or all of the heat required for drying.

There are advantages as well as limitations to each dryer type and a careful technoeconomic evaluation is necessary when a final selection is to be made. The well-known Carver–Greenfield process (discussed elsewhere in this handbook), utilizing the multiple-effect evaporation concept as well as several indirect dryers provides competing technologies for drying of sludges. It should be noted that use of high-temperature air can lead to significant fire and explosion hazards in drying of sludges of all types. Atmospheric emissions of organic volatiles and fine particulates are problems of special concern as well. For very large cities in developed countries, steam-drying technologies, which utilize dried sludge as fuel and yield vitrified, easy-to-handle dry product for landfills, appear to be especially attractive. Vitrified sludge particles do not leach their undesirable chemicals into the soil, which is a major advantage of such a process.

Although no details are available, an impinging stream dryer for sludges using super-heated steam as the drying medium has been operated at pilot scale successfully in Russia. Mujumdar has presented additional information on these special types of dryers [1] (see chapter on impinging stream drying in this handbook).

19.2.2 DRYING OF COAL

Coal is a raw material for many chemical syntheses as well as a fuel. Depending on its initial moisture content, coal is dried to increase its calorific value and simplify loading, unloading, transport, and improve boiler combustion efficiency. Although not commonly required in North America, drying of low-grade coals containing high levels of moisture is necessary in many parts of the world. Coal is also dried for briquetting, coking, gasification, carbonization, liquid fuel synthesis, and so on. Coke oven

efficiency can increase 30–50% in preheating and 10–15% in drying, if coal is predried. Direct dryers (e.g., rotary, pneumatic, fluid bed, vibrating fluid bed, shaft dryers, etc.) are used commonly with hot air or combustion gas at 700–900°C before the dryer and at 60–120°C after the dryer. Rotary dryers with indirect heating are used for hard coals. These have higher energy efficiency, about 3100 kJ/kg water evaporated. For air fluidized bed dryers, the corresponding figure is 3100–4000 kJ/kg water evaporated. A commercial vibratory dryer for hard and brown coals (manufactured by Escher-Wyss of Switzerland) uses a vibrational frequency of 50–100 Hz and an amplitude of 0.5–3 mm, giving a conveying velocity of 0.01–0.3 m/s with an angle of inclination of 5° to the horizontal. Low gas velocities are needed, as vibration suspends most of the pseudofluidized beds. The efficiency is better than in a conventional fluid bed employing high gas velocities. Attrition is reduced and gas-cleaning requirements are minimized in a vibrated bed dryer.

In pilot trials, Potter et al. have shown that extremely favorable heat transfer rates as well as drying efficiencies are obtained when drying brown coal in a steam-fluidized bed with internal heat exchanger tubes immersed within it [4,5]. Typical processing conditions were reported as follows:

Heating tube temperature	140–170°C
Bed temperature	110–127°C
Minimum fluidizing velocities	0.57 m/s (approximately)
Steam temperature	130–150°C
Steam velocity	0.20–0.30 m/s
Coal feed rate (wb)	40–70 kg/h
Product (wb)	16–28 kg/h

Using steam exhausted from one dryer stage as carrier steam for another stage, multiple-effect operation (similar to that common to evaporators) can be achieved yielding a steam economy of 1.9 for a triple-effect dryer. Potter et al. used a continuous fluid bed dryer for drying Victoria brown coal [5]. The fluid bed dryer was 0.3 m by 0.3 m by 3 m (high) with four bubble caps to distribute steam. The disengaging region was 2.5 m. Both horizontal and vertical tube bundles were tried.

Faber et al. have compared drying rates in air and in steam-fluidized beds of pulverized coal [6]. They confirmed existence of the inversion temperature above which steam drying is faster than air drying. Above about 180°C, the steam drying rate in the constant rate period in fluid bed drying exceeds that in (dry) air drying. For a 2000-kg/h dryer for alumina, they found the capital cost to be 20% lower for the steam dryer, whereas the total energy cost was 15%

lower. No credit was given to the steam produced in the steam dryer.

Faber et al. also report on a successful industrial installation using a steam dryer for activated carbon pellets (2000 kg/h, dry basis) from an initial moisture content of 50% to 2% (dry basis) [6]. The pellets are dried to 8% (dry basis) before they are fed to an evacuated chamber in which the final moisture content of 2% is achieved. The steam enters the dryer at 300°C and leaves at 150°C. The steam discharged is used to preheat the feed. The authors report smooth operation of the dryer since 1985. The installed cost of the steam drying steam was 40% lower than that for a conventional air dryer. The air dryer can operate at a maximum temperature of 125°C to avoid combustion in the dryer. The energy costs (1986 data) were estimated to be about \$3.6 per ton of dry product in South Africa.

Woods et al. have reviewed steam-drying technologies for coal and presented interesting results on steam drying of 1–13-mm coal particles and the evolution of volatiles during drying [7]. They note that, in steam drying, the drying time (actually residence time in dryer) does not affect the volatiles' liberation, unlike air drying. Further, they found that, under the conditions of their experiment, the constant rate drying period is 6–7 times longer in steam drying than in air drying and the heat transfer rate in steam drying is 1.7–2 times that in air drying. They also report favorable industrial experience with steam-fluidized bed drying of brown coal with an evaporative capacity of 25 t/h. No details are given about the use of steam produced by the dryer.

Black coal generally has low water content, whereas brown coal may have 60–80% (wet basis, wb) moisture. Regardless of whether brown coal is burned, gasified, coked, or liquefied, the wet raw coal should be dried to 5–10% moisture for economic utilization. For some gasifiers, the additional requirement of free flowability for a well-metered feed rate means that the pulverized coal must be 5% or below in moisture content. Coal drying in a steam-fluid bed of coal containing 65% moisture (wb) is estimated to reduce energy wastage in the subsequent combustion step by two thirds, resulting in some 15% increase in overall power plant efficiency, a similar level of reduction in CO₂ emission, and a 30% reduction in flue gas generation for the same thermal output.

In some reports, steam drying is claimed to reduce power consumption for milling as the grindability index is increased due to steam drying. At elevated degrees of superheat and prolonged exposure times, some studies have reported a reduction in the sulfur content of coal. This is not necessarily an advantage as the sulfur will then enter the dryer-exhaust steam. Additional data are needed to evaluate this aspect in detail.

Following are some of the key advantages claimed for steam-fluidized bed drying of brown coal:

- Better energy utilization by condensing steam generated in the dryer; overall efficiency up from 36%, typical for a brown-coal-fired power plant, to 42%.
- Ability to couple the dryer with power plants to use latent heat of condensed low-pressure steam (cogeneration potential).
- Coal moisture is discharged as liquid water rather than as dusty vapor. No biological or chemical treatment is needed for the condensate water.
- Large-capacity dryers (up to 15 t/h) are feasible.
- The dryer is compact (e.g., heat transfer coefficients in the order of 200 W/(m² K) versus 20–50 W/(m² K) in steam-tube rotary dryers).
- The operation is safer (reduced insurance costs).
- Product is dried to greater uniformity and better briquette strength.

The coal dryer itself is not the most expensive component of the drying system; coal grinding prior to drying, milling after drying, and cleaning the effluents are by far the most expensive steps in the process. Commercial-scale dryers for this application are operational in Australia and Germany.

19.2.3 DRYING OF BEET PULP

Drying of wet, pressed beet pulp after extraction of sugar in a conventional high-temperature dryer requires 9 kW h/100 kg beet, which amounts to over 33% of the energy consumption in a sugar factory. BMA AG of Germany has developed a high-pressure superheated steam dryer for beet-pulp drying that consumes only 2900 kJ/kg water evaporated, compared with nearly 5000 kJ/kg in conventional air drying. The BMA dryer is a horizontal pressure vessel made of mild steel and is 6.5 m in diameter and 37 m long [8]. Special airlocks are provided at entry and discharge of feed. The feed pulp is preheated to 105°C in a screw conveyor and then deposited on the drying screen—this avoids condensation and corrosion problems. Three traveling screens of stainless steel have a total area of 240 m² for a throughput of 32 t/h of pressed beet pulp with 30% solids to be dried to 90% solids. The pulp bed is between 40 and 120 mm, which affects the dryer throughput. The steam is made to circulate through the bed of beet pulp, causing it to dry as the screen moves. Nine parallel fans capable of handling 200,000 m³/h of steam circulate the steam in the dryer. Heat exchangers are used to reheat the recirculating steam. A mean

residence time of up to 720 s is needed for a bed of 120 mm with effective evaporation capacity of 16–20 t/h. The pulp is brighter than air-dried pulp. However, the dryer costs are high, with payback periods of up to 6–7 years.

Recently, Niro A/S of Denmark has successfully commercialized a pressurized-steam-fluid bed dryer for particulate and sludgelike or pulpy materials [9]. The fluid bed has a special “cellular” design and operates at 3 bar. The fluid bed is driven by superheated steam flow that is recycled through a heat exchanger by a fan and blown up through the fluid bed. Capacities of 2–40 t/h water evaporation are available. Initiated at pilot scale in 1982, the full-industrial-scale dryer (6-m diameter) for drying beet pulp came on stream in Denmark in 1985.

Compared with conventional rotary dryers, energy savings of up to 90% are feasible. The product quality is found to be better in steam drying (i.e., appearance and texture, as well as digestibility by cattle).

Although existing dryers using this concept are currently operational in Europe for drying of beet pulp, pilot tests indicate that this dryer can successfully dry the following products as well:

- Spent grain from brewery
- Grass, alfalfa
- Fish to produce fish meal
- Peels and pulp from citrus fruits
- Apple pomace
- Bark, wood chips
- Bagasse from sugarcane
- Municipal sewage sludge

Among some of the specific advantages claimed of this dryer are

- Closed-system drying eliminates odor emission
- Automatic, reliable, maintenance-free operation feasible
- No danger of fire or explosion
- Purer product; no contamination or oxidation in flue gases as in conventional dryer

It is interesting to compare the energy flows in the rotary dryer versus the pressurized-steam dryer for 32 t/h of water evaporation. If 16-bar-gauge steam is supplied to the steam dryer, it will generate 3-bar-gauge steam, which could have produced 2 MW. Instead, it is available as low-pressure steam, although the 2 MW is not used by the dryer. Rotary drying will consume 26 MW of fuel none of which can be recovered. Of course, all the advantages are gained at increased capital costs; as the cost of energy rises, the high-energy efficiency of steam drying becomes

economically more attractive. Potential for significant energy savings exists only when the steam produced by the dryer can be used effectively elsewhere in the process or at a nearby location.

19.2.4 DRYING OF LUMBER

Drying is an important step in processing lumber. Atmospheric steam dryers have been studied for drying green softwoods and hardwoods. Effects on resin exudation, surface discoloration, loosening of knots, and other quality parameters have been studied to varying extents for a variety of wood species. High-temperature drying of lumber (above 100°C) in air-steam mixtures causes greater defects, such as collapse, honeycomb, and checking in high-temperature-dried lumber than in lumber dried at lower temperatures.

Rosen has given pilot-scale data on pressure steam drying of lumber for 27-mm-thick, green, yellow poplar, and red oak [10]. He also made a technoeconomic study based on a number of plausible assumptions. Essentially, he showed that the capital cost of the steam dryer was about the same as that of a conventional kiln, although the drying times were reduced. The electrical energy usage was only about 0.82 kW h/kg (2.95 MJ/kg) water evaporated for the electric heater. Because steam flow rates used were higher than those in conventional kilns, the blower power was about 40% higher for steam dryers. However, as the dryer system is a closed one (in which dryer exhaust is recirculated after electrical heating), this energy is also useful for evaporation. It is important to note that the electrical energy requirements for drying depend on the type of lumber and the moisture range. Rosen dried yellow poplar from 100% moisture content to 5% (dry basis) in 28–30 h, whereas red oak was dried from 19% moisture content to 5% (dry basis) in 21 h. About 4% of the dollar value of the lumber was lost because of degradation associated with drying. The total energy requirements for red oak were estimated to be triple those for yellow poplar.

In the past decade, a low-pressure steam dryer for timber has gained momentum in Southeast Asia and Europe. The so-called Moldrup process marketed by Iwotech Limited of Denmark (1992) is carried out in an enclosed autoclave in which the sawn timber is stacked on trolleys for easy feeding and discharge [11]. The drying process is initiated by evacuation of the autoclave with a low-power vacuum pump, which takes 1–2 h. It is then filled with superheated steam, which is heated in the range of 50–90°C. Heat for drying is supplied by recirculating the steam at velocities up to 20 m/s at 4 millibar pressure using a series of fans within the chamber. The drying steam

is heated by steam or hot water coils located in the autoclave to a temperature above the saturation temperature of steam. At the operating pressure (around 3–5 millibar pressure) used, the boiling point of water is lowered to below 50°C. The process is controlled automatically.

Among the advantages claimed for the process are

- Fast drying rates (2–5 times faster)
- Simple process control
- No risk of fire or explosion
- Staining, mold attack avoided
- No oxidative discoloration (e.g., oak, beech, etc., do not discolor)
- Minimal danger of corrosion
- Minimal removal of extractables
- Reduced stresses, cracking, warpage, and the like

The autoclaves come in diameters up to 4 m and lengths up to 24 m. The autoclave is equipped with a rubber air tube that is pressurized to hold against the top part of the timber stack with a pressure of 1000 kg/m² to reduce danger of deformation. The moisture coming out of the timber is condensed. The drying time ranges from one to several days depending on the wood species and the thickness of the wood. At the end of the drying cycle, prior to unloading of the charge, the vessel is filled with atmospheric air.

The faster drying rates result in energy savings of about 50% over conventional kilns. One of the limitations of the process is that the autoclaves are much smaller than the conventional hot-air kilns. However, as the drying cycles are several-fold shorter, this is not a major limitation. In fact, the shorter drying times give the user flexibility in drying different species or sizes of wood while reducing the cost of inventory, especially for wood species requiring several weeks of drying time in air drying. Lack of oxygen in the dryer may also help kill microorganisms or insects in wood, although experimental verification is needed. Because the drying temperatures may be lower, thermal distribution may not be adequate in some instances. This process appears to be especially attractive for species and timber dimensions that take very long (weeks) to dry in conventional kilns.

19.2.5 DRYING OF PEAT

Peat is usually dried at power plants and briquette factories with flue gases (300–600°C) from the boiler. More recently, steam flash dryers have been commissioned in Sweden using back-pressure steam. The first industrial pulp dryer utilizing superheated steam as the carrier for conveying the pulp to be dried was

installed in 1979 at the Rockhammars in Sweden. Drying is accomplished by indirect heating with steam extracted from a turbine at 10 bar. Steam formed by evaporation within the dryer tubes is circulated in the dryer, where it acts as a convective drying medium. A part of the steam generated is used elsewhere in the mill as the dryer operates at 5 bar. Even after allowing for the energy loss in the turbine, it is found that this dryer consumes less than half of the energy used in flue gas drying—the conventional technology. Drying times are 10–30 s. This type of dryer has also been used successfully for drying of peat, sawdust, and forest biomass that, when pulverized, can be used as fuel in existing oil-fired thermal power plants.

According to Munter, flash dryers for pulp can be designed and operated economically at evaporation capacities of the order of 400 t/d [12]. If the steam produced is used elsewhere in the process, then the “net” energy consumption chargeable to the steam dryer can be very low. Currently, no full-scale industrial installation appears to exist that utilizes vapor recompression technology for the steam flash dryer similar to that commonly employed in multiple-effect evaporators. The higher capital costs associated with steam flash dryer units coupled with the lower energy consumption result in the cost of drying pulp by air or steam about the same, that is, of the order of U.S.\$23 per ton of water removed (in Quebec in 1989).

Possible operating problems for such dryers include deposit formation and erosion of the flash tubes due to high-velocity transport of particles that may contain abrasive impurities such as sand. Erosion, particularly in bends, can be minimized or taken care of by using special materials or replaceable elbows. Deposits may form within the dryer walls if the transporting steam temperature falls to saturation level. Corrosion is also a problem to be considered. Often, stainless steel construction is required. It is important to disintegrate the solids prior to feeding in the flash dryer as the transport steam does not suspend large lumps and the like.

Briquetting industries in Ireland and the former Soviet Union have used the so-called Peco dryer for peat for over seven decades [13]. The Peco dryer is a two-stage dryer. The latent heat of evaporated water from the second stage is used as the drying energy in the first stage. The dryer contains five columns, each containing about 500 pipes of about 70-mm diameter. The milled peat is carried by airflow or air–steam mixture in the pipes, while the energy for drying is supplied by condensing steam outside the pipes (second stage) or from hot water (first stage). Peat is separated by cyclones after each stage. The energy consumption for the Peco dryer is 1.7–1.8 MJ/kg

water evaporated, which is much lower than that for a flue gas dryer (3–4 MJ/kg water). Electrical power demand for the compressors used for pneumatic conveying is high.

Another dryer for peat (or lignite) is the tubular steam dryer generally used in Germany. It is a rotating inclined cylinder in which the material flows in 100-mm diameter pipes, whereas steam flows in the shells at a pressure of 0.60 MPa. Air is drawn through the pipes to remove evaporated water. The energy consumption is reported to be 2.9–3.2 MJ/kg water removed. The exhaust energy is generally not utilized. It is interesting to note that the flow properties of peat depend on the presence of wood and fiber content. If there is low content of fiber and wood with a bulk density of 400 kg/m³, such dryers operate well.

The pressurized-steam flash dryer originally developed at the Chalmers University of Technology, Gothenburg, Sweden, in the early 1970s is ideal for drying peat as well as pulp, bark, and so on. This dryer is a closed, pressurized system in which the peat is exposed to indirectly heated superheated steam. The dryer consists of transport ducts, heat exchangers, a cyclone, and flans. The superheated steam recirculates at a pressure of 2–6 bar. The primary heating steam is condensed (usually 8–15 bar) on the shell side.

Dry steam and material are separated in a cyclone and the steam is recirculated, whereas the excess steam is bled off. If the material is dried from 45–50% dryness to 85–90% dryness, about 1 ton of steam is generated per ton of material processed. This steam is available as process steam at 2–6 bar pressure. Typical drying time is 10–30 s. The transport velocities are 20–40 m/s. If excess steam is used (e.g., for district heating), then the net dryer energy consumption is only 0.5–0.7 MJ/kg water removed, which may be compared with 3.5–4.8 MJ/kg common to flash dryers using air. Dust and explosion hazards are eliminated in steam drying, thus reducing the insurance costs as well.

A recent installation in Sweden uses peat (replacing coal) as the fuel in a 440 MW thermal power plant. The peat is dried, briquetted (3 million tons/year) at source about 400 km from the power plant, and then milled for combustion. Peat is dried from 60% to 10% (dry basis) moisture for briquetting at an output level of 20 t/h per drying unit. Two dryers are used in parallel. The total tubular heat exchanger area is 5500 m². The electricity demand of the compressors and blowers is 10 MW. About 3.6 MW district heat is produced, giving a net energy demand of the dryer of about 0.50 MJ/kg water, which is only about one sixth to one seventh the energy consumption of typical flue gas drying systems.

At Helsinki University of Technology, successful demonstration of fluidized bed drying of peat has been reported at the pilot-plant level (100 kg/h of water evaporation). A tubular heat exchanger immersed in the bed provides indirect heat for drying. It uses condensing steam at 0.8 MPa. Steam evaporated from peat is used as the fluidizing medium. Wet peat is fed into the lower zone of the fluid bed while dry peat is withdrawn from the upper zone by a screw. The milled peat is about 1 mm in average size. The bed-to-tube heat transfer coefficient is rather low at 100 W/m² K (i.e., of the same order as that in a flash dryer but very low for a fluidized bed due to the poor thermal properties of peat). Sand may be used as a bed material to increase the drying rate but separation of peat from sand is not very effective. If a fluidized bed combustor with sand is used, then this is not a problem. Fine-size magnetite as a bed material increases the bed-to-tube heat transfer coefficient to 350 W/m² at the same time allowing easy magnetic separation from dry peat [14]. This dryer is claimed to be economically justified for use on new, large power plants. The investment costs are high and a demonstration plant is necessary to study the reliability and technoeconomics of the process. The fluid bed installation is likely to be more compact than the flash dryer but with net energy consumption of the same order.

Aside from differences in the flow characteristics, similar conclusions should apply for drying of lignite, biomass, and the like in steam flash or steam-fluid bed dryers. The cost and net energy consumption should be of the same order for the same production capacity (i.e., in the order of 0.5–0.7 MJ/kg water evaporated), provided the steam generated by the dryer is used elsewhere.

One important consideration in drying of peat is the evolution of organic compounds when flue gases are discharged into the stack or if the energy is recovered and the condensate sent as a waste stream. Indeed, such a problem exists whenever the material that is dried can result in evolution of organic compounds due to heating or due to interaction with steam at elevated temperatures. The quality of the condensate is affected by drying conditions (e.g., time, residence time in dryer, etc.), by the amount of moisture removed, type of peat, and so on. Acetic acid, formic acid, and furfurals are the main organic compounds in the condensate. The average biochemical oxygen demand (BOD) is 140–150 mg/kg dry peat, chemical oxygen demand (COD) is 500–850 mg/kg dry peat, and the total organic carbon (TOC) is 90–300 mg/kg dry peat in different dryers. Table 19.1 is a summary of the effluents from steam peat dryers based on the Fagernäs and Wilen data [13]. Note that drying of bark in steam-fluidized beds

TABLE 19.1
Effluents from Steam Peat Dryers

	Fluid Bed Steam Dryer	Pilot Fluidized Bed Dryer	Industrial Peat Dryer
Temperature, °C	100–140	140–170	170
Pressure, bar	—	2–8	5.7
pH of condensate	4.1	5.7	3.8
Solids, mg/l	80	220–400	70–170
NH ₃ -N, mg/l	1.2	28.0	11.0
P, mg/l	0.04	1.60	0.07
BOD, mg/l	520	130–190	—
COD, mg/l	880	470–630	440–1300
TOC, mg/l	310	90	310–450

BOD, biochemical oxygen demand; COD, chemical oxygen demand; TOC, total organic carbon; P, phosphorus

yields higher solids content (~400 mg/l), higher COD (~2600 mg/l), and higher TOC (~450 mg/l) than corresponding figures for peat; other effluent levels are comparable.

In general, increased bed temperature leads to increased load of organics in the condensates. Increasing bed temperature from 110°C to 120–130°C can increase BOD, COD, and TOC values nearly threefold. The longer residence time in a fluidized bed as compared with a flash dryer is likely to result in higher organic loading of the condensate. Finally, a novel high-pressure steam flash dryer (25 bar) has been operated successfully in Finland for drying of milled peat, which is then fed continuously to a high-pressure gasifier. Clearly, such a sophisticated dryer is not warranted for a drying application.

19.2.6 DRYING OF PAPER AND TISSUE

No commercial dryers exist for direct drying of paper webs with superheated steam. For the first time, Mujumdar proposed a variety of possible configurations for direct and direct and indirect drying of paper using superheated steam [15]. These include

1. Pure impingement with steam jets with web supported on a heated roll
2. Impingement and through drying (so-called Papridryer configuration but using steam rather than hot combustion gases, which pose serious fire hazards)
3. Pure through drying with steam drawn under suction with web (assumed permeable at reasonable pressure drop) supported on a perforated roll or honeycomb roll

All of the above processes have been tested at pilot scale or are already in commercial operation using hot air or hot combustion gases. All of these are highly energy-intensive processes requiring 4800–5700 kJ/kg water evaporated. Also, the electrical power consumption is high for the fan power that is needed for jets, throughflow, as well as the high recirculation ratios (up to 90% in some high-performance impingement dryers). Such dryers are in use for drying of highly permeable grades such as tissue and toweling. The Papridryer process, which is of above-mentioned second type, has been tested at pilot scale for newsprint [16,17]. The author has discussed the superheated steam drying of paper in some detail including its history, current status, and potential [18].

Aside from the higher drying rates and lower net energy consumption in a superheated steam dryer, two additional major advantages of interest to the papermaker are the elimination of fire hazard and improvement in product quality as measured by its strength and optical properties. Cui et al. [19] and later David [20] confirmed this in laboratory-scale static drying tests. Cui and Mujumdar [21], and Loo and Mujumdar [22] employed simple models to calculate the drying rates (under several assumptions later validated experimentally) and energy consumption for different steam-drying configurations. They concluded that if the steam produced in dryers is fully utilized elsewhere (without vapor recompression), the net energy consumption for paper drying can be as low as 1500 kJ/kg water evaporated. This energy requirement is mainly to account for leakage, sensible heating of the web, other losses, and energy for steam recycle. Pilot and mill tests are needed before the results can be confirmed.

Bond [23] and Poirier [24] have presented valuable data and analyses of steam drying of handsheets in static laboratory test rigs. No pilot-scale data exist at this time on steam drying of paper or tissue. Application to tissue drying (impingement and through drying) appears to be the more imminent industrial-scale operation. Another potential near-term application is as a booster dryer to increase the productivity of a dryer-limited mill.

19.2.7 DRYING OF WOOD PARTICLES AND WOOD WAFERS

Salin has reported on a significant study on steam drying of small wood particles for particleboard to very low moisture levels (2–3% water) [25,26]. The conventional process uses direct contact flue gas drying or indirectly heated air drying; both processes do not permit economic energy recovery from dryer exhaust. Wood particles were dried in a

pressurized-steam dryer in which the particle–steam mixture is conveyed through a series of vertical heat exchangers in which the mixture is heated to maintain sufficient superheat. Upon exhaust, the particles are separated from steam in cyclones and recirculated, except for an amount equivalent to that produced by the dryer. A 13-m long single-tube dryer was used in the pilot tests, giving a residence time of 12–15 s; a second pass was needed to achieve the required dryness in 25–30 s. The final desired dryness of 2% was achieved for both softwood and hardwood (birch) particles. About 40°C superheat was found to be adequate in the dryer. Fouling of heat exchanger surfaces was minimal, although this should be examined closely for industrial-scale operations. For bark drying in a similar dryer, no problems have been reported on heat exchanger surface encrustation even after six years of operation in Sweden.

Salin compared the quality of particleboards manufactured from steam-dried versus air-dried wood particles [26]; the bending and tensile strength, and water absorption properties of the former were found to be superior. This results in a 9% reduction in glue consumption in particleboard making. Optimal selection of dryer temperature and heat treatment can further enhance the board quality. No industrial-scale steam dryers appear to be in operation anywhere at this time for this application.

Figure 19.1 classifies superheated steam dryers based on their operating pressure. The product temperature necessarily exceeds the saturation temperature of steam at the corresponding operating pressure. So, for products that may undergo undesirable physical transformations (e.g., melting) or chemical transformations (e.g., hydrolysis) at elevated temperatures, a low-pressure operation is desirable.

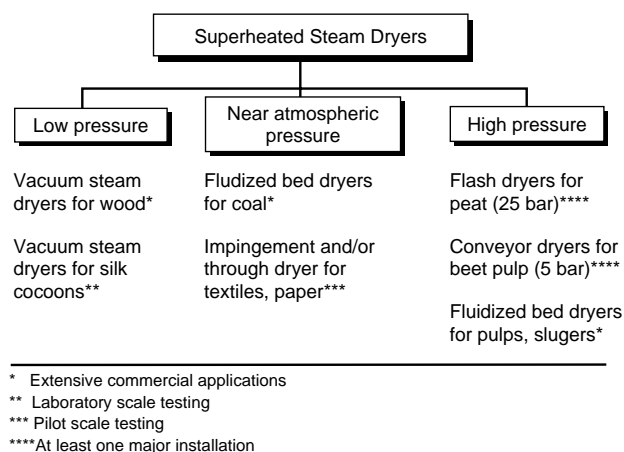


FIGURE 19.1 Classification of superheated steam dryers based on their operating pressure.

The vapor evolving from the product may be withdrawn from the chamber, condensed, and the latent heat recovered. Alternatively, the vapor is reheated within the drying chamber by tubular or plate heat exchangers and recirculated as a convective drying medium to enhance the drying rate. Such a system is used commercially, with very attractive results, to dry timber. Although the net energy consumption is reduced several fold, the product quality is enhanced and the environmental problem of emission of the volatile organic compounds (VOCs) produced during drying is also eliminated. The volatile organic components (boiling point ranging from room temperature to over 200°C) are condensed out with the steam. Because the two are immiscible, simple decantation allows recovery of the condensables (e.g., terpenes, essential oils), which may be sold separately.

In general, superheated steam drying is worth considering as a viable option only if one or more of the following conditions apply:

- Energy cost is very high; product value low or negligible (e.g., commodities like coal, peat, newsprint, tissue paper, waste sludges, which must be dried to meet regulatory requirements).
- Product quality is much superior if dried in steam rather than air (e.g., newsprint, which yields superior strength properties in steam and permits use of lower chemical pulp content to attain same strength and runnability).
- Risk of fire, explosion, or other oxidative damage is very high (e.g., coal, peat, pulps). Lower insurance premiums may partially offset added investment costs of a steam dryer.
- Quantity of water to be removed as well as production capacity required is high. This affords economy of scale. Clearly, such dryers are worth considering only for a continuous operation because of the inherent problems associated with start-up and shutdown when water condensation on the product as well as presence of noncondensables (air) cause problems.

19.2.8 QUALITY CONSIDERATIONS

It is impossible to generalize the effect of steam drying on product quality. Table 19.2 gives a summary of tests carried out at McGill University on batch drying of handsheets in a static apparatus in which the samples were dried under an impingement flow of superheated steam. Note that the effects are all positive relative to air drying or, at most, comparable. In Finland, similar tests on paper dried on a dynamic pilot-scale apparatus have yielded comparable results.

TABLE 19.2
Summary of Tests Carried Out at McGill University

Property	Effect of Steam Drying (Relative to Air Drying)
Bulk	Slight increase
Surface roughness	Marginal increase
Burst	20% enhancement (typical)
Tensile index	20–30% improvement
Stretch	No significant effect
Folding endurance	No noticeable difference
Scott internal bond strength	10–20% higher
Wet strength	Little effect
Permeability	Little effect
Crystallinity	No measurable change
Brightness	Above 300°C, a little worse
Fiber strength (zero span)	Little change

Extensive results have been reported in the literature on drying of wood as well. Again, all quality indicators are positive. Several vendors worldwide have already commercialized the vacuum-superheated steam-drying systems. It is noteworthy that besides enhancement of the product quality, the drying times are reduced two- to five-fold resulting in significant reduction in cost of inventory in the dryer.

For drying of silk cocoons, laboratory-scale testing in the People's Republic of China has proven the enhancement in quality of the silk produced (e.g., brightness, strength of fiber, etc.) in steam drying at a temperature of around 45°C. For certain food or vegetable products, the yielded porous structure of the product dried in superheated steam (due to evolution of steam within the product that enhances porosity) is a desirable characteristic. This decreases the bulk density of the product while enhancing its rehydration characteristics. The color and texture of the product may also become more desirable. Thus, the increased cost of steam drying may be offset by the additional credit received for the better quality of the dried product.

As noted above, any direct dryer, in principle, can be converted to a superheated steam dryer (e.g., flash, fluidized bed, spray, impinging jet, conveyor dryers). Thermal efficiencies can be improved and the unit size reduced by supplying a part of the heat indirectly (e.g., by conduction or radiation). Note that the inversion temperature is lowered in the presence of indirect heat supply, which is a further benefit. For example, in the presence of an appropriate radiant heating, the “inversion temperature” dropped from 250°C to 170°C for a specific case reported in the literature.

It is not necessary, however, to always operate an SSD above the inversion temperature to benefit from

TABLE 19.3
Factors Contributing to the Feasibility of Superheated Steam Drying

Factor	Description of Impact
<i>Product-Related Factors</i>	
Low temperature sensitivity	To avoid higher steam temperature, more expensive vacuum systems are required
High moisture content	As the latent heat of the moisture can be recovered with SSD, the bigger the portion it represents, the greater will be the efficiency improvement
High thermal resistance	Higher surface temperature with SSD reduces heating and drying times
High sensitivity to oxidation	Lack of oxygen with SSD improves product quality
Undesirable taste or aroma	SSD strips more of the acids which contribute to bitter tastes
High product values	Drying time reductions provide biggest inventory cost savings with higher value products
<i>Process-Related Factors</i>	
Other uses of steam available	Energy consumption for the process is small and the capital costs are minimized
Environmental emissions from dryers	SSD provides easier recovery of solvents and particulates
Combustion and explosion hazards	Lack of oxygen significantly reduces fire and explosion hazards
Expensive source of thermal energy	Thermal energy savings with SSD will offset greater the energy costs than with waste fuels (e.g., hog fuel)

the advantages of SSD. Aspects other than higher drying rates or lower energy consumption, such as quality or safe operation, may dominate the selection procedure in most cases. For relatively low-value products (e.g., sludges, coal, peat, hog fuel) that are readily combustible in hot air and dried in large tonnages, the reduced net energy consumption in SSD is particularly advantageous, as this also reduces the environmental emissions of greenhouse gases (e.g., CO₂) as well as toxic gases (NO_x, SO_x). However, the excess steam produced must have a viable application in or near the process.

Table 19.3 summarizes the key factors contributing to the feasibility of superheated steam drying. Both product- and process-related factors must be considered. Laboratory testing is required before proceeding with any decision to consider superheated steam drying.

19.3 MISCELLANEOUS APPLICATIONS

Furukawa et al. reported on a commercial steam drying and deodorization plant to process soy sauce cakes as feedstock [27]. An agitated trough steam dryer (1.5 t/h dry product capacity) was used. Deodorization commenced at a moisture level below 4% (dry basis) and a temperature of 135°C. The oils and fats distilling from the cake result in deodorization. Indeed, the fats and oils were used as auxiliary fuels for the process. Concurrent steam drying and deodorization of rice bran, fish meal, and pupa were also reported by Akao [28]. In these products, the

odor is due to oxidation of lipids. Akao also showed that the quality of green tea and vegetables dried in steam (e.g., color, odor, etc.) was acceptable.

Takahashi et al. dried a slurry of fine silica (0.1–5 μm) in a steam-fluidized bed using inert particles as bed material. The quality of the dried silica particles was as good as that obtained commercially in air drying. Also, it was noted that food powders can be sterilized in superheated steam conveyors. Among other applications of steam drying developed in Japan, one may cite drying of potato slices in which color and vitamin C are preserved due to lack of oxygen. Lack of formation of a case-hardened skin on the product in solvent drying was applied to advantage in dry spinning of synthetic fibers in superheated solvent vapor. Yoshida and Hyodo reported that stronger and finer fibers without surface wrinkles were obtained in this process [29].

19.4 UTILIZATION OF EXHAUST STEAM

Evaporated moisture in steam dryers, assuming no losses, becomes excess superheated steam at lower specific enthalpy. Economic use of this steam generally holds the key to the success of the steam-drying process. This steam is typically at near-atmospheric pressure and is likely to be dusty (e.g., in paper drying it will pick up fibers and dust). It must be cleaned for reuse. In high-temperature drying of tissue paper, for example, the fibrous debris picked up by the drying gases gets incinerated prior to recycling. This cannot occur in steam-drying systems. The degree of cleaning

required (using bag filters, cyclones, or electrostatic precipitators, etc.) depends on how the steam is recycled or utilized in the process. It is important to ensure that no condensation occurs in the vapor-cleaning systems. Following is a brief discussion of potential systems for steam utilization for dryer exhausts. A detailed discussion is beyond the scope of this chapter. The significance of this aspect, however, cannot be overemphasized.

Vapor recompression technology is widely used in continuous evaporators. Billet has discussed this subject in considerable detail [30]. In the pulp and paper industry, there are currently two major application areas in which mechanical vapor compressors are employed to achieve significant energy savings. One is use of compressors in conjunction with the black-liquor evaporators, which results either in a decrease in the number of evaporators or in increased evaporator efficiency. The compressor is used to compress the vapor leaving the evaporator and, under appropriate conditions, the mechanical energy (or electrical power) required by the compressor is significantly lower than the recoverable heat value in the vapor. The second major application area is in the thermo-mechanical pulp (TMP) plants in which a compressor can be used to recover heat from the TMP vapor normally vented to atmosphere. In this system, heat exchangers are needed to generate low-pressure steam using the latent heat of the TMP vapor. The steam compressor boosts the clean low-pressure steam to the pressure required in the paper mill. According to van Gogh, either screw or centrifugal compressors can be used [31]. Table 19.4 is a summary of the key features of these two types of compressors for vapor (steam).

Note that other types of compressors can also be used for steam (e.g., lobe, reciprocating, axial, turboblowers, etc.). Benstead has reported on some preliminary studies at the Electricity Council Research Center, Capenhurst, Chester, England, on a pilot steam-spray dryer for a ceramic [32]. The

dryer-exhaust steam was split off and compressed; the high-pressure steam gave up its latent heat to the incoming steam. He modeled two designs for comparison. One used a standard compressor that withstood the superheat. The other considered a compressor with water injection into the suction vapor to cool the steam during compression. The latter is about 10% more efficient due to reduction in steam volume at the reduced temperature.

Anthony has summarized the virtues of vapor-compression evaporation (VCE) followed by crystallization to treat waste brines from chemical process plants [33]. Anthony compares VCE followed by a crystallizer with other options, such as reverse osmosis and crystallizer, crystallizer plus centrifuge plus air-flash dryer, steam-heated crystallizer, and so on. He also compares the process performance, utility requirements, and initial and utility costs of treating cooling tower blowdown at utility plants for five different methods utilizing VCE. It is interesting to note that the capital costs of all combinations are within 10% of each other although the operating utility costs of steam, electricity, and natural gas or oil can vary by a factor as high as 5. A reverse osmosis/VCE/crystallizer/solar pond combination consumes only one fifth the electrical power consumed by a combination of VCE and spray dryer. The selection among different systems is thus dictated by the relative costs of fossil fuel versus electricity.

It is impossible to arrive at generalized guidelines at the present time regarding the technoeconomics of steam compression by thermocompression (e.g., steam jet ejectors) versus mechanical recompression for steam-drying plants. The costs of installation of such systems are such that they should be considered only for large-scale continuous steam dryers, that is, those with evaporation capacities of 100 t/d or higher. The steam compressors for such service are expected to cost over \$500,000 (1990 estimate for North America). To this cost, one must add the costs of spares and maintenance.

TABLE 19.4
Characteristics of Screw versus Centrifugal Compressors

Feature	Screw	Centrifugal
Suction capacity	500–25,000 m ³ /h	2000–100,000 m ³ /h
Maximum pressure ratio	6:1	2:1
ΔT	~60K	~20K
Capacity control	Variable-speed driver	Large variation possible without loss of efficiency
Other	Water droplets, dust may cause damage. No surge line.	Dry steam needed to avoid impeller erosion. Bypass control to avoid compressor surge.

It is important to note that existing steam-compressor technologies may not meet the requirements of some of the steam-drying systems, especially if very high degrees of superheat are needed. Often, two centrifugal compressors will be needed to boost the steam pressure to a level that is useful (e.g., 3 bar absolute from nearly atmospheric pressure). To increase the discharge pressure typically entails increasing the rotational speed of the impeller; in some cases, this may exceed the design speeds of commercially available compressors. Thus, some new design problems may arise to obtain discharge steam pressures desired.

Detailed design and technoeconomic calculations in close association with manufacturers of compressors, steam jet ejectors, and steam dryers and with utility companies are required to make confident conclusions regarding the optimal steam reuse and recycle systems for steam dryers.

19.5 CONCLUSIONS

Although touted as the drying technology of the future some two decades ago, the continuing low-to-moderate cost of energy has kept this technology from developing faster as low net-energy consumption is one of the key advantages of this technology, aside from quality enhancement and safe operation due to lack of fire and explosion hazards. Further, there are both product and process limitations to be overcome.

The cost of electricity as well as fossil fuels clearly drives the choice of the system chosen. There are significant capital costs involved in both superheated steam operation as well as in the energy recovery process. Development in newer and more efficient compressors for steam will give a boost to vapor recompression technology for steam dryers. Cleaning of the steam is another obstacle to be overcome. Overall, it appears that there are some niche markets for superheated steam drying but much laboratory, pilot scale as well as engineering design studies are needed before the technology can become widespread, e.g., drying of paper and tissue using superheated steam is a new technology that shows promise as shown by Kiiskinen and Edelmann [34]. The energy savings depend greatly on process integration and the cost of energy. Depending on available heat sinks the energy savings potential in paper drying can range from 15 to 85%. Despite significant R&D of the process many major problems remain, e.g., cleaning of the exhaust steam, fouling of heat transfer surfaces, and sealing of the dryer at the high speeds currently in practice.

ACKNOWLEDGMENTS

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20 Special Drying Techniques and Novel Dryers

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20.1 INTRODUCTION

Increasing demands for new and high-quality products, energy-efficient processes, environment protection, and the like have stimulated progress in drying science and technology. Mujumdar has presented a summary of the motivation for development of new drying technologies and identified a number of trends, which include [1]:

- Use of superheated steam in direct dryers
- Increased use of indirect (conduction) heating
- Use of combined (or integrated) heat transfer modes
- Use of volumetric heating (microwave [MW]/radio frequency [RF] fields) in specialized situations
- Use of two-stage (or multistage) dryers
- Use of intermittent heat transfer
- Use of novel combustion technologies (e.g., pulse combustion for flash drying)
- Use of novel gas–solids contactors (e.g., two-dimensional spouted beds, intermittent or rotating spouted beds, etc.)
- Design of flexible, multiprocessing dryers
- Combination of different dryer types, and so on

Many of the above-mentioned topics are covered in detail elsewhere in this handbook. There are numerous other technologies that have reached various stages of maturity—ranging from concepts and pilot-scale demonstrations to large-scale industrial applications. Several of these have been reviewed in the recent literature [2].

The focus of this chapter is on special drying techniques and selected novel dryers. In general, these technologies have specific applications and may offer advantages over conventional drying technologies under certain conditions. To a major extent, this chapter updates a number of other chapters in this hand-

book as it covers drying of various types of feeds (e.g., pastes, sludges, granular solids, etc.) and concentrates on the more recent developments not covered adequately in other chapters. Some of the dryers discussed here have found limited applications to date. The application base will likely increase in the future.

Superheated steam drying technologies have emerged as the drying technologies of the future in view of the numerous advantages (and challenges) they offer. This area is important enough that a separate chapter is devoted to this subject. Also microwave drying and radio frequency drying, which show promise of increasing industrial acceptance, are presented in depth in separate chapters and hence only MW and RF assisted dryers are included here.

Most of the novel dryers and special drying techniques are developed to meet one or more of the following criteria:

- Enhanced product quality
- Reduced energy consumption
- Higher productivity or capacity
- Reduced environmental emissions
- Production of new materials/products
- Enhanced flexibility
- Ability to combine other operations in the same unit
- Easier control
- More attractive technoeconomics, and so on

Often a trade-off is required between capital and operating costs. So, the final selection must be made not on the basis of novelty but on the overall technoeconomics of the drying system. It is important to stress that one should examine the drying system rather than the dryer in isolation. Furthermore, novel dryers have limited field experience, which makes laboratory and pilot testing especially important. Lack of prior experience should not, however, be taken as a negative factor against novel dryers.

Examination of new technologies reveals that most feature one or more of the following characteristics:

- Combined modes of heat transfer (e.g., combined dielectric or radiation with convection or conduction)
- Combination of conventional drying technologies (e.g., pneumatic-fluid bed, spray-fluid bed, spray-conveyor dryer, etc.)
- Novel gas–solids contacting (e.g., spout-fluidization, centrifugal fluid beds, impinging streams, intermittent spouting, etc.)
- New or more efficient methods of heat generation and utilization (e.g., pulse combustion)
- Combination of mechanical and thermal dewatering (e.g., filter-dryers)
- Use of superheated steam as a drying medium
- Use of nonconventional drying methods (e.g., supercritical drying)

Often the new technologies have evolved as a result of special application needs in a specific industrial sector. This does not preclude potential applications in other sectors, however.

In view of the large number of dryers and drying techniques covered, the present discussion is necessarily brief. The interested reader is referred for details to cited literature as well as to our recent book on *Novel Drying Technologies* [3]. It should be noted that not all types of dryers mentioned are available commercially at this time. It should also be noted that all proprietary and trade names used in this chapter are cited exclusively for identification purposes. There is no explicit or implicit endorsement of the products mentioned.

20.2 SPECIAL DRYING TECHNIQUES

20.2.1 DRYING TECHNOLOGY TO PRODUCE AEROGELS

Recent experimental studies on supercritical (also termed hypercritical) drying have revealed fascinating physical properties of the so-called aerogels (a term first proposed by Kistler in 1932) [4], that is, they are gelatinous materials that acquire the fractal structure of a solid skeleton due to replacement of the liquid moisture by a gas, frequently air. Among others, aerogels possess the following advantageous properties: small pore size (on average 2–20 nm) and large pore volume, which result in high porosity (0.90–0.97) and thus large specific surface area (in the order of 700 m²/g); relatively low density (70–250 kg/m³); low thermal conductivity (0.012–0.02 W/m·K); high

acoustic impedance (10^4 – 10^5 kg/m²·s); low sound velocity (100–120 m/s); low permeability (10^{-17} – 10^{-18} m²); and low refractive index (1.015–1.055) [5,6]. As applied to the catalytic processes, for example, drying in supercritical carbon dioxide to obtain polymer-based aerogel catalysts increases catalytic activity and selectivity, alters the pore structure toward a higher proportion of macro- and mesopores, and elevates the temperature stability limits over those of the conventionally dried polymers [7].

Aerogel technology is becoming increasingly important because of its numerous novel applications, such as for:

- Thermal insulators, both in domestic and industrial use (e.g., in solar collectors, multilayer windows)
- Acoustic transducers (e.g., in ultrasonic meters, sound-delay lines, sound absorbers)
- Components of high-quality glass (e.g., in optic elements, light pipes)
- Catalyst carriers (e.g., with oxides or noble metals deposited on silica-based aerogels)
- Liquid thickeners (e.g., to convert liquid nitrate acid into thixotropic fluid)
- Insecticides (living cell dewatering due to high adsorption capacity of aerogels)

Porous materials from gels can be obtained by [8]:

- Supercritical drying from the native solvent, usually alcohol
- Solvent replacement and subsequent supercritical drying using CO₂ or other low-critical-point solvent
- Convective drying at very low drying rates or using the drying control chemical additive (DCCA)
- Freeze drying

The first two methods yield a highly porous (porosity above 0.9) and transparent aerogel; the other methods produce less perfect and more dense porous solids called *xerogels*. The principles and practice of the above methods are summarized in the following sections.

20.2.1.1 Supercritical Drying from Native Solvent

In supercritical drying from a native solvent, an inorganic gel is synthesized in such a way that the alcohol fills pores just before drying. This is usually done by the hydrolysis of an alkoxide of a selected element (Si, Al, Zr, Ti, V, etc.) using a stoichiometric amount of

water. Unfortunately, this reaction is followed by condensation, which leads to sorption of the water molecules released from the polymeric network in a pore liquid. The pore liquid after condensation and aging is finally a mixture of alcohol with a certain amount of water. Kistler's idea was to eliminate surface tension leading to destruction of the gel's subtle structure by evaporating pore liquid above its thermodynamic critical point. This requires heating the alcohol-saturated gel in an autoclave above the critical temperature (at the same instant, pressure has to build up above its critical value) and then slow evacuation of the autoclave, maintaining the temperature still above the critical one.

One of the following options can be selected to ensure pressure buildup without evaporating liquid from the gel when below the critical temperature:

- Inject an extra solvent into the autoclave so at the critical temperature its vapor raises the pressure to the critical one.

- Fill and pressurize the autoclave with an inert gas (nitrogen, argon, etc.) so very little or no evaporation of solvent is necessary to reach the critical pressure at critical temperature. Depending on the initial pressure, the amount of additional solvent varies from the critical load to zero [9].

Technological considerations for this method of drying are as follows:

- Critical parameters of alcohol–water mixtures increase significantly with water content; it is advisable to wash out water before drying.

- Heating rate has to be slow enough not to induce cracks due to solvent thermal expansion.

- Extremely low permeability of gels hinders expulsion of expanded liquid from the pores; as a result, pressure builds up inside the solid.

- Evacuation of supercritical fluid has to be slow for the aforementioned reasons.

- Cooling of the autoclave before discharge must be accompanied by flushing with an inert gas to avoid condensation of the residual vapor.

This method requires more sophisticated autoclaving but leads to high-quality aerogels. Further, this technology is limited to production of thermally resistive gels.

20.2.1.2 Solvent Replacement with Carbon Dioxide

Smaller demands on the autoclave and lower temperatures are inherent to solvents such as CO₂, N₂O, Freons, and so on, of which carbon dioxide is the

most convenient to use in actual solvent replacement. They all have critical temperatures below 100°C and critical pressures below 100 bar.

Extraction with supercritical CO₂ is a well-known unit operation and may be employed to extract solvents from a gel. However, the supercritical CO₂ can be replaced by liquefied CO₂, which lowers the technological demands on the apparatus. The only condition to be met is that the original solvent be soluble in liquid carbon dioxide, which is the case of aliphatic alcohols, acetone, amyl acetate, and others.

The degree of solubility has a strong effect on the transparency of the aerogel. For example, at a certain concentration, the mixture of ethanol and liquefied CO₂ exists as a two-phase system. In such a case, cracking and product opacity may occur when the interface of these two liquids penetrates the gel [10]. When the solvent in a gel is fully replaced with liquefied CO₂, the temperature of the autoclave is raised above 31°C to cross the critical point. Evacuation time of the autoclave is determined by the solid size since large monolithic aerogels may crack if the evacuation rate is not low enough to release the tensile stresses. Experiments indicate that gels produced from CO₂ replacement are more opaque and more hygroscopic than those obtained from the native solvent evacuation at high temperatures, but the process is technically much easier to accomplish.

20.2.1.3 Convective Drying

Transparent and crack-free gels may be obtained at very low drying rates, controlled, for example, by diffusion through a few small holes made in a cover of the sample container. Surprisingly good results may be obtained by controlling the rate of convective drying by the so-called DCCAs [6,11]. The assumed role of the DCCA is to lower an evaporation rate from the solid surface via lowering of the surface tension in the surface layer of the solid. DCCAs' substances such as formamide or diethylene glycol are added to the sol before gelation. Initially, the pore liquid is evenly distributed but, due to selective evaporation, the surface layer becomes quickly depleted of the native solvent, leaving behind the DCCA, which does not evaporate so quickly because of its low volatility but forms a diffusion barrier. Although drying with DCCAs produces transparent, crack-free, partly shrunk xerogels, drying times are very long, being in the order of days even for gels not thicker than 10 mm.

20.2.1.4 Freeze Drying

Although freeze drying eliminates the problem of liquid menisci in the pores, the ice crystals usually

damage the subtle structure of the gel. In dense gels, the solid form is preserved during drying but less dense gels collapse and break into powder. The xerogel (or cryogel) monoliths produced in this way are inferior to dry gels produced by other methods; therefore freeze drying finds little application in aerogel technology.

20.2.2 CONTACT-SORPTION DRYING

Direct contact of a wet material with a heated particulate medium is an efficient method of solids drying. The particle-to-particle heat transfer mechanism is responsible for moisture evaporation when using hot inert particles like sand, salt, steel balls, and so on. In contrast, contact mass transfer is responsible for moisture removal if the wet solid is mixed with highly hygroscopic particles such as silica gel. The most promising method, however, is simultaneous contact heat and mass transfer, which occurs when the wet solid is exposed to heated hygroscopic particles. Both natural adsorbents, such as bentonite, zeolite, chabazite, and the like, and synthetic materials of high sorption capacity and resistance to thermal shock like alumina or molecular sieves, can be used for contact-sorption drying.

20.2.2.1 Sorption Drying of Particulates

A continuous process apparatus for sorption drying of particulate materials by contact with hot inert particles is depicted in Figure 20.1. The unit consists of three coaxially located conical chambers with a central heating chamber in the form of a conical drum supported

axially by drive and stationary shafts. The stationary shaft is made as a hollow cylinder, thus permitting gas feed to the propane burner situated on the longitudinal axis of the heating chamber. The gas burner, fed additionally by compressed air for complete combustion, generates a flame that extends into the heating chamber and directly heats the inert particles whereas they traverse the heating chamber. For better heating, the inert particles are showered by scope-like flights staggered at the chamber wall.

Wet particulates fed to the dryer are mixed with hot inert particles discharged from the outlet end of the heating chamber and conveyed continuously by the auger disposed between the heating chamber and the outer conical drum. This drum is thus rotated with the heating chamber and this rotation causes the mixture to be displaced in the drying section. As the granular mixture is conveyed, the moisture evaporates due to sensible heat stored in the hot inert particles. The tail end of the outer drum is screened to permit the inert particles to pass through and to be collected in an annular cone-shaped collector. The end of the collector is a peripheral annular chute equipped with buckets to transport the inert particles back to the feeding chute.

The residence times of the inert particles in the heating chamber and then in a drying section are relatively short, in the order of 15 to 20 s. Thus, the inert particles recycled to the heating chamber are still hot. While traveling through the heating chamber, the initial temperature of the recycled particles of about 100°C is increased by approximately 100 to 200°C. Therefore, particles enter the drying section at an operating temperature usually in the range of 200 to 300°C.

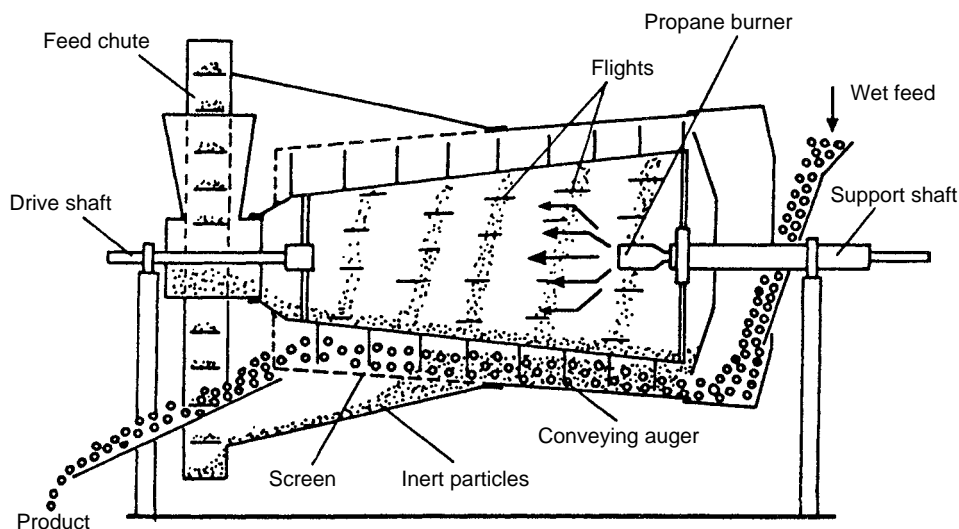


FIGURE 20.1 Continuous contact-sorption dryer for particulate materials.

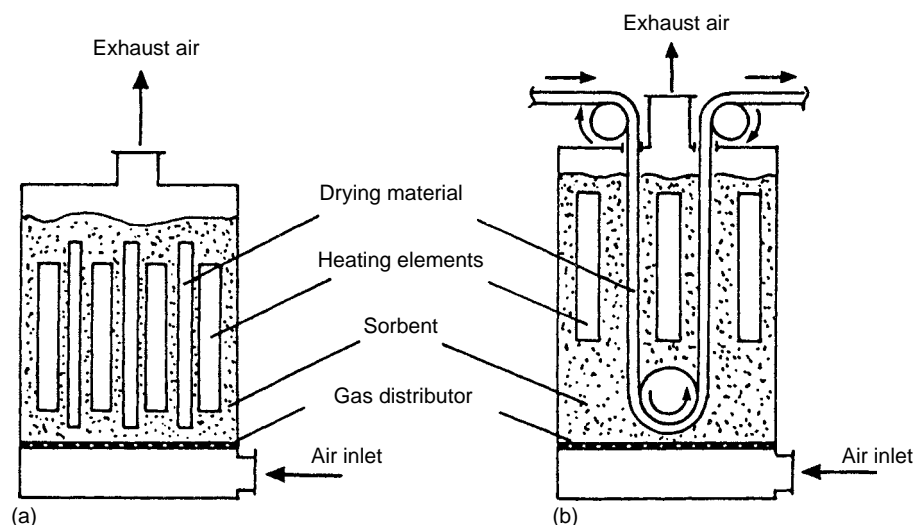


FIGURE 20.2 Contact-sorption dryer with active fluidized bed: (a) batch; and (b) continuous.

Although originally developed for drying corn with sand as the heat transfer medium, it can also be used for drying grains by solid sorbents, parboiling rice, roasting soybean to remove trypsin and improve protein digestibility, processing ground nuts and oil seeds, and more [12]. With a length of 1.2 m and diameter of 0.8 m, a contact dryer can process 750 kg of corn per hour by raising its temperature from 15 to 75–90°C. The sorbent-to-particulates mass ratio, which affects the drying rate, can be controlled by the wet feed rate, mass of sorbent in the system, and angular velocity, with the recommended value ranging between 100 and 160 rpm [13].

20.2.2.2 Drying in an Active Fluid Bed

A novel technique for contact-sorption processing of large sheet-type solids such as textiles, leather, or veneer is drying in an active fluidized bed [14,15]. In this technique, the material to be dried is immersed in a fluidized bed of highly hygroscopic ('active') particles that extract moisture from the wet material and release it to the fluidizing air (Figure 20.2). In a continuous system, the band of drying material is transported in a serpentine fashion through the fluidized bed of sorbent particles. Heat required for moisture release from the sorbent to fluidizing air is supplied by immersed heaters. The use of fluidized beds of active particles enhances not only the convective drying but also material–sorbent, sorbent–sorbent, and sorbent–heater contact heat and mass transfer rates. Comparison of the active fluid bed dryer with the cylinder dryer is given in Table 20.1.

As compared with convective drying, the application of fluidized silica gel for drying 1.5 to 2.5-mm thick cowhides from 60 to 20% moisture at 60°C allows reduction of the drying time from 7 h to 15–20 min, reduction of the steam consumption by 30%, and lowering of the capital cost by a factor of 3. The dryer is of modular type with up to 20 modules that form a compact assembly 2.1 m long, 2.2 m wide, and 2.0 m high [15].

20.2.2.3 Sorption Drying-Granulation

A fluidized bed of active particles can also be used for drying-granulation of highly adhesive materials. The fine particles of a suitable sorbent are fluidized by hot

TABLE 20.1
Comparison of Active Fluid Bed Dryer with Cylinder Dryer for Polcorfam^a

Parameter	Fluid Bed Dryer	Cylinder Dryer
Drying medium	Air	Steam
Temperature, °C	100	160
Drying time, min	10	37
Unit heat consumption, kJ/kg H ₂ O	4200	22,300
Serpentine length, m	12	56

^aUnwoven fabric for artificial leather.

Source: From Technical information. Institute of Chemical Engineering, Warsaw Technical University, Warsaw, Poland.

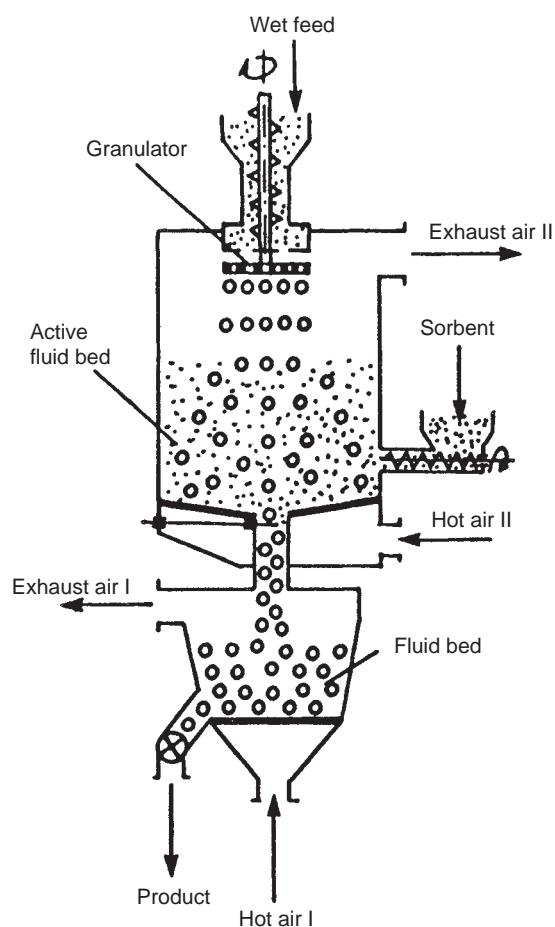


FIGURE 20.3 Contact-sorption fluid bed dryer for pasty materials.

air in the upper chamber of a two-stage dryer-granulator (Figure 20.3). Wet granules formed in a disk-type granulator are coated by the hot, pulverized sorbent as they fall through the fluid bed. Due to the high rates of the contact heat and mass transfer, the surface layers of the granules are rapidly desiccated. The resulting porous (and thus vapor-permeable) dry coating facilitates further moisture diffusion and does not allow the lumps and granules to aggregate. Drying of the coated loose granules to the final moisture content is then carried out in a standard fluid bed formed in the lower chamber of the dryer [16].

20.2.2.4 Sorption Drying Using Filler Materials

With respect to the product, the solid sorbent can be neutral (placebo) or active. In the first case, the sorbent is used only as a carrier for heat-sensitive and otherwise hard-to-dry materials. Liquid bioproducts such as antibiotics, enzymes, yeasts, amino acids, and the like, which, when conventionally dried, can lose up to 70% of their biological activity, are typical examples. Sawdust, activated carbon, and ground straw or hay are typical solid carriers. After drying, the solid sorbent is either separated from the product or left as is for further utilization, as it is in the case of fodder antibiotics [17]. Figure 20.4 shows the co- and counter-current modes of contacting solid carriers with liquid biomaterials in a spray dryer configuration.

An interesting option for contact-sorption drying is the use of a sorbent that can be incorporated into the drying material as an integral and essential part of the final product. Wheat bran, starch, powdered skim milk, casein, peat, corn flour, and the like

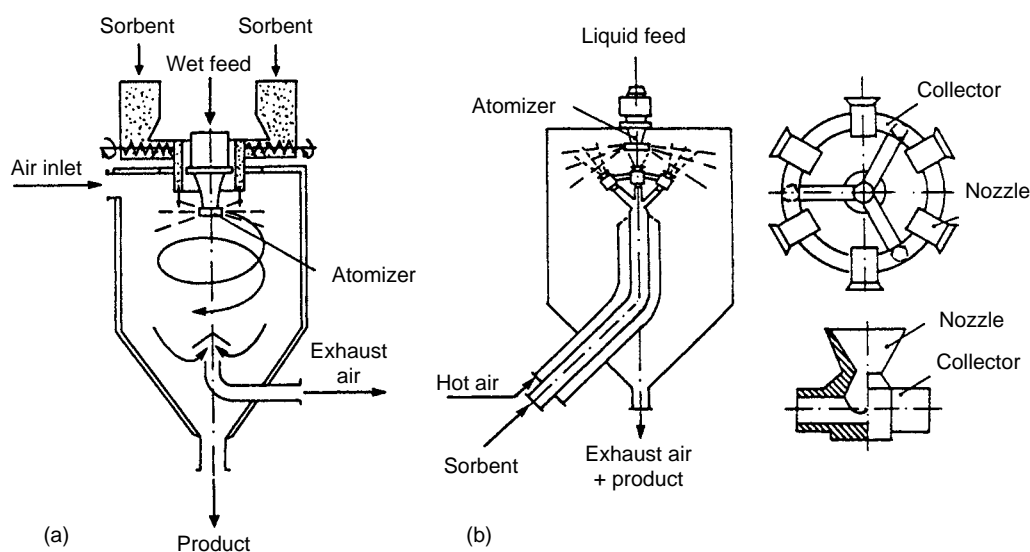


FIGURE 20.4 Contact-sorption spray dryers: (a) cocurrent; (b) countercurrent.

are reported to be good fillers for drying liquid bio-products in fodder, fermentation, pharmaceutical, and similar industries [18,19]. As applied to drying of lysine, for example, the use of wheat bran as an active filler gives the following advantages:

- Reduces by two the relative hygroscopicity of the product and thus lowers three- to fourfold the rate of moisture sorption during storage
- Shifts the equilibrium humidity from 50 to 65–70% in the temperature range 20 to 50°C
- Eliminates otherwise necessary postdrying operations such as mixing or tempering
- Stabilizes moisture evaporation during drying, which in fact takes place from a porous material instead of from a highly viscous droplet
- Eliminates product buildup on dryer walls
- Lowest material temperature during drying, thus preserving up to 98% of its biological activity in the liquid state

It may be noted that the sorption capacity of organic sorbents can be improved significantly (even by 40%) in relatively small (5–10% w/w) admixtures with salt or lactose [18].

Figure 20.5 shows a schematic of a contact-sorption dryer for biological materials containing living cells from the lactic bacteria group (*Lactobacterium plantarum*, *Lactobacterium acidophilum*, *Streptococcus lactidiatensis*, etc.). Liquid biomass at solid concentrations of 0.01 to 0.05 kg/kg and solid sorbent-filler are fed at the top of the dryer by a pneumatic feeder of a special design shown schematically in Figure 20.6 (performance data for such a feeder are given in Table 20.2). The chopped or pulverized filler is forced through the annulus of the feeder into the mixing chamber where it adsorbs the liquid biomass dispersed within it by compressed air. The semisolid mixture flows out of the feeder as a spray of fine granules. These granules are predried convectively in free fall, and dried to the required moisture content in a moving bed fluidized additionally by compressed air supplied to the dryer via an air distributor. An inclined grid supports the granules and facilitates continuous discharge of the dry product. Spraying of water or raw biomass in a dryer above the feeder level (in the amount of 1 to 2% of the main feed) reduces entrainment of fines with exhaust air.

20.2.3 DRYING ON INERT PARTICLES

20.2.3.1 Fluid Bed Drying

Drying of liquid and pasty materials on the surface of heated inert particles is one of the newer drying

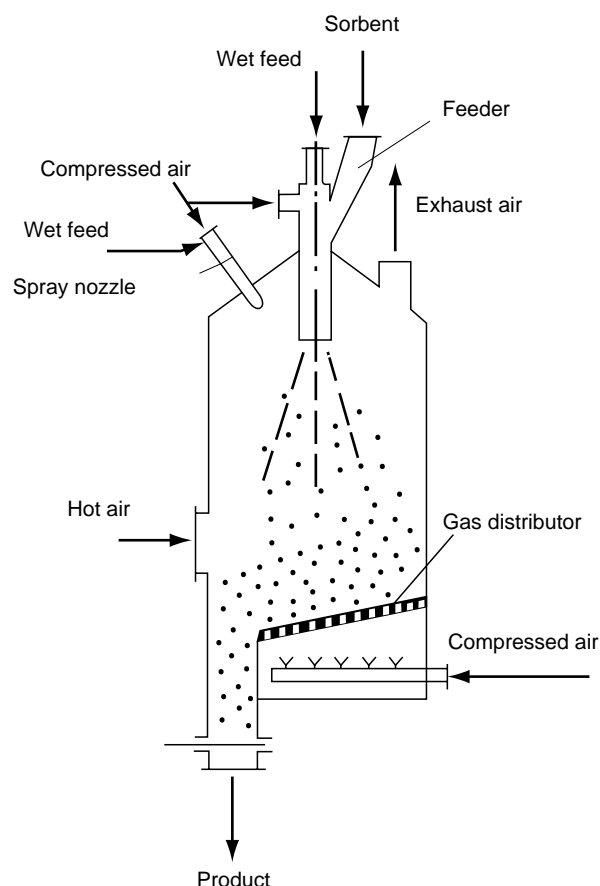


FIGURE 20.5 Contact-sorption dryer for liquid biomaterials.

techniques. The inert material serves not only as a carrier for the liquid film but also as a heat transfer medium. The size of the inert particles may be from 20 to 40 times larger than that of the dispersed material dried, which allows for higher gas velocity and thus increased dryer productivity. Moreover, intensive motion of the inert particles (e.g., in fluid or spouted beds) results in good dispersion of the liquid feed, so coarse spray nozzles can be used. As compared to the standard spray dryer, the use of a fluid bed of inert particles to dry slurries results in 15- to 17-fold higher volumetric evaporation rates under identical thermal conditions. Table 20.3 presents some drying performance data for a fluid bed dryer with inert particles shown schematically in Figure 20.7 [20].

For drying of liquid or pasty materials that change their physical properties significantly with moisture content (e.g., lactic sugar, egg albumen, L-lysine, hemoglobin), a two-chamber fluid bed dryer (Figure 20.8) appears to be more suitable. Controlling the liquid feed rate with simultaneous change of inert particle flow from the internal (predrying) chamber to

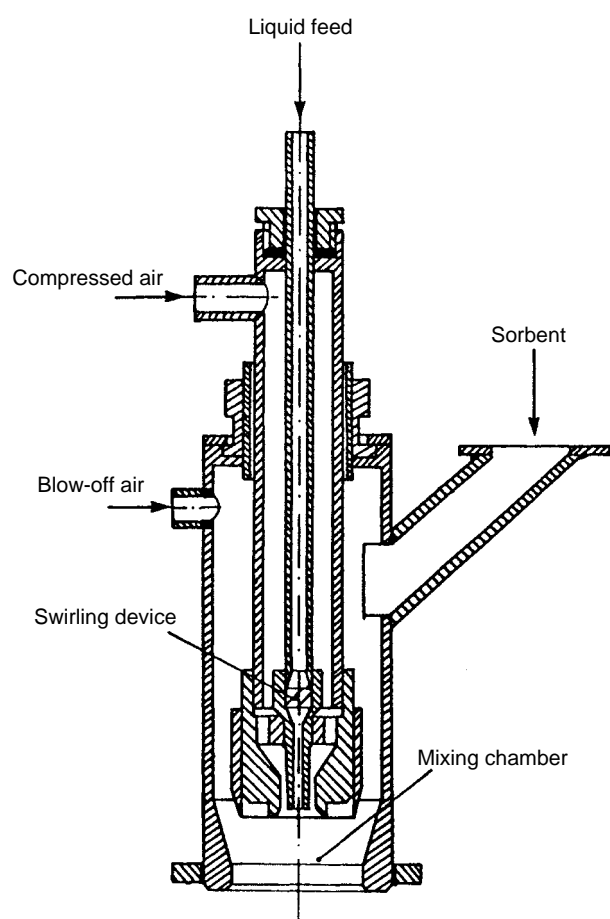


FIGURE 20.6 Three-stream pneumatic feeder.

the external (final drying) one enables control of the heating rate, residence time, and kinetics of product release from the inert surface.

20.2.3.2 Drying in a Vibro Fluid Bed

Intensive motion of particles in a vibro fluid bed (VFB) appears to have the potential for drying of pasty materials when sprayed onto a bed of vibrating inert particles. Figure 20.9 presents such a novel design, originally developed for drying of whole egg but successfully used for drying of other biomaterials such as egg white, animal blood, casein, yeast, soya protein, and so on [21]. The dryer is comprised of a vertical cylindrical chamber with a set of horizontally located and oppositely directed pneumatic nozzles. A grid-type gas distributor supports inert particles. In contrast to the conventional VFB dryers, this grid is freely mounted with a small clearance to the chamber wall, and vibrated vertically by an eccentric drive; this allows for reduction of the power requirements. The feeding pump forces the liquid or pasty feed toward the two-stream pneumatic nozzles, which spray it into a vibrating bed of polytetrafluoro ethylene (PTFE) cubes, fluidized additionally by a through flow of hot air.

As it coats the inert surface, the wet film is dried due to rapid convective heat and mass transfer and, to certain extent, conductive heat transfer from the bed of inert particles. As a result of the intense, high-frequency collisions between particles, the dry shell

TABLE 20.2
Characteristics of a Three-Stream Pneumatic Feeder

Parameter	Type			
	AF-1	AF-2	AF-8	AF-10
<i>Throughput, kg/h</i>				
Sorbent	400	800	1000	800
Liquid	700	1500	2000	1500
Air consumption, kg/h	700	1500	2000	1500
Air pressure, MPa	0.3	0.3	0.3	0.3
Swirling device	Spiral	Spiral	Tangential	Tangential
Solids concentration in sprayed liquid, % mass	<40	<40	<40	<40
Spray angle, °	60–180	60–180	60–180	60–180
<i>Dimensions, m</i>				
Length	0.9	1.0	1.1	1.0
Width	0.12	0.16	0.16	0.15
Height	0.16	0.21	0.21	0.17
Mass, kg	13	14	14	13

Source: From *Drying and Thermal Processing of Wet Materials*. 1990. Nauka i Tekhnika, Minsk (in Russian).

TABLE 20.3
Performance Data for a Fluid Bed Dryer with Inert
Particles for Selected Materials

Parameter	Material	
	Sludge of Phosphorous Fertilizer	Suspension of Fluorosilicates
Throughput (per dry product), kg/h	2500	1500
Grid area, m ²	4	2
Moisture content, % wb		
Initial	70–80	25–20
Final	1.0–1.5	0.05
Inlet gas temperature, °C	650–700	650
Bed temperature, °C	220–230	160
Air consumption, N m ³ /h	33,000	14,000
Volumetric evaporation rate, kg/m ³ h	170	28
Diameter of inert particles, m	0.001 ^a	0.0016–0.002 ^b

^aGranules of dry product.

^bCorundum.

Source: From Kaganovich, Yu.Ya., *Industrial Dewatering in Fluidized Bed*, Khimiya, Leningrad, 1990 (in Russian).

formed on the inert surface breaks and disintegrates into powder, which is carried away with the airstream to the cyclones. Then, the inert surface, which is free of dry film, is covered once again with a wet film, and the coating-drying-shell-removal process repeats until a dynamic equilibrium, ensuring continuous operation, is attained. Since adhesion is the governing parameter in the drying mechanisms, only materials with low adhesiveness toward the inert particle's surface can be processed in this system. In fact, the VFB dryer has successfully been applied for egg white, egg yolk, animal blood, casein, yeasts, soya protein, meat-and-bone broth, and more. Negative results were obtained in drying of protein hydrolysates, whole and skim milk, coffee, tea, and fruit pulps, which tend to form firm and highly adhesive films even at the PTFE surface [21].

The industrial system incorporates a dryer, feed unit, product receiver, air blower with steam heater, and cyclones with powder collectors. Table 20.4 presents the main operating parameters of a VFB dryer with inert particles for drying whole eggs.

Compared with spray drying, the VFB dryer with inert particles is more efficient; in particular, it allows for reduction of the electrical energy by a factor of 2 and 5- and threefold reduction of space and height requirements, respectively (Table 20.5). Moreover, the egg powder is free flowing, does not agglomerate while stored, and has a high degree (over 98%)

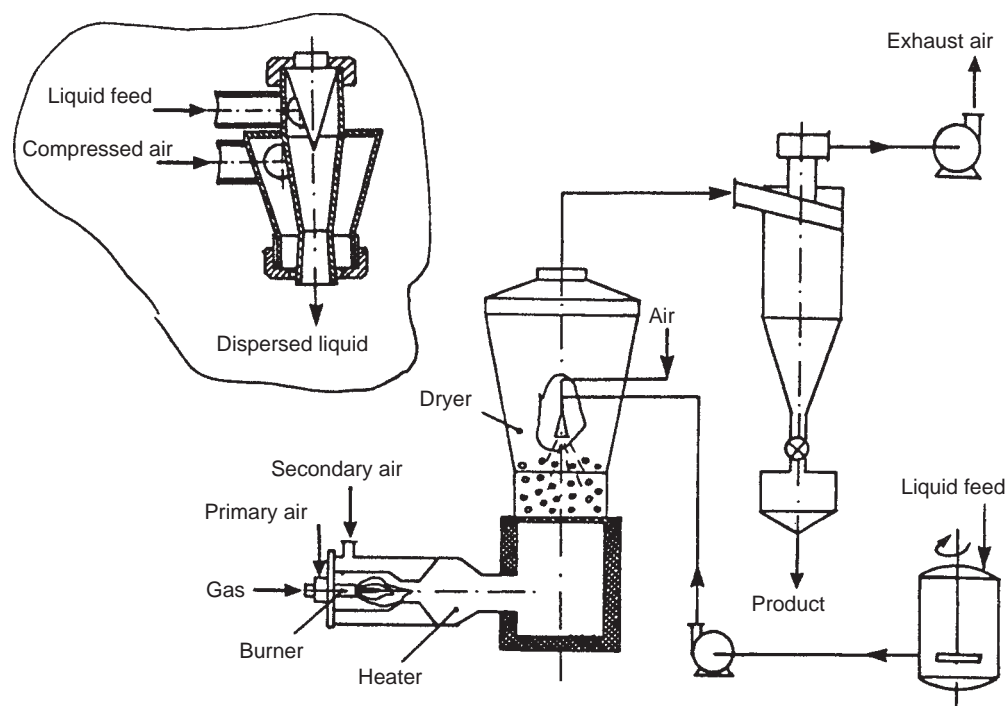


FIGURE 20.7 Fluid bed dryer with inert particles.

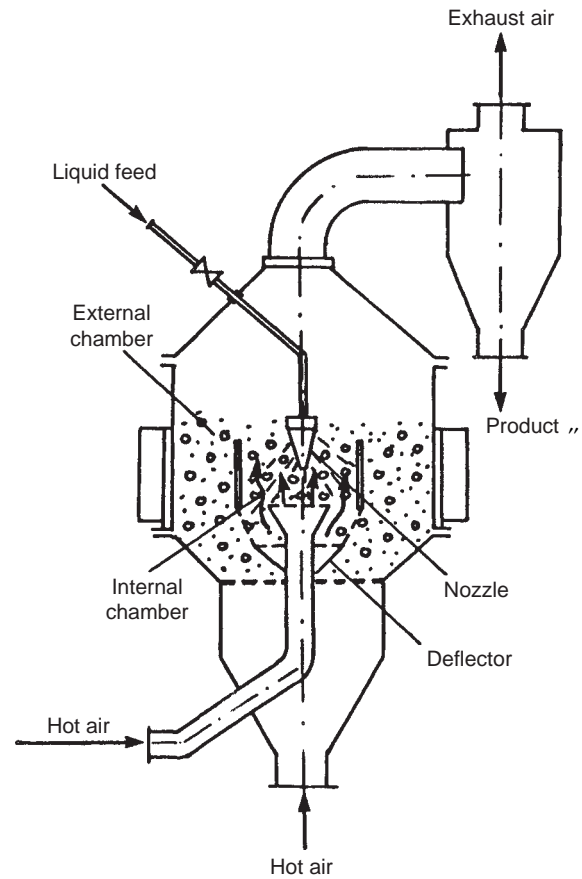


FIGURE 20.8 Two-chamber fluid bed dryer with inert particles.

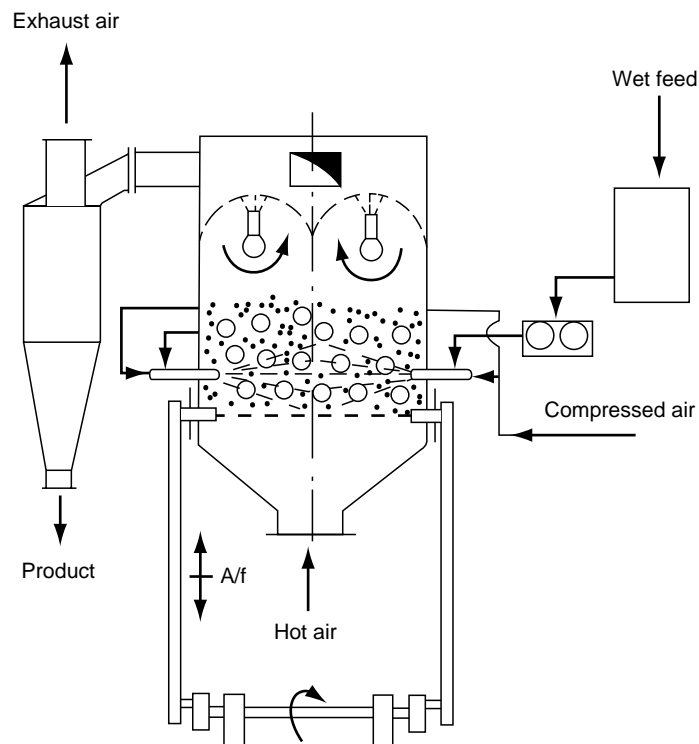


FIGURE 20.9 Vibro-fluidized bed dryer with inert particles. (From Technical information, NPO “MIR” (Scientific-Industrial Association), 123308 Moscow, Marshal-Zhukov Prospect 1.)

TABLE 20.4
Characteristics of Industrial Vibro Fluid Bed Dryers with Inert Particles for Drying Whole Eggs

Parameter	Unit		
	A1-FMU	A1-FMYa	A1-FMB
Throughput, kg feed/h	75–80	200	410
Initial moisture content, % wb	74	74	74
Final moisture content, % wb	5–7	5–7	5–7
Evaporation capacity, kg H ₂ O/h	59–76	150	300
Air consumption, m ³ /h	<5000	<12,500	<25,000
Steam consumption (0.3–0.4 MPa), kg/h	200	500	920
Electric energy consumption, kW h	20	40	83
Ice-water consumption ^a (3°C), kg/h	300	500	800
Grid area, m ²	0.32	0.8	1.6
Vibration frequency, Hz	7.5	7.5	7.5
Amplitude of vibration, m	0.008	0.008	0.008
Air inlet temperature, °C	120	120	120
Mass of inert (4-mm PTFE cubes), kg	55	150	300
<i>Overall dimensions</i>			
Length, m	2.6	4.5	7.0
Width, m	2.1	2.9	4.5
Height, m	3.2	4.0	4.5

^aWhen cooling of product is required.

Source: From Technical information, NPO “MIR” (Scientific-Industrial Association), 123308 Moscow, Marshal-Zhukov Prospect 1.

of solubility in water (Table 20.6). The quality of spray-dried product is generally different from that obtained with the VFB dryer.

20.2.4 COMBINED FILTRATION AND DRYING

Since in many batch-drying technologies a product to be dried is previously filtered and washed, conducting these multiple unit operations in a single-process vessel can offer the following benefits:

Improved product quality due to combination of various operations in one apparatus (high purity, no product contamination)

Improved economics and production efficiency (no wet material transport, no product loss, lower consumption of wash liquid, shorter batch time)

Increased process flexibility and control (normal, countercurrent, or reslurry washing; operating under vacuum or overpressure)

TABLE 20.5
Dryer Indices for Spray Dryer and Vibro Fluid Bed Dryer with Inert Particles

Index ^a	A1-FMU	VFB Dryer with Inert Particles (Russia)		Spray Dryer RS-150 (Yugoslavia)
		A1-FMYa	A1-FMB	
Energy consumption, kJ/(kg h)	3600	2772	2736	7416
Steam consumption, kg/kg	10.0	9.6	8.4	15.9
Throughput per unit floor area, kg/m ²	2.13	2.0	2.75	0.79
Throughput per unit volume, kg/m ³	0.53	0.5	0.6	0.08

^aPer kg of product.

Source: From Technical information, NPO “MIR” (Scientific-Industrial Association), 123308 Moscow, Marshal-Zhukov Prospect 1.

TABLE 20.6
Selected Quality Indices for Whole Egg Powder
Dried in a Spray Dryer and in a Vibro Fluid Bed Dryer
with Inert Particles

Parameter	Spray Dryer	VFB Dryer
Shape	Spheres	Scales
Size, mm	0.02–0.05	0.05–0.06 thick; 0.2–0.4 long
Adhesion, N/m ²	4000	2000
Bulk density, kg/m ³	300–350	450–500

Source: From Technical information, NPO “MIR” (Scientific-Industrial Association), 123308 Moscow, Marshal-Zhukov Prospect 1.

Reduced environmental pollution and minimized personnel exposure to hazardous media (closed system allows also for operation with toxic products and organic solvents)

Reduced investment and running costs (no separate filter, centrifuge, wash tank, dissolver, etc.)

Compact design (reduced space requirement, easy integration into existing building, lower installation and construction costs)

Higher drying rate (convective or contact drying with surface renewal due to filter cake agitation)

Flexibility (cold or hot air blowing, pressure or vacuum drying, drying in inert atmosphere)

Filter drying is especially appropriate for sterile batch processes in which solids have to be separated from the mother liquor mechanically and thermally in an enclosed system. It is particularly suitable for process plants in which frequent product changes are made (e.g., for the manufacture of pharmaceutical intermediate and final products, dyes and pigments, fine chemicals, agrochemicals, foodstuffs, etc.).

There are two basic configurations in industrial use for the combined filtering and drying method: a vertical filter with a heated agitator and a horizontal filter with a revolving housing.

20.2.4.1 The Vertical Filter-Dryer

The key to the vertical filter-dryer is a heated paddle agitator (Figure 20.10) designed as a three-arm structure with aerofoil blades that provide for increased stability at high torque levels and extended heat transfer area. This unique arrangement promotes axial and radial mixing and uniform drying by continuous exposure of each particle to the heated surface of

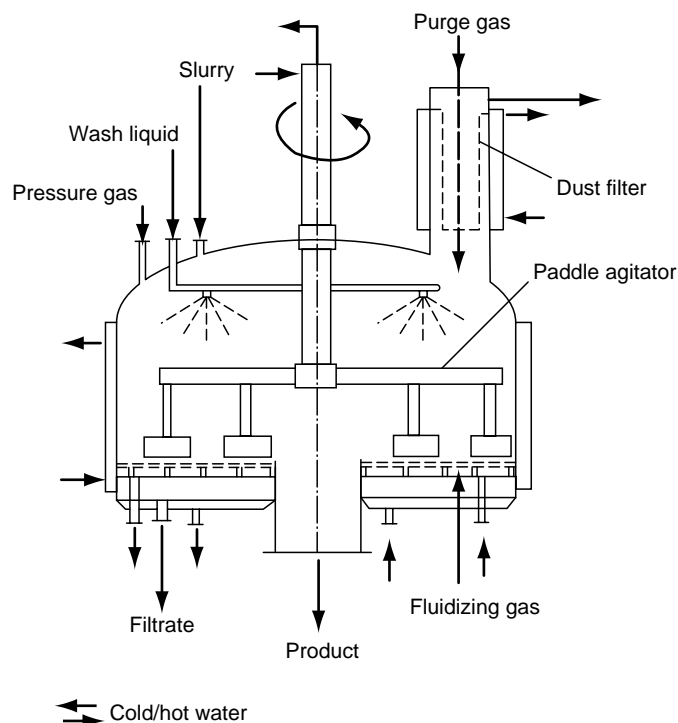


FIGURE 20.10 Vertical filter-dryer. (Courtesy of De Dietrich Process Systems, Inc., Charlotte, NC. With permission.)

the interior vessel walls, filter plate, paddle agitator, and the individual particles themselves.

Generally, after filtration and washing, most drying processes begin with a cold blow-through step to reduce the initial moisture content by liquid entrainment. Convection drying with hot gas adds thermal energy into the cake for evaporation and continues the mechanical liquid entrainment. Finally, vacuum applied both from above and below the filter cake with intense agitation combined with contact heating from the agitator, vessel walls, and filter plate brings the product moisture to its required value, frequently below 0.5%. The representative results for a 8-m² filter-dryer are [22]:

1. The use of a heated agitator increases the heating rate by up to 70% over that provided by a filter-dryer with no agitator.
2. The heat transfer coefficient from the heated paddles is three to four times larger than that from the filter plate and up to 1.5 times larger than that of the vessel walls.
3. The estimated heat input contributions by component for the heated paddle agitator, vessel walls, and filter plate are 45, 30, and 25%, respectively.

Typical throughput of a 8-m² vertical filter-dryer is 1250 kg/batch at a cake depth of about 0.4 m and an agitator speed of 6.5 rpm. Clearly, these depend on the feedstock.

Figure 20.11 presents a schematic of a vertical filter-dryer with a helix-shaped filter element. During the filtration and cake-washing stages, the unit operates as a conventional filter with a filter cake formed on the upper surface of the hollow helix that serves as the filter grid. Then, steam or heated gas is introduced into the vessel in which it passes through the cake into the helix interior. Alternatively, a heating medium can be introduced into the helix fitted with internal tubing whereas a vacuum is applied to the vessel. The discharge solids are loosened from the filter element by spinning the helix, which further acts as a screw conveyor to transport dry and loose cake to the bottom of a pressure vessel. A scraper, which is integral with the helix, then discharges the cake through the outlet into a closed receptacle. The helix filter dryer finds applications in pharmaceutical and fine chemical manufacturing processes, from coarse salts to fine antibiotics. Typical feed slurries may have solids concentrations of less than 1% by weight. Due to the ease of automatic discharge, these filters have practical applications with solids concentrations up to 10% or

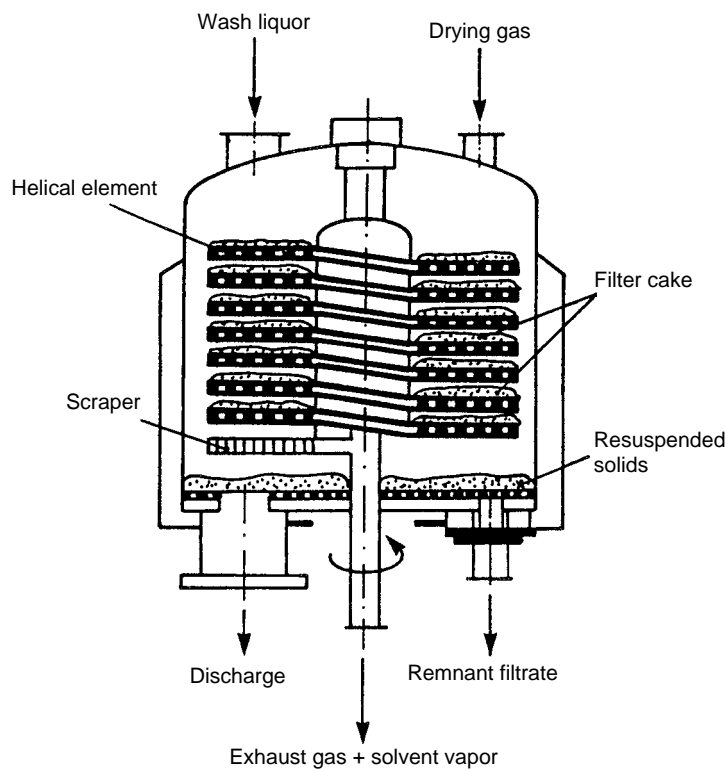


FIGURE 20.11 Filtroba helical filter-dryer.

more. A choice of sintered metal or cloth filter media allows for particle capture down to 0.5 μm . The ability to agitate rapid-settling slurries by rotating the helix makes it possible to also process coarse particles [23].

20.2.4.2 In-Line Filtration and Drying

A combination of gravity and mechanical and thermal dewatering is the major feature of the energy-efficient system developed by Atlas-Stord, Incorporated (Greensboro, NC), for dewatering sticky, viscous, difficult-to-process materials such as fish meals, municipal sludges, or wastewater from petrochemical, pharmaceutical, pulp and paper, and other industries. The system consists of a mechanical sludge process and an indirectly heated ROTADISC® dryer that utilizes steam, oil, or superheated water as the thermal medium.

Basically, the ROTADISC dryer consists of a set of hollow disks that form the heat transfer surface and are mounted on a central hollow shaft as depicted in Figure 20.12. The heat transfer medium flows through the rotor (disks/shaft), which is completely submerged in the product processed. The rotor assembly revolves within a fixed cylindrical stator, which may also be heated to increase the heat transfer area by up to 12%. The wet material contained within the stator is transported through the unit in a plug flow, passing through the annulus formed by the tips of the disks and the stator. The moisture evaporated due to contact heat transfer is collected in the upper part of the stator, which extends into a plenum to form a vapor dome, and is condensed in a heat exchanger to provide makeup water for boilers or other plant needs. Paddles on the disk tips aid progression; they can be adjusted to optimize the dryer for each specific product. These paddles in combination with

the scraper bars welded to the stator between each disk create vigorous material motion within the dryer. The resulting “turbulence” improves contact heat transfer, self-cleans the heating surfaces, releases the vapor into the vapor dome, and eliminates the need for recycling even with evaporator concentrate added.

As applied to processing wastes from the pulp and paper industry, for example, the primary and secondary sludges from the paper mills are fed at a controlled rate into a blend tank. The homogenized mixture is then pumped to a rotary screw thickener for initial gravity dewatering. The thickened sludge, at approximately 5% solids, is transported by gravity to the twin-screw press for mechanical dewatering. The pressed material at 25 to 35% solids is fed to the ROTADISC processor for thermal drying (scalping). The waste heat of vapor generated from the scalping of the sludge is used to preheat the incoming slurry before the addition of any polymer conditioners. The scalped sludge can be incinerated in a fluid bed combustor equipped with boilers for generating high-pressure steam [24]. The system is flexible enough to handle various sludges and still maintain a consistent feed material for incineration. Operating data for several types of ROTADISC dryers are given in Table 20.7.

20.2.5 PULSE COMBUSTION DRYING

Whereas the idea of pulse combustion has been known since the turn of the century, interest in this technique to enhance energy-intensive processes has grown in recent years because of its potential to increase productivity, reduce pollutant emissions, and produce significant fuel savings. The term pulse combustion originates from the fact that the combustion and heat-release processes vary periodically in time. Such periodic combustion of a liquid or gaseous fuel

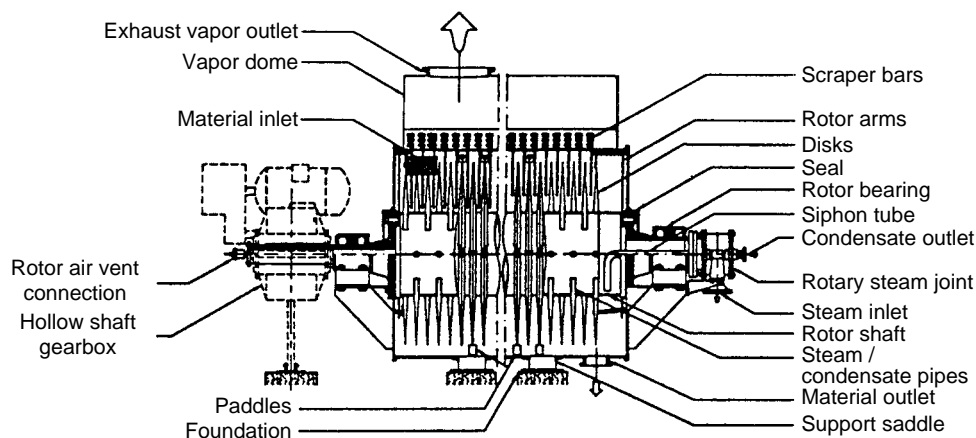


FIGURE 20.12 ROTADISC filter-dryer. (Courtesy of Atlas-Stord, Inc., Greensboro, NC. With permission.)

TABLE 20.7
Operating Data on Several Types of ROTADISC Dryers for Sludge Processing

Parameter	Dryer Type		
	TST-70R	TST-60R	TST-20R
Sludge/process type	Municipal (digested primary and waste activated)	Pharmaceutical (secondary only, aerobic fermentation)	Municipal (secondary only, waste activated)
Throughput, m ³ /h	10,833	250	5625
Moisture content, kg/kg db			
Before dryer	0.77	0.85	0.70
After dryer	0.50–0.57	0.05–0.10	0.55–0.60
Sludge-conditioning method	Thermal and polymer	Polymer and FeCl ₃	Polymer and FeCl ₃
Mechanical dewatering	Centrifuges	Centrifuges	Plate/frame filter
Heating surface area, m ²	480	816	77
Reflux ratio	None	10:1	None
Dry sludge use	Fuel (fluid bed combustor)	Fertilizer (bulk sales)	Fuel (fluid bed combustor)
Evaporation rate, kg H ₂ O/h	6000	8480	1540
Steam			
Parameters	6 bar at 165°C	4 bar at 155°C	7.5 bar at 173°C
Consumption, kg/h	8,000	10,430	2,080
Steam ratio (generated/demand)	1.13	n/a	3.43

Source: From Stord, Inc. 1993. *The Stord Report*, Stord, Inc., Greensboro, NC.

creates pressure, velocity, and, to a certain extent, temperature waves propagated from the combustion chamber via a tailpipe to the process volume (applicator). Further, these waves affect the momentum and heat and mass transfer rates, which finally result in enhanced rates of such processes as liquid dispersion or moisture evaporation.

Although some processes can be carried on in a tailpipe of the pulse combustor, the typical pulse combustion system consists of a combustor and an applicator in which the pressure waves are amplified by acoustic resonance. The frequency of the pressure pulsations is generally close to the frequency of the fundamental acoustic mode of the combustion chamber and the applicator. If properly tuned, the pulsed combustor can excite large-amplitude pulsations in a process (e.g., drying, calcining, or incineration) carried out downstream of its tailpipe.

Numerous studies on a variety of pulse combustor designs have demonstrated that a pulse combustor can offer the following advantages over the conventional (i.e., continuous) combustion systems [25–27]:

- Increased heat and mass transfer rates (by a factor of 2 to 5)
- Increased combustion intensity quantified by the gas mixing index (by a factor of up to 10)
- Higher combustion efficiency with low excess air values

- Reduced pollutant emissions (especially NO_x, CO, and soot) (by a factor of up to 3)
- Improved thermal efficiency by up to 40%
- Reduced space requirements for the combustion equipment

Furthermore, some pulse combustors are self-aspirating so they do not require a separate fan to supply combustion air. Because of the high temperature and large volume of the combustion products, there is in general no need for a blower.

Industrial applications of pulse combustion were limited so far to space and water heating mainly because of the difficulty of resonating the process volume at a fixed frequency if its acoustic characteristics vary with operating conditions. However, the progress in tunable-frequency pulse combustors that can be operated over a range of frequencies has made it possible to use this technique for energy-intensive industrial processes such as drying, calcining, heating of steel, solid and hazardous wastes incineration, coal and black liquor gasification, carbon activation, cement production, and the like. As applied to drying, the use of pulsed combustion enables one to [25–27]:

- Increase (by a factor of 2–3) the drying rate
- Eliminate property (e.g., temperature) distribution within the dryer, which improves the product quality

- Lower gas and product temperatures during processing
- Handle sticky materials that form lumps or aggregates without mechanical mixing or disintegration
- Atomize most pumpable liquids without the need for an atomizer or high-pressure nozzles
- Reduce unit air consumption by 30 to 40%
- Lower air volumes discharged to the atmosphere
- Eliminate an air blower from the system

The operation of a pulse combustor is controlled by a complex interaction between an oscillatory combustion process and acoustic waves that are excited inside the combustor (Figure 20.13). Ignition of the fuel–air mixture by a spark plug is followed by explosion-like burning and rapid temperature and pressure increases. Because of closed valve, that pressure initiates a flow of combustion gases into the tailpipe. As long as the pressure rise produced by the combustion process is larger than the pressure decrease due to the gas flow off the tailpipe, the combustion pressure increases. When the combustion-generated

pressure becomes lower than the outflow-produced pressure drop, the combustor pressure starts decreasing (Figure 20.14). Owing to inertia, the exhaust of the combustion products continues for a quarter of the pulsation period after the combustion pressure drops below the average pressure level (cooling of the combustion products). When the combustion pressure reaches the minimum, the flow reverses direction and gases from the tailpipe reenter the combustor. As the combustor pressure drops, the valve opens and admit new charges of fuel and air into the combustor chamber. These are ignited either by the spark plug or by contact with hot burning gases left over from the initial cycle.

An alternative design of a pulse combustor is a so-called valveless combustor in which the mechanical valves are replaced with an aerodynamic diode in the form of a profiled orifice in the inlet pipe, contoured diffuser, or a shrouding duct [28] (Figure 20.15). Similarly to the combustor with mechanical valves, the high-temperature gases from a combustion chamber start to flow just after a

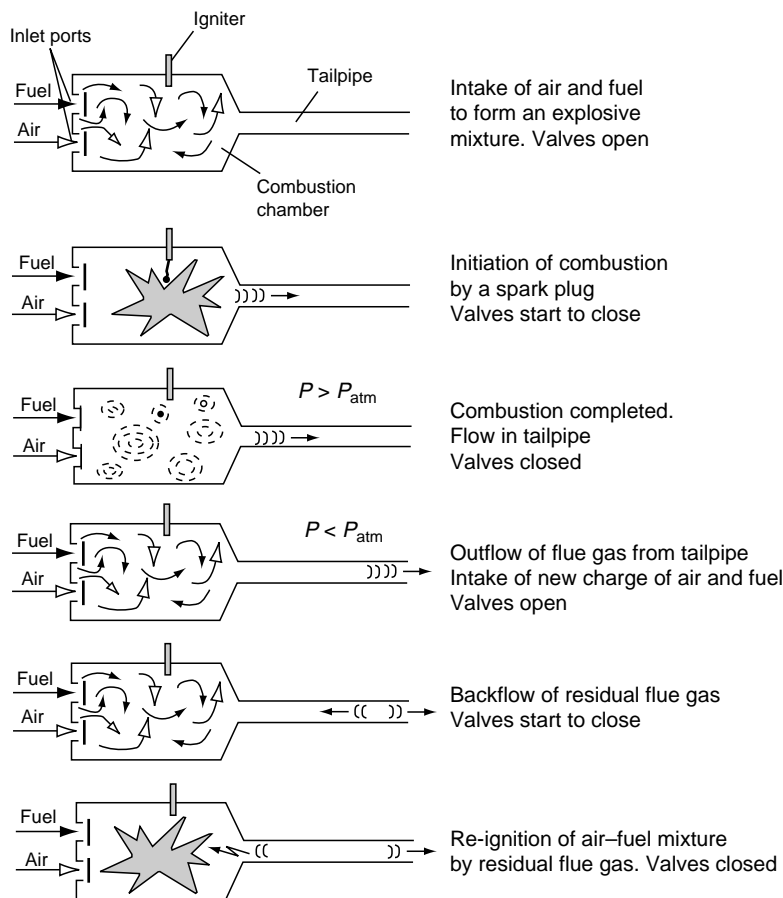


FIGURE 20.13 Operation of a pulse combustor.

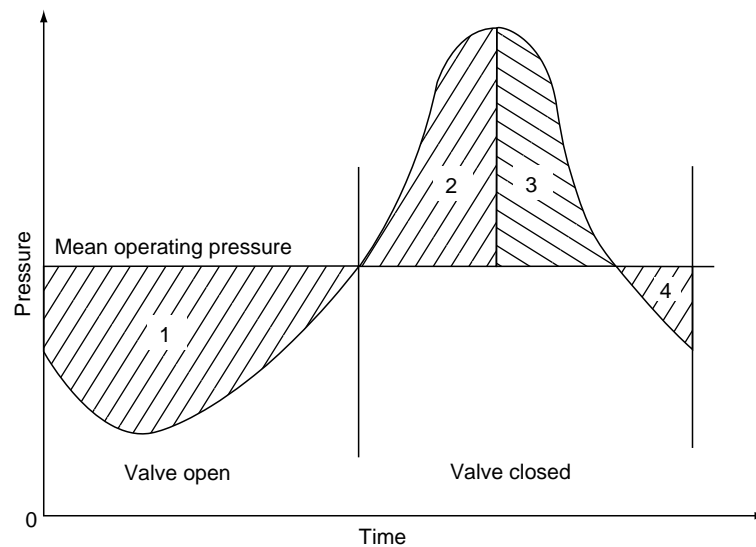


FIGURE 20.14 Pressure trace in a pulse combustor with rotary valve: 1, air and natural gas enter the combustion chamber; 2, fresh charge ignited, pressure rises as combustion gases heat up, stop of natural gas flow; 3, combustion complete, pressure after peak decreases as flue gases are vented; 4, momentum of exhausting gases creates negative pressure in the combustion chamber.

high-pressure period both forward into the tailpipe and backward into the inlet pipe. In a valveless combustor, however, this backflow of the flue gases during a half cycle of the pulsation dams up the flow of a fresh fuel–air mixture. After one quarter to one third of a combustion cycle, the gas flow in an inlet pipe reverses, which results in suction of the fuel–air mixture into the combustion chamber in which the process of ignition, combustion, and so on repeats itself.

Typically, pulse combustors oscillate with frequencies that vary from 20 to 150 Hz. Pressure oscillations in the combustion chamber of ± 10 kPa produce tailpipe velocity oscillations of nominally

± 100 m/s and the gas jet velocity at the tailpipe exit pulsates from approximately 0 to 100 m/s [27]. The input power for commercially available pulse combustors ranges from 70 to 1000 kW.

As applied to drying, the major function of the pulse combustor is to supply heat for moisture evaporation, and to generate large-amplitude, high-frequency pressure pulsations within a drying chamber, which enhances the drying rate [29]. The advantage of a strong oscillating gas jet from a tailpipe has also been applied in spray drying to promote liquid dispersion [25,27,30].

To excite large-amplitude pulsations in a drying chamber with a reasonable power input into the pulse

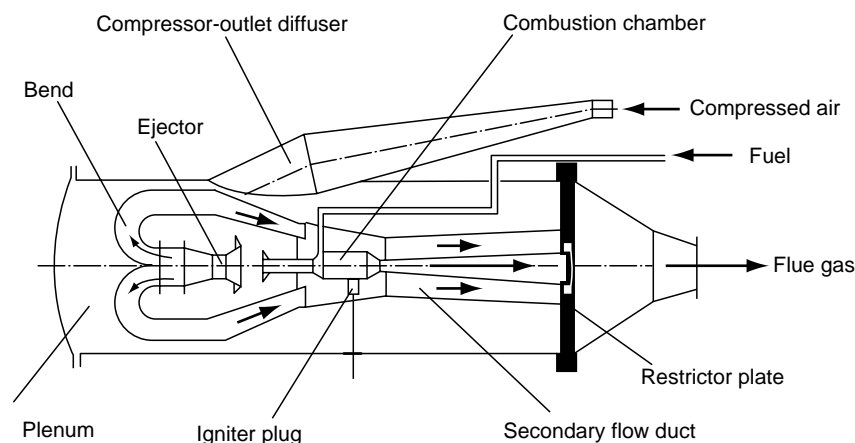


FIGURE 20.15 Valveless pulse combustor. (Courtesy of Kentfield, J.A.C., *Proceedings of the International Symposium on Pulsating Combustion*, Monterey, California, Paper A-6, 1991. With permission.)

combustor, the combustor must operate at a frequency that equals one of the acoustic frequencies characteristic of that chamber. Then, the pressure pulsations inside the combustor and the pressure pulsations generated inside a drying chamber are in resonance [31]. In practice, the acoustic mode of industrial dryers depends on the chamber geometry and material load as well as the operating conditions (e.g., temperature, moisture content, or air humidity). Consequently, the pulse combustor should be tuned actively to the actual resonant frequency of the chamber. Resonance-driven drying is achieved either by a trial-and-error method in a pilot-scale apparatus, or by varying the combustor frequency while monitoring the amplitude of pulsations inside the dryer by a pressure transducer.

In practical applications, the pulse combustor can be used alone or in combination with a conventional burner. The primary function of the pulse combustor is then to serve as a “speaker” that excites resonant pulsations in a drying chamber, thereby improving the dryer performance. In such a case, the pulse combustor supplies from 10 to 30% of the process energy.

The total amount of energy supplied to the burner and pulse combustor by a fuel is equal to or lower than the amount of energy supplied to the dryer with a conventional burner alone [26].

Figure 20.16 presents a block diagram of the pulse combustion spray dryer (PCD) in which the sound energy generated in a pulse combustor is used for atomization of the liquid feed and for enhancing the rate of convective drying. The liquid or pasty feed material is introduced into a gas stream from a pulse combustor in the feed chamber situated at the top of the drying chamber. The dryer and product-collection system is held under a slight negative pressure by an exhaust fan to eliminate possible leakage. The gas temperature at the dryer inlet is controlled by the fuel feed rate to the combustor, whereas the outlet temperature is adjusted by varying the material feed, which generally floats according to the evaporative load. Normal operating temperatures for the pulse combustor are in the range of 810 to 1470 K with a frequency range of 125 to 150 MHz. The maximum power output of one of the current units is 235 kW, which allows evaporation

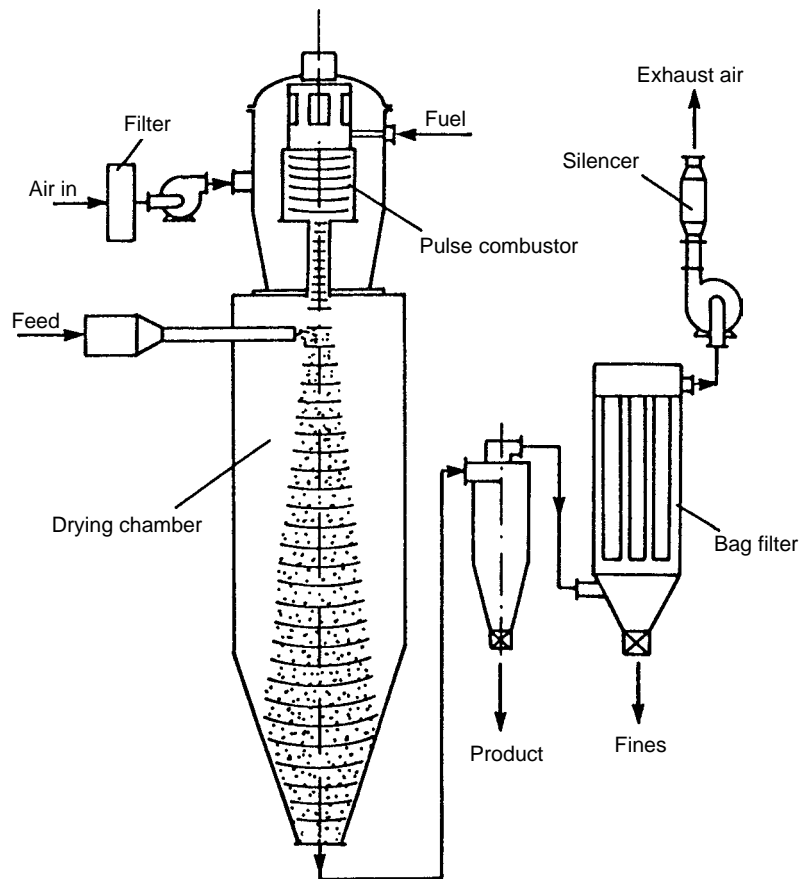


FIGURE 20.16 Pulse combustion drying system.

of up to 300 kg H₂O/h with drying air supplied at 600 kg/h [27]. The dryer can handle slurries (with solids up to 2 mm in diameter) with viscosities up to 16,000 cP as well as solutions with viscosities up to 300 cP. If desired, nozzles can be built in to atomize highly viscous liquids. The particle size distribution is generally finer than that normally obtained in a spray dryer using conventional nozzles, and varies from 30 to 60 μ m. The following list of materials dried in the Unison PCD compares well with the products from a spray dryer in regard to product quality and unit-cost basis [25]: animal antibiotics, citrus peel, corn syrup, spices, vegetables, protein, fibers, whole eggs, food colorings, caramel, biopesticides, spent brew yeast, acrylic latex, metallic sulfates, flame retardants, and so on. Performance data for selected products are given in Table 20.8.

Figure 20.17 is a schematic of a sewage sludge dryer with an afterburner and a pulse combustor

(developed by IMPULS—Foundation for Industrial Research, The Netherlands) operating at a frequency of 175 Hz, a sound pressure level of 136 dB, and a heat load of 2×10^8 W/m³. The sludge from a sieve belt containing about 30% solids is fed via a twin extruder to the upper part of a drying chamber in which the free-falling granules are exposed to heat supplied by an afterburner. The profiled baffles extend the residence time up to about 0.4 s, which is sufficient to remove the free water, thus preventing the granules from agglomerating. These granules trickle down over a number of mixing vanes while contacting the gas stream, which is a mixture of the flue gases from a tailpipe and the gas recirculated in the sewage plant. Approximately 50% of the moisture is evaporated in this section by the dryer. The remaining, mostly capillary water, is removed in a vibrated bed fluidized additionally by the pulse combustor flue gas. The

TABLE 20.8
Comparison of Unison Pulse Combustion Dryer versus Conventional Spray Dryer for Selected Groups of Materials

Material	Conditions	Spray Dryer	Pulse Dryer
Antibiotics	Inlet temperature, °C	400	704
	Outlet temperature, °C	127	104
	Product moisture content	Base	Same
	Atomization	Rotary	By pulse
	Air consumption	Base	20% less
Vegetables	Throughput	Base	Up to 11% higher
	Inlet temperature, °C	232	704
	Outlet temperature, °C	82	88
	Product moisture content	Base	Same
	Atomization	Nozzle	By pulse
Protein	Air consumption	Base	30% less
	Flavor	Base	2–3 times more flavorful
	Inlet temperature, °C	300	840
	Outlet temperature, °C	82	90
	Product moisture content	Base	Same
Acrylic latex	Atomization	Nozzle	Nozzle
	Air consumption	Base	25% less
	Protein level	Base	Same
	Inlet temperature, °C	204	760
	Outlet temperature, °C	71	77
Whole eggs	Product moisture content	Base	Same
	Atomization	Nozzle	By pulse
	Air consumption	Base	33% less
	Inlet air temperature, °C	260	871
	Outlet air temperature, °C	79	82
	Product moisture content	Base	Same
	Atomization	Nozzle	By pulse
	Air consumption	Base	40% less
	Protein level	Base	Same

Source: From Ozer, R.W., *Proceedings of the Powder and Bulk Solids Conference/Exhibition*, Chicago, Illinois, 1993.

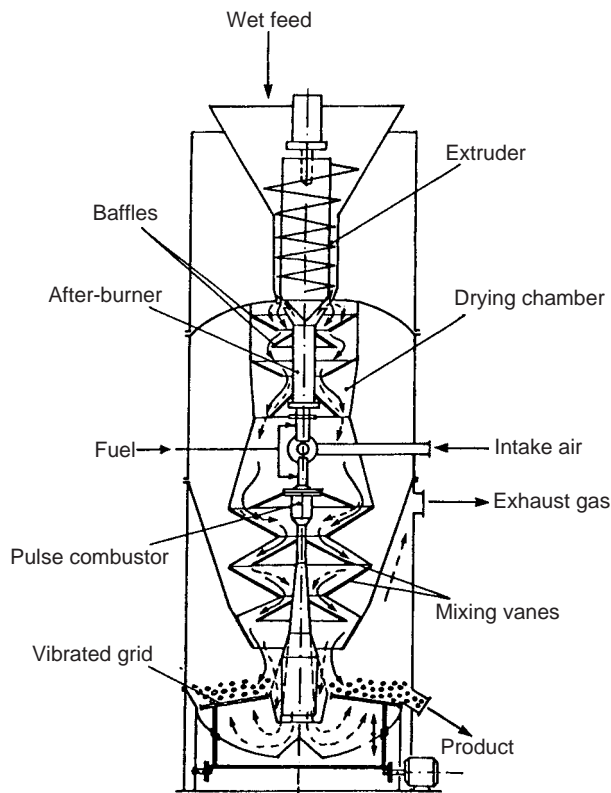


FIGURE 20.17 Schematics of pulse combustion dryer for sewage sludge.

final moisture content of about 10% db is achieved within 4 min, which, at the product temperature of 90 to 120°C, prevents excessive emission of CO, NO_x, and SO₂ due to product gasification. Granules produced (with diameters from 1 to 3 mm) are extremely porous with a specific weight of 0.4 as compared with the typical value of 0.7. A dryer 2 m in diameter and 6.5 m in height, equipped with a 1000-kW pulse combustor and 3000-kW afterburner, can handle 20,000 t/y of the sewage sludge [32,33].

20.2.6 DRYING WITH INDUCTION HEATING

The principle of induction heating is based on heat generated in the bulk of a conductive material due to eddy current (also termed Foucault current) (Figure 20.18). In a rotary dryer, for example, the dryer shell acts as the secondary winding of a ring transformer with laminated core (Figure 20.19). The exciting primary “winding” of this transformer is the center conductor in the form of a stationary bar or tube. The primary winding is fed from a step-down transformer at about 5 V/m of dryer length [34].

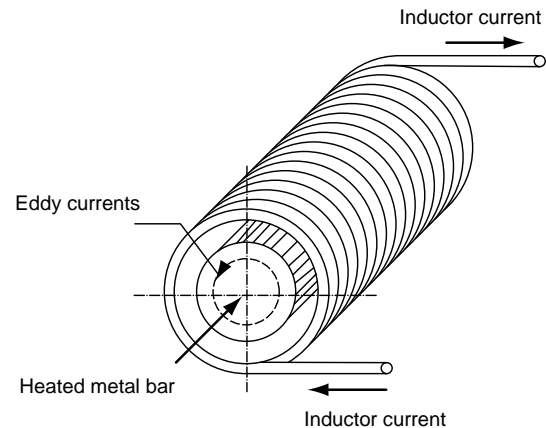


FIGURE 20.18 Principle of axial flux induction heating.

The drum itself is a composite structure consisting of a series of laminated ring cores situated inside a double-walled tubular shell. To maximize the heating effect, the electrical resistance of the inner shell is at least ten times that of the outer skin. In order to minimize heat losses, the dryer is internally or externally insulated (Figure 20.20). In the external mode of thermal insulation, the core material and the outer shell reach a temperature comparable with that of the inner shell; this means that the process temperature is limited by the thermal resistance of the construction materials or by the Curie point (670°C). The internal mode of insulation enables higher process temperature, even up to 1000°C within a very short time constant. However, in this configuration, the outer shell should be made of good conducting materials such as copper, brass, or aluminum. To enhance the heat transfer, the thin metal flights or fins may be fitted longitudinally to the inner surface of the shell.

This heating and drying technique, developed by EA Technology [35], was tested for drying of powders, crystals, granules, or slurries that are fragile or heat-sensitive. The energy consumption of the dryer depends on the type of the material and its moisture content, but in general ranges from 1.25 to 1.60 kWh/kg of water evaporated at the electrical efficiency exceeding 90%. The drying material occupies about 20% of the drum volume, thus, at the retention time of about 400 s, the output capacity varies from 0.55 to 3.64 m³/h.

20.2.7 CARVER–GREENFIELD PROCESS

Dewatering and drying of waste sludges is becoming an increasingly important area of concern globally. Most sludge is disposed of in landfills [36], but

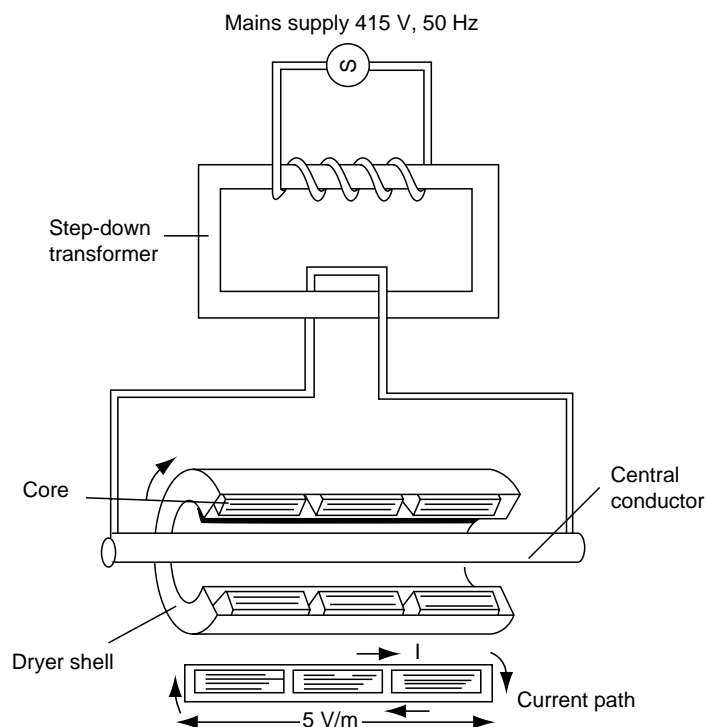


FIGURE 20.19 Principle of rotary dryer with axial current.

increasingly there is a trend toward utilizing the calorific value of the waste and burning it in boilers. In the late 1940s, an innovative dehydration process was proposed that involved use of a nonvolatile liquid medium as a carrier for dispersed wet solids to be effectively dried in multiple-effect evaporators. This is the so-called Carver–Greenfield (C–G) process. In multiple-effect evaporation, external heat is supplied to the last stage (effect) and the vapor produced there is used as a heat source to the preceding evaporator. The energy utilization can be further enhanced by mechanical vapor recompression (MVR) of vapor from an evaporator and then using it as the heating medium in the same evaporator [37]. [Figure 20.21](#); shows a schematic of a process configuration.

In the Carver–Greenfield process, first commercialized in 1961 to dry ground slaughterhouse waste, the ground meal is contained in indigenous carrier oil (tallow). This liquid mixture was processed in a two-stage, multiple-effect evaporator. The tallow and dried solids were separated by centrifuging. The tallow was recycled to maintain an oil-to-product ratio of 8:1. Some 60 plants of this type still operate in the United States. Pluenneke and Crumm [37] have reported on a highly efficient new version using a light oil that can be recovered by evaporation rather than by mechanical pressing. Light oils also extract oils present in the feed material (e.g., greases and fats

from sewage sludge, oils from peat and lignite drying, etc.). Further, an MVR system is included to obtain energy consumption of the order of 50 kJ/kg water evaporated.

In a recent version of the MVR C–G process, the feed is very dilute (2% solids) to quite high in solids content (50%). Drying of such sludges is accomplished in one stage. The solids' size must not exceed

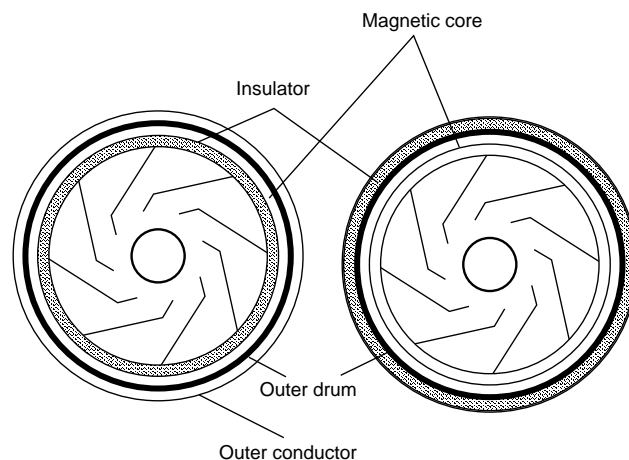


FIGURE 20.20 External and internal mode of thermal insulation.

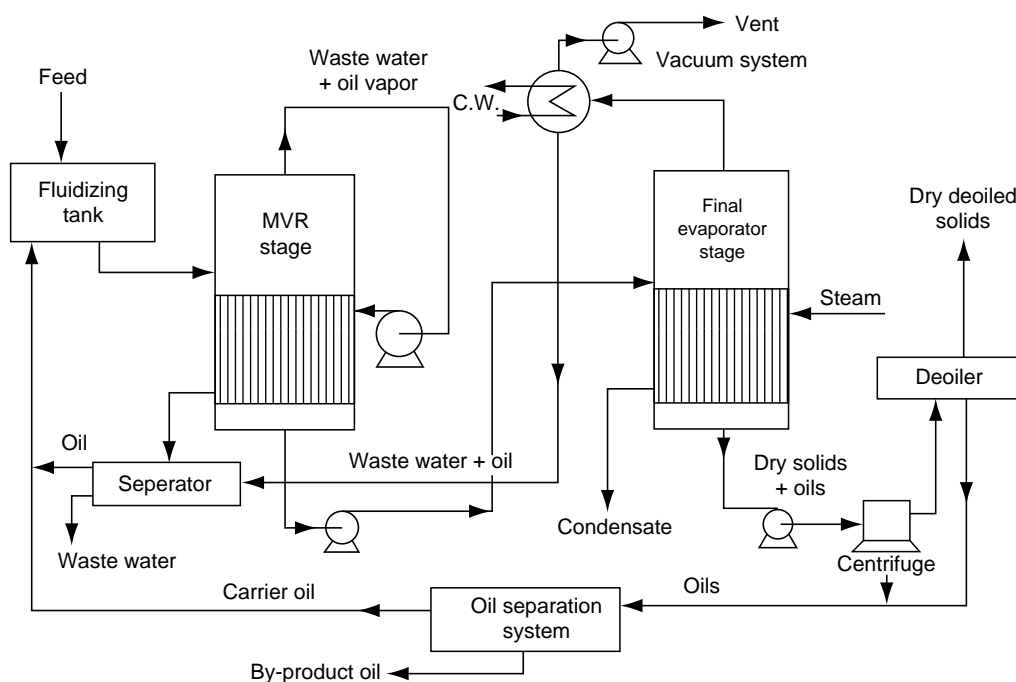


FIGURE 20.21 Carver–Greenfield process configuration.

7.5 mm along any axis. Existing plants do not typically use MVR yet. Falling-film evaporators can be used for low solids concentration, whereas forced flow is required at high concentrations. Up to five stages (effects) are used in various C–G installations to dry sewage sludge.

Among current applications of the C–G process are drying of activated sludge from wood pulp plant, wool processing waste from textile manufacturing facility, instant coffee processing waste, dairy product waste, sewage sludge, and others. Products like lignite and peat have been dried successfully at the pilot level with energy consumption in the order of 135 kJ/kg water evaporated. The process is particularly attractive for drying of peat. One may expect that this process and its variants will find new and major applications around the world as engineers and managers concerned with decision-making will become familiar with the technology and as experience accumulates on its industrial operations. Several major installations exist in Japan that use the C–G process for drying municipal waste sludges.

20.2.8 DRYING IN A VIBRO-ROTATIONAL BED

One of the major limitations of classical vibro-fluidized bed dryers is the potential for mechanical problems and increased energy consumption due to large inertia of the dryer, which must be accelerated.

An interesting design allowing vibration of the bed exclusively is the vibro-rotational bed (VRB) in which the oscillatory motion of particles is generated by rotation of a mechanical mixer of special design immersed in the bed.

The chamber of a VRB dryer is equipped with an agitator (Figure 20.22a) in the form of a horizontal disk with one or more inclined blades. While rotating, the blades not only force the bulk of particles to undergo a circulatory motion but also lift periodically a fraction of the bed that is just above the blade. This double action of the blades results in a characteristic pattern of particle circulation, viz: particles in the near-wall region rise in a spiral trajectory toward the bed surface, move to the bed center and then fall axially to the agitator region. When the rotational speed of an agitator increases the circumferential component of the particle velocity, the bed starts to oscillate in the vertical direction; the particle motion is then composed of a vertical oscillation and onward movement along the flow contour. The most intense oscillation occurs at agitator speeds in the range 3 to 7 1/s, depending on the size of the agitator [38]. The vertical circulation of particles with simultaneous vibration (as shown in Figure 20.22a) exists only if the ratio of the bed height to agitator diameter is less than 0.6 to 0.7. In deeper beds, particles in the upper layers do not circulate but oscillate locally in the vertical direction only. Air circulation above the bed surface

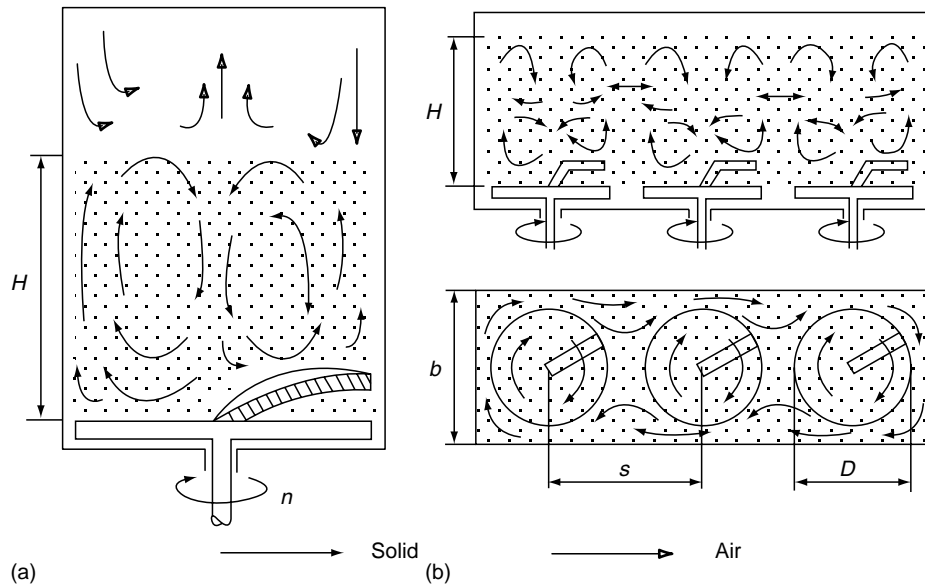


FIGURE 20.22 Flow pattern in a vibro-rotational bed: (a) cylindrical chamber with a single agitator; (b) rectangular chamber with a series of agitators.

(open arrows) caused by bed vibration is opposite in direction to the particle flow pattern mainly because of the suction-pumping effect.

Whereas the frequency of bed vibration is given by the product of the rotational speed of the agitator and the number of its blades ($f = nN$), the amplitude of vibration depends both on the bed height and the rotational speed of agitator, and it decreases with increasing height of the bed. For shallow beds ($H/D < 0.7$), however, the amplitude depends

solely on the rotational speed of the agitator with a maximum around $n = 5\text{--}7$ 1/s (Figure 20.23) [38].

Single VRBs can be arranged in series in a single rectangular vessel. The interaction of the particle flow pattern generated by the individual agitators corotating at the same rotational speed results in longitudinal transportation of particles via moving layers formed near the chamber wall. The thickness of these layers depend on the system geometry, and reaches a minimum of $0.1D$ if the chamber width is equal to the

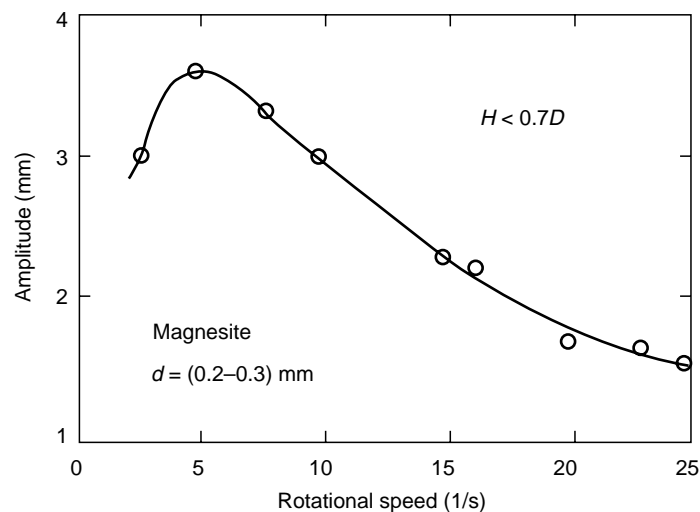


FIGURE 20.23 Amplitude of vibration in shallow vibro-rotational beds. (From Nikitin, V.S. and Puchkov, G.F., *Heat and Mass Transfer in Dispersed Systems*, Minsk, AN BSSR, 1983, pp. 116–123 (in Russian).)

agitator diameter. For $b/D > 1.4$ stagnation zones in the vicinity of the wall develop. The particle velocity in the near-wall layer depends greatly on the agitator speed, which in the order of 0.01 to 0.015 m/s at a rotational speed 6.5 1/s.

The longitudinal transportation of particles develops only for s/D (1.5–1.6) otherwise the entire VRB separates into individual ones with no interbed transfer of particles. A similar bed separation occurs at any speed with counterrotating agitators.

The vibration of the bed by mechanical mixing results in the transformation of electrical energy into heat due to interparticle and bed-to-wall/agitator friction (self-heating effect). The degree of energy conversion can reach 50%, which is approximately double that for the vibro-bed; the resulting bed temperature in the absence of the air bed-through flow may be as high as 60 to 80°C [39].

Intensive mixing of the bed and the continuous transfer of particles between the moving near-wall layers and cores of the vibrating beds obviously favors heat and mass transfer processes. The wall-to-bed heat transfer coefficient increases gradually with agitator speed, attaining an asymptotic value determined by particle characteristics. The heat transfer coefficient from submerged surface to the bed also increases with the agitator rpm. However, if particles are small enough to produce air flow and bed expansion due to the so-called “pumping effect,” the heat transfer coefficient exhibits a distinct maximum at a rotational speed characteristic of a fully developed vibratory motion (Figure 20.24) [40]. The absolute values of both heat transfer coefficients decrease

with increasing particle diameter, gas flow through the bed and bed height since all these parameters affect the contact of the heat transfer surface with moving particle.

Figure 20.25; presents immersed wall-to-bed heat transfer coefficients in vibro-rotational and vibrating beds of the same geometry and at the same amplitude–frequency of vibration [41]. The superiority of the VRB results from particle circulation superposed on bed vibration, which improves the contact of the particles with the heat transfer surface. An additional advantage of the VRB is its efficient operation under vacuum; the heat transfer coefficient at 0.2 atm is only 7 to 11% lower than that at atmospheric pressure [39].

Drying kinetics in a VRB are essentially similar to convective drying (Figure 20.26) because of air percolation through the bed, intense mixing due to vibration, and particle circulation. Hence, the drying rate is affected by air velocity, bed temperature, and rotational speed of the agitator. Self-heating of the bed, along with relatively high heat transfer rates at rarefied conditions makes the vibro-rotational dryer especially suitable for vacuum processing of thermolabile materials of low moisture content. Vacuum drying of nylon crumbs from 3 to 0.2% wb in a VRB, for example, reduces the drying time by 6 as compared to vacuum drying in a fixed bed [39].

Based on a broad spectrum of materials processed in laboratory- and pilot-scale dryers (millet, semolina, graphite, corundum, sand, nylon, polystyrene, polycarbonate, magnesite, alumina), the following features of the VRBs have been identified [38,42]:

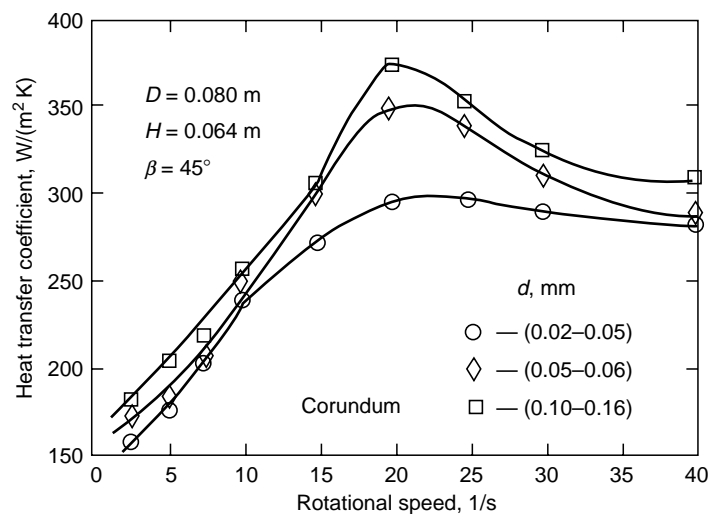


FIGURE 20.24 Immersed surface-to-bed heat transfer coefficient versus rotational speed of the agitator (10-mm flat bar inclined at 45°). (From Puchkov, G.F., *Heat and Mass Transfer in Dispersed Systems*, Minsk, AN BSSR, 1982, pp. 29–33 (in Russian).)

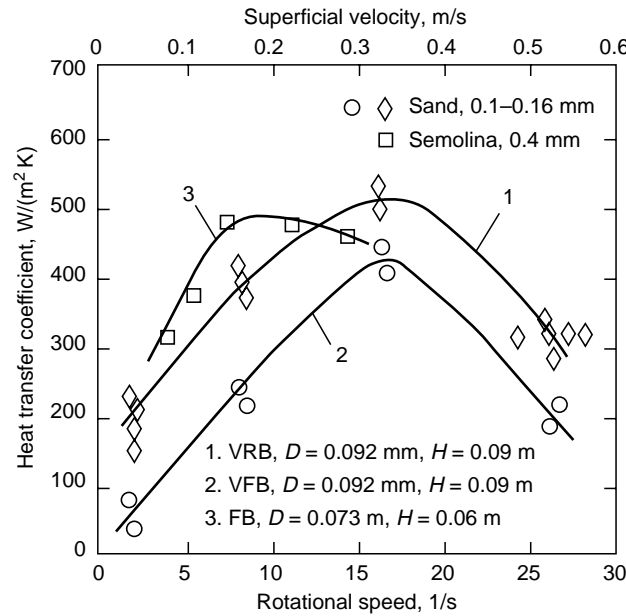


FIGURE 20.25 Wall-to-bed heat transfer coefficient in vibro-rotational bed (1), vibrated bed (2), and fluidized bed (3). (From Nikitin, V.S. and Martenov, O.G., *Heat and Mass Transfer in Dispersed Systems*, Minsk, AN BSSR, 1982, pp. 130–135 (in Russian).)

- Ability to process a wide range of highly poly-disperse particles (from 0.002 to 5 mm)
- No need of internal heater due to strong self-heating effect
- Isothermal conditions throughout the bed due to intensive mixing of particles
- Easy control of the bed temperature via internal cooler because of high heat transfer coefficients
- Efficient operation under vacuum

- No effect from moisture content on particle motion
- No vibrating elements

Table 20.9 presents the characteristics of the industrial VRB dryer in application to batch drying of polycarbonate beads. The cylindrical drying chamber is 0.55 m in diameter and 0.72 m high. The nominal static bed height is 0.36 m (62 kg of polycarbonate

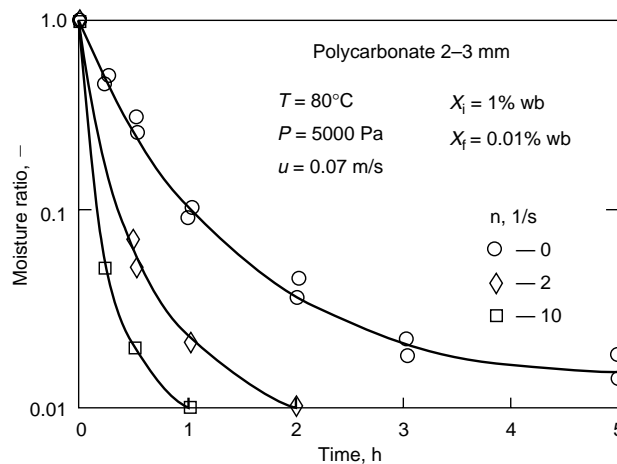


FIGURE 20.26 Drying kinetics in a vibro-rotational bed. (From Martenov, O.G. and Nikitin, V.S., *Transactions of the Academy of Sciences BSSR*, Minsk, AN BSSR, No. 2, 1983, pp. 69–74 (in Russian).)

TABLE 20.9
Performance Data for Drying of Polycarbonate Granules in the Vibro-Rotational Bed Dryer

Parameter	Value
Throughput	25 kg/h
Initial moisture content	0.2% wb
Final moisture content	0.015% wb
Bed volume	0.1 m ³
Pressure drop	<5000 Pa
Drying temperature ^a	Up to 140°C
Drying time ^a	Up to 2.5 h
Rotational speed of agitator	300 rpm
Installed power (3 × 380 V)	16 kW
Energy consumption per kg of dried material	0.04 kWh/kg
Overall dimensions	1.5 × 0.73 × 1.32 m

^aMaximum value depending on drying kinetics and initial and final moisture content.

beads) and the dynamic bed height is 0.42 m. A vacuum pump with controllable power evacuates evaporated moisture from the drying chamber to the condenser.

20.2.9 IMPINGING STREAM DRYERS

20.2.9.1 Introduction

Impinging stream dryers (ISDs) is a generic term for an emerging class of dryers in which moisture evaporation from wet particles or liquid droplets occurs in an impingement zone that develops as a result of “collision” of two oppositely directed high velocity gas streams, at least one of which contains the dispersed material to be dried (Figure 20.27). At the outset a distinction should be made from the well-known impingement dryers in which gas jets are directed onto the web- or slab-like materials, or “jet-zone” dryers in which a layer of particulates is “pseudofluidized” by a multiplicity of high-velocity airstreams [43]. Due to the hydrodynamic characteristics of the impingement zone, and the large inertia of the solid–liquid phase, the particles flowing originally with a gas stream oscillate about the impingement plane with damped amplitude until their velocity drops to the terminal velocity to be entrained with the outgoing gas stream. The high intensity of turbulence in the impingement zone and the rapid, unsteady particle motion enhance significantly the heat and mass transfer processes and hence reduce the drying times [44]. A typical set of operating data for impingement stream dryers in coaxial configuration is listed in Table 20.10.

TABLE 20.10
Range of Operating Parameters for Coaxial Impinging Stream Dryers

Gas velocity, m/s	10–150
Particle diameter, mm	0.2–3.0
Volumetric concentration of solids, m ³ /m ³	Up to 0.0035
Particle residence time, s	0.5–15.0
Production rate, kg/h	
Of dry material	50–2000
Of water evaporated	Up to 5000
Volumetric throughput, kg/(m ³ h)	
Of wet material	1,500–30,000
Evaporative capacity, kg/(m ³ h)	Up to 1700
Pressure drop, kPa	1–10

The main features of the impinging streams’ configuration that result directly from their special hydrodynamics are:

- High intensity of drying, especially when surface or weakly bound water is to be removed
- High product quality
- Simple design and operation (in general, no moving or rotating parts)
- Compactness
- Possibility of combining drying with other operations (e.g., granulation, disintegration, heating, cooling, chemical reactions, etc.)

Because of the high gas velocities, solids loading ratios, and momentum loss in the collision zone, the pressure loss in ISDs is much greater than in

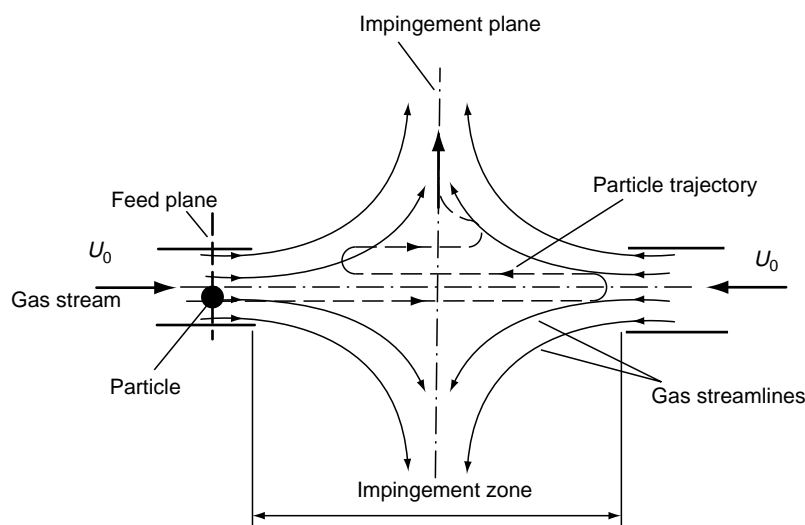


FIGURE 20.27 The principle of impinging stream dryers (ISDs).

TABLE 20.11
Performance of Selected Dryers for Granules of Aluminum Alloy^a

Parameter	Vibro-Fluidized Bed	Fluidized Bed	ISD
Drying time, ^b s	300–480	600	1–20
Unit gas consumption, ^c N m ³ /kg	8	13	1.2–2
Unit energy consumption, ^c kJ/kg	1380	3240	850
Product quality ^d	Decreased	Decreased	Unchanged

^aReference productivity, 250 kg/h; $d_p = 0.65\text{--}2\text{ mm}$; $X_1 = 0.17\text{ kg/kg}$.
^bTo reach $X_2 = 0.01\text{ kg/kg}$.
^cPer kg of dry solid.
^dEstimated from oxidation tests.

pneumatic dryers but it is comparable with that of fluidized and spouted bed dryers [44,45]. The impinging stream configurations can, however, compete in various aspects with the classical systems for drying of particulates and pastes (Table 20.11).

20.2.9.2 Characteristics of Impinging Stream Systems

When two gas streams collide, the initial velocity profiles characteristic of a free flow deform in the vicinity of the impingement plane and additional components of velocity (radial, axial, or circumferential, depending on the impinging streams' configuration) appear as a result of this deformation (Figure 20.28 and Figure 20.29).

If a solid or liquid particle flows with one of the impinging streams then in the impingement zone, the particle penetrates into the opposite stream due to its inertia and decelerates to a full stop some distance of penetration within the domain of the opposing jet flow (cf. Figure 20.27). Thereafter the particle accelerates in the opposite direction and penetrates the original gas stream. Thus, the process of deceleration and acceleration repeats itself. Because of energy dissipation, after several damped oscillations, the particle leaves the impingement zone and flows with an outlet gas stream into the discharge chamber.

Because of such oscillatory motion, the residence time of a single particle in the impingement zone is longer than that of the gas stream. This residence time is usually reduced for a number of particles because of interparticle collisions that lead to enhanced energy

dissipation. When solids are present in both gas streams, the rate of collision and the resulting loss of momentum is much higher than that for a single-stream feed, which might result in such significant decrease of the residence time and penetration depth that the beneficial effects of impingement are dramatically reduced.

The above disadvantage is practically eliminated in an impinging stream system with mobile impingement zones [46]. In this arrangement, the impingement plane is made to move between the two locations I and II by alternate switching of the gas flows from left to right and then from right to left (Figure 20.30). The particles fed into the accelerating flow duct pass through the central (reverse flow) tube that links the two impingement chambers. Flowing with the original gas stream into the first impingement chamber the particles collide with the secondary gas stream, penetrate it up to the stagnation point, and then begin to accelerate in the opposite direction. At this moment the gas outlet of the first impingement chamber is closed whereas the outlet of the second impingement chamber is open. This procedure results in the flow of a secondary gas stream with accelerating particles toward the second impingement chamber where the process of jet collision, penetration of particles into the original gas stream, and the following acceleration toward the first impingement zone is repeated. The period of oscillatory motion of the particles is controlled by decreasing the inertia force due to reduction of particle size in the course of processing, by reduction of particle mass due to moisture evaporation, or by placing a suitable limiting grid at the outlet ducts, which restricts particle entrainment.

There are another configurations of IS such as a semicircular IS, two- and four-IS, and a variety of multistage and combined systems [3,47]. One of such systems that combines coaxial IS and curvilinear IS and allows particle segregation and recycling of over-size (wet) particles into impinging streams is shown in Figure 20.31.

20.2.9.3 Heat Transfer and Drying Characteristics

Due to the inherent hydrodynamics of ISDs the heat transfer rates from gas to particle depend on time as well as space. Reported data indicate, however, that the average heat transfer coefficient for ISD differs by less than 10% from its local value, which is sufficient for dryer design [48,49].

In general, the average gas-to-particle heat transfer coefficients for ISDs are much higher than those in classical dryers that operate under similar hydrodynamic regimes. For example, when the heat

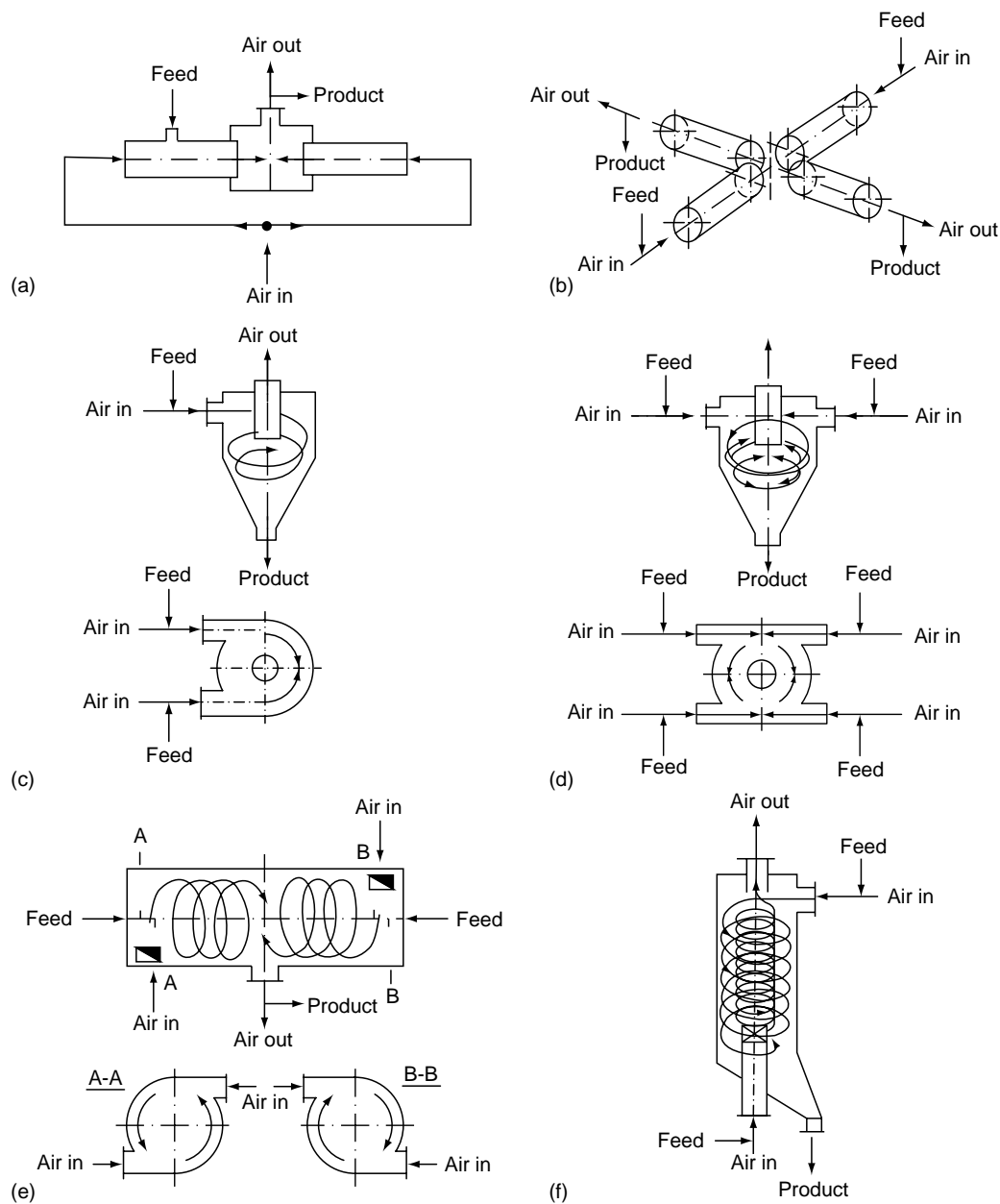


FIGURE 20.28 Basic types of impinging stream flows: (a) coaxial; (b) mutually perpendicular X configuration; (c) curvilinear countercurrent; (d) four impinging streams; (e) counterrotating countercurrent; (f) corotating countercurrent.

transfer coefficient in a coaxial ISD is $850 \text{ W}/(\text{m}^2 \text{ K})$, the values for pneumatic dryers calculated for the same operating conditions from several equations compiled by Strumillo and Kudra [50] ranged from 300 to $520 \text{ W}/(\text{m}^2 \text{ K})$. For the TIS dryer with coaxial impinging streams the heat transfer coefficient is 1.1 to 1.8 times higher than that for the geometrically similar spouted bed under the same operating conditions [51]. The volumetric heat transfer coefficients for ISDs are also higher, reaching $125,000 \text{ W}/(\text{m}^3 \text{ K})$, which is 2.5 to 3 times higher than the values for

spouted bed dryers [51] and 15 to 100 times higher than those for spray dryers [44].

Drying in impinging streams is purely convective. Thus, typically both constant and falling drying rate periods can be observed if the processing materials have both internal and external resistances to mass flow. Generally, the period of constant drying rate is short because of the high heat and mass transfer rates.

Figure 20.32 shows the time variation of moisture content, temperature, and location for 1-mm aluminum beads covered with a thin layer of surface water,

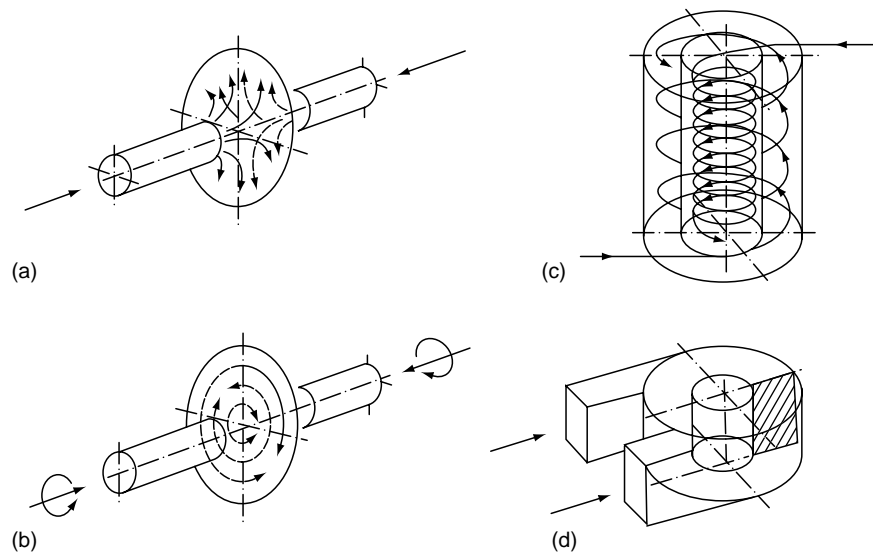


FIGURE 20.29 Impinging plane geometries: (a) planar with radial flow; (b) planar with circumferential flow; (c) tubular; (d) annular.

and dried in an ISD with mobile impingement zones (see Figure 20.30) [52]. It can be seen that the heat transfer rate differs significantly in the period of particle deceleration and acceleration due to the different gas-particle relative velocities. Despite the stabilizing effect of oscillatory motion this difference can be as high as 35 to 40%. Because of surface water evaporation, the drying rate is controlled by the rate of external heat transfer. Thus it is higher in the deceleration period than in the acceleration period. The

particles are completely dry within 0.6 s, that is, within less than one cycle of oscillation.

Figure 20.33 presents the drying kinetics of crystalline lysine with an initial moisture content of 15.2% in an ISD with mobile impingement zones using hot air at 120°C and flow velocity of 20 to 23 m/s; the frequency of reversing motion is 1.0 to 1.2 Hz [53]. Curve 1 represents drying of monodisperse crystals (0.4-mm mean diameter) at mass concentrations of 0.2 to 0.5 kg/kg of air. In this case the surface water is

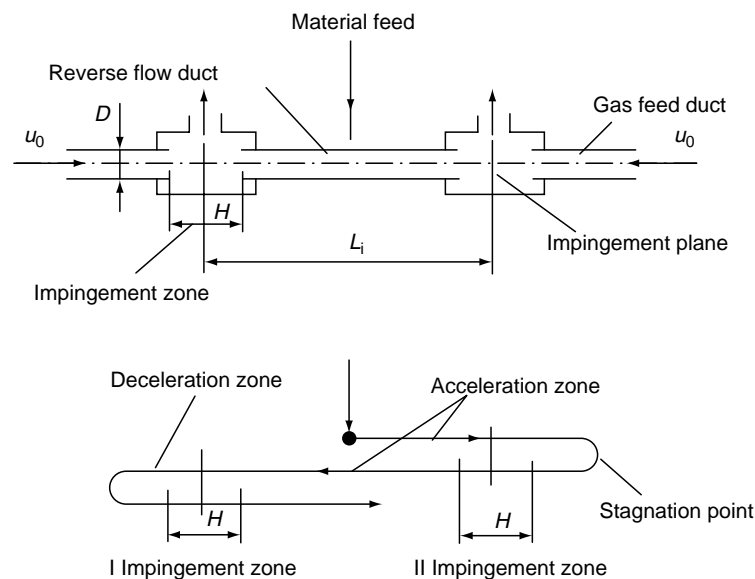


FIGURE 20.30 Coaxial impinging stream dryer with a mobile impingement zone.

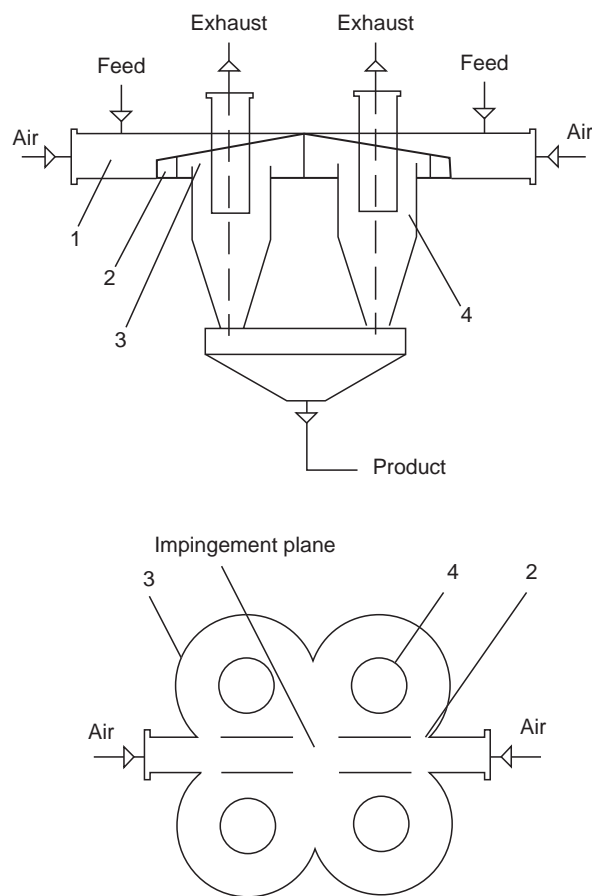


FIGURE 20.31 Combined impinging stream dryer. (From Fiodorov, G.S. and Shulyak, V.A., Russian Patent 1,768,897, 1992.): 1, accelerating duct; 2, slot; 3, swirling chamber; 4, cyclones.

removed within one period of motion (2 to 3 s). The desired moisture content level of 1% can be achieved within five periods of oscillation. Curve 2 shows the drying kinetics for polydisperse lysine with a mean diameter ($d = \sum x_i d_i$) equal to 1 mm with simultaneous grinding due to the oscillatory motion of the inert material (mixture in 1:1 mass ratio of 2 mm steel and 3 mm aluminum beads) at a mass concentration of 1.0 to 1.5 kg of inert per kg of drying material. Compared with curve 1 the drying rate with simultaneous disintegration is appreciably higher in the first period of reverse motion mainly due to a reduction of the heat transfer resistance inside the material and increase of the mass transfer area due to particle disintegration. After a certain number of oscillations the drying rate decreases and follows a similar decrease in the intensity of crystal disintegration (curve 3).

Empirical correlations for heat transfer coefficients, pressure drop, penetration depth, residence time, scale-up rules as well as results of mathematical modeling and computer simulation that are necessary for process calculation and equipment design can be found elsewhere [3,54,55].

20.2.9.4 Some Industrial Impinging Stream Dryers

Figure 20.34 shows a schematic of an industrial ISD for drying of dewatered sewage sludge in a two-stage configuration. The first stage is a coaxial ISD in which the feed is dispersed and a significant amount of surface moisture removed. The second stage is comprised of flash drying in the pneumatic duct and

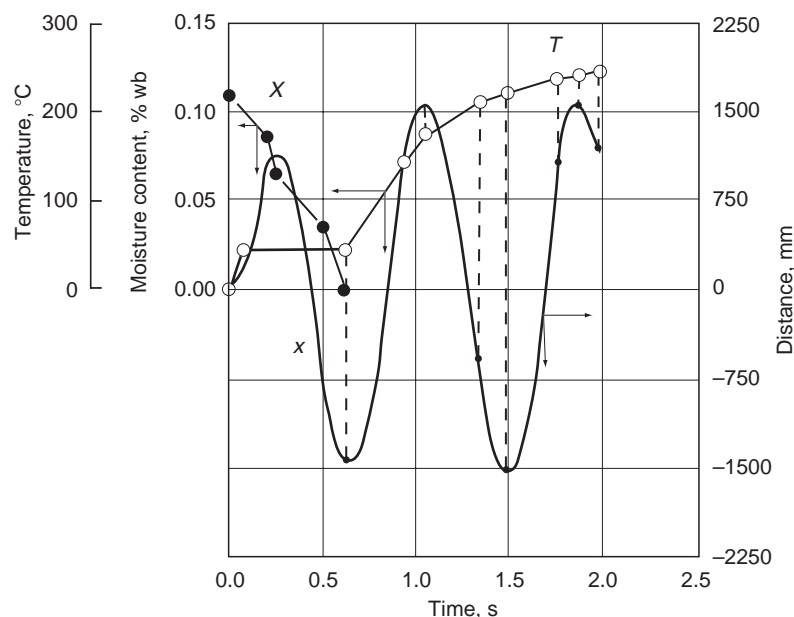


FIGURE 20.32 Time variation of axial position, moisture content, and particle temperature in an impinging stream dryer with a mobile impingement zone ($T_{g1} = 250^\circ\text{C}$; $u_0 = 30$ m/s; $X_1 = 0.11$ kg/kg).

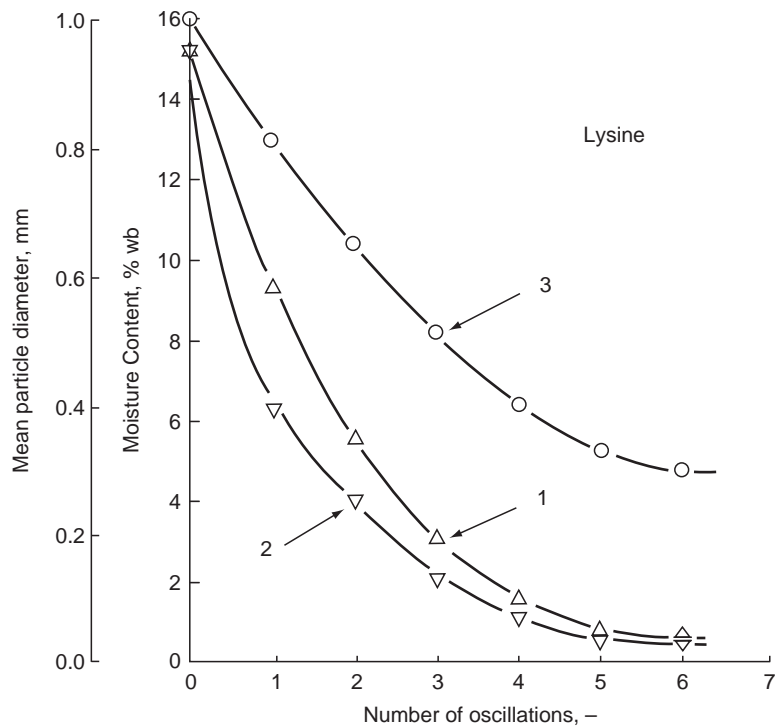


FIGURE 20.33 Drying and grinding characteristics for lysine: 1, drying kinetics for monodisperse crystals; 2, drying kinetics for polydisperse crystals; 3, grinding curve for polydisperse crystals.

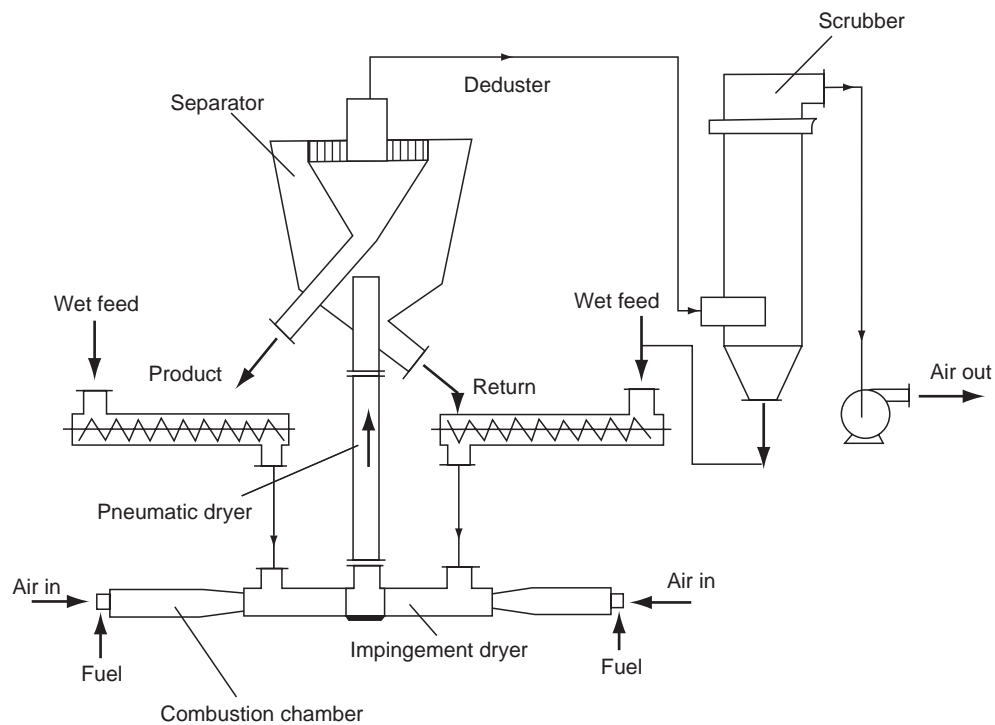


FIGURE 20.34 Impinging stream dryer for waste sludges.

drying in a swirling stream during segregation in the centrifugal separator. The mixture of air and furnace gases from combustion of liquid or gaseous fuels (oil or natural gas) is used as the drying agent at inlet temperatures of 560 to 700°C. The outlet gas temperature is in the range of 90 to 140°C. The gas velocity in ISD tubes and the pneumatic duct is in the order of 25 to 50 m/s, which ensures stable transport of the dispersed material without settling or sticking to the tube walls. The gas velocity in the profiled nozzle of the ejector-type feed is typically 200 to 250 m/s, which is sufficient to discharge a screw feeder, disintegrate lumps, and disperse particles well in the feed section. The mechanically pre-dewatered sludge at initial moisture content 72 to 83% wb is blended in the ratio 3:1 with under- and oversize solids from the centrifugal separator. This backmixing with the initial sludge allows the moisture content to be adjusted to the feed moisture content to about 50%, which facilitates operation of the ISD (lower gas velocities can be employed with a relatively dry feed) and also reduces the energy consumption from 3300 to 3800 kJ/kg to about 2900 kJ/kg evaporated water. The specific air consumption per kg of water evaporated is 4.5 to 5 Nm³, whereas the electric energy consumption varies from 0.02 to 0.03 kWh/kg. The final moisture content depends on the particle size in product, and ranges from 2% for the 0.25-mm particles to 55% wb for 5-mm particles. The average moisture content of the 1 to 2-mm granules is 19.5% wb.

Coaxial ISDs with mobile impingement zone as presented in Figure 20.30 are especially suitable for drying quartz and foundry sand, metal granules, polymers, and like materials with surface or loosely bound moisture. Table 20.12 gives the characteristics of an industrial unit used for drying of metal granules. The same type of ISD with mobile impingement zones but furnished additionally with constraining

grids at the outlets from the impingement chambers has been used in the former Soviet Union for drying with simultaneous grinding of otherwise hard-to-dry products containing tightly bound moisture.

An example of such a product is crystalline lysine, a highly heat-sensitive material with tendency to lump and with a large amount of bound water (up to 17% wb). The crystal size after centrifugation is 0.2 to 2.5 mm. These conditions along with the additional requirements for final moisture content below 1 to 1.5% wb with crystal dimension to be less than 0.5 to 0.6 mm are met by the ISD dryer shown in Figure 20.35. The dryer consists of air conveying ducts with two impingement chambers situated at the ends of a central drying duct 1.5 m in length. Both chambers are connected via outlet ducts and cyclones to a flow-switching device. The outlet ducts are constrained by grids with a grid opening of 1.3 mm. A mixture of steel (2 mm) and aluminum (3 mm) beads in 1:1 mass ratio is used to disintegrate the crystals and break lumps. The mass concentration of the inert beads in the drying duct is maintained at 1 to 1.5 kg/kg. The frequency of reverse motion between the impingement chambers is 1 to 1.2 cps. Hot air at 135°C and 20 to 23 m/s enters both ends of the impingement chambers. A switching device moves the impingement zone periodically between the two impingement chambers. This results in an oscillatory motion of both the inert beads and the wet particles, which are continuously fed to one end of the duct only. After performing four to five oscillations, which are the equivalent of 10 to 15 s of drying time, the material is well ground and dry (see Figure 20.33) so it is carried away through the constraining grids to the cyclones. The volumetric evaporative capacity of the dryer is about 1700-kg H₂O/(m³ h) whereas air and heat consumption (without heat recovery) per kg of water evaporated are 200 to 250 Nm³ and 3 to 3.6 MJ, respectively. This shows that the performance of the ISD is superior to that of a fluidized bed dryer [53].

Figure 20.36 presents an industrial setup with semicircular impinging stream ducts developed for thermal processing of grains (e.g., drying, puffing, and carrying out certain thermally induced biochemical reactions). In this case superheated steam is used as the carrier and drying medium. A multiplicity of semicircular tubes (type a ISD as in Figure 20.28) are placed within a cylindrical chamber with a conical bottom. The product is collected at the bottom of the cone whereas the exhaust gas is led to a cyclone to collect the fines.

Exhaust steam is partially recycled after mixing with fresh superheated steam. Grains treated in this way are suitable for direct use as cattle feed with high digestibility. Table 20.13 presents the operating

TABLE 20.12
Performance of an Impinging Stream Dryer^a
for 0.5–2.5-mm Aluminum Alloy Granules

Production rate, kg wet material/h	150–200
Moisture content, % wb	
Initial	10–12
Final	0.01–0.02
Inlet air temperature, °C	300
Drying time, s	5–20
Air consumption, N m ³ /h	250–300
Electric power, kW	70
Overall dimensions, m	3 × 2.2 × 3.5

^aModel USV produced by ITMO, Minsk, Belorussia.

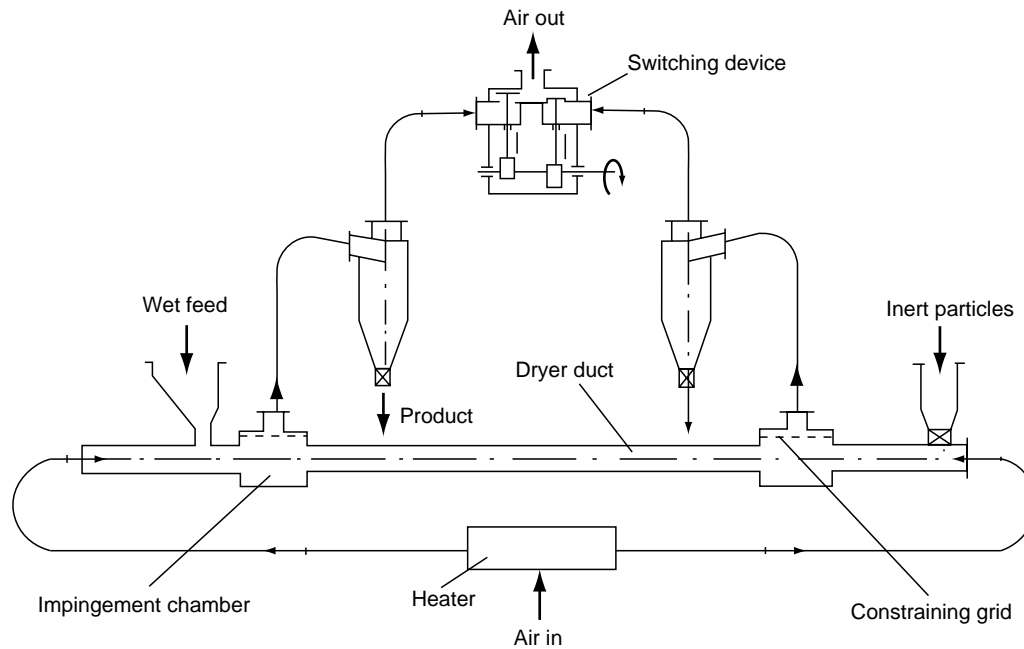


FIGURE 20.35 Impinging stream dryer with mobile impingement zone for drying of crystalline lysine.

characteristics of such ISD installations used for thermal processing of grains.

Figure 20.37 shows a schematic of a vortex spray dryer (VSD) for suspensions and slurries that operates essentially as a countercurrent, counterrotating ISD. The dryer is made as a horizontal cylinder 1.2 m in diameter and 4 m in length with two fluid nozzles

located at each end of the dryer. The primary streams of hot air are introduced axially with the nozzles. Secondary airstreams are fed tangentially to the primary ones but in opposite directions, which results in a counterrotation of the primary airstreams. Thus, the drying material in the form of liquid or suspension is not only atomized but also brought into a swirling

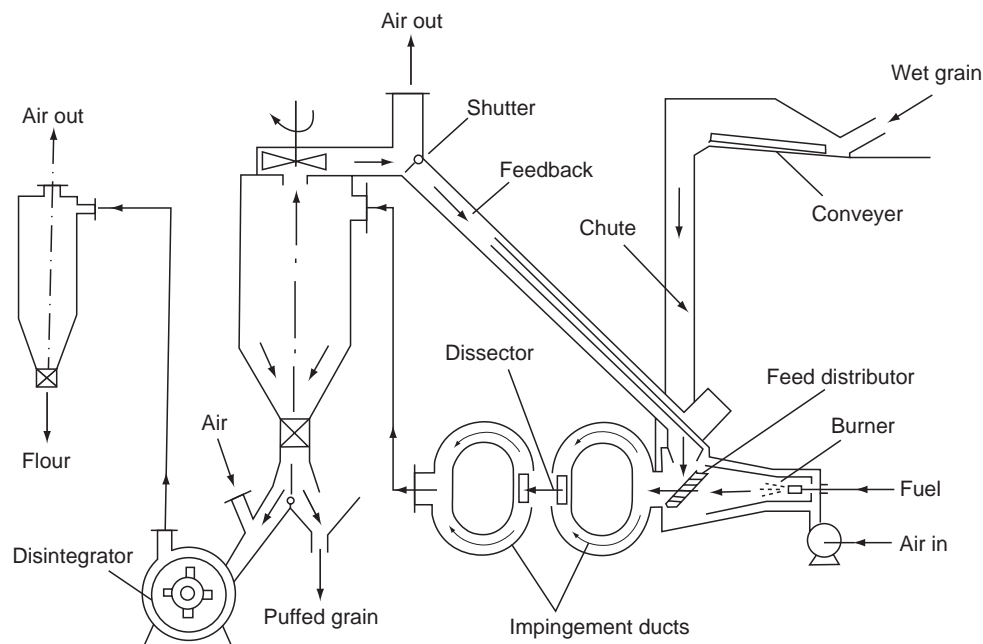


FIGURE 20.36 Semicircular impinging stream dryer (Vortex spray dryer, 0.8/1.2 type) for thermal processing of grains in superheated steam.

TABLE 20.13
Semicircular Impinging Stream Dryers for Drying/Puffing of Grains

Parameter	Vortex Spray dryer, 0.15 ^a	TD, 3 ^a	TRD, 3 ^b
System	Closed ^c	Closed ^c	Open
Throughput (dry product), kg/h	260	1000	1200
Material ^d			
Initial moisture content, % wb	5–25	12–16	<35
Final moisture content, % wb	3–15	6–12	6–12
Degree of dextrinization, %	30–45	25–40	35–45
Heat carrier	Air	Superheated steam	Flue gas
Inlet temperature, °C	<380	380–420	<420
Consumption, kg/h	400	180 (at 130°C)	30 (fuel oil)
Electric power (total/heater), kW	43/26	70/81	85/103
Energy consumption, kJ/kg dry product	455	252	310–370
Pressure drop, kPa	5	10	5
Overall dimensions, m	2.4 × 1.7 × 2.7	2.4 × 1.7 × 3.0 ^e	3.5 × 2.0 × 1.0

^aProduced by Kislodmash, Odessa, Ukraine.

^bProduced by ITMO, Minsk, Belorussia.

^cWith release of excessive vapor.

^dCorn, rye, barley, wheat, rice.

^eWithout steam heater.

motion by the rotating airstreams. To prevent falling of the sprayed droplets onto the dryer wall (which can result in material overheating), supplementary airstreams are introduced tangentially (in the same direction as the secondary air) through the dryer wall. This part of the dryer wall at both ends is made as a slotted grid formed of suitably directed vanes.

After impingement in the middle zone of the dryer, the airstream carrying dry, powdery material is directed to the separation system. Due to intensive evaporation in the impingement zone the inlet air temperature of about 150°C falls rapidly to about 70°C, which allows drying of thermosensitive materials like antibiotics or microorganisms. At this inlet air temperature

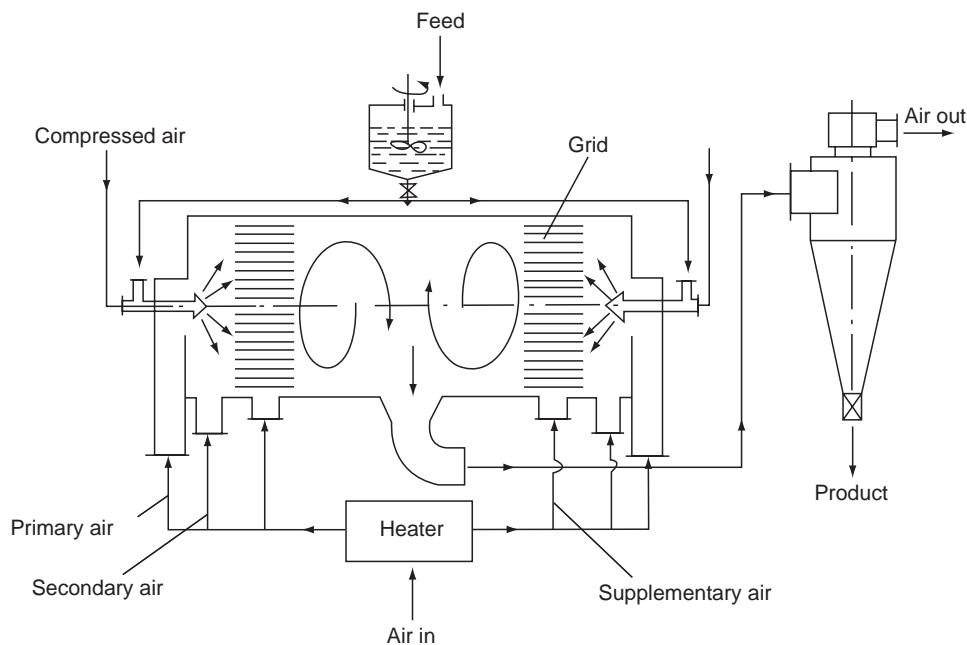


FIGURE 20.37 Counterrotational, countercurrent impinging stream dryer for liquids, slurries, and pastes.

the evaporative capacity is in the order of 28-kg H₂O/(m³ h), which gives about 100 kg of evaporated water per hour. As compared to the classical spray dryers operating under similar conditions, the impinging stream spray dryer has seven times greater evaporative capacity (Table 20.14), mostly because of the high turbulence in the impingement zone and higher loading of the air.

20.3 NOVEL DRYERS

20.3.1 PULSED FLUID BED DRYERS

Pulsed fluid beds (PFBs) are a variant of the fluidized bed technique in which the fluidizing gas is introduced in a periodic (pulsed) manner. In a large rectangular bed, the region of pulsed fluidization can be alternated within some advantages if periodic relocation of the gas stream is used. Although the idea of gas “relocation” to eliminate channeling in a fluidized bed of fine particles was originated over 30 years ago [56], it did not find a wide application.

Figure 20.38 shows the principle of a PFB dryer with periodically relocated gas stream. Hot air flows through a rotary valve-distributor that periodically interrupts the airstream and directs it to different sections of the plenum chamber below the supporting grid of a conventional fluidized bed. Air in the “active” chamber fluidizes that segment of the bed located above the active chamber. This fluidized segment of the bed becomes almost stagnant when the air is directed to the next chamber. In practice, because of gas

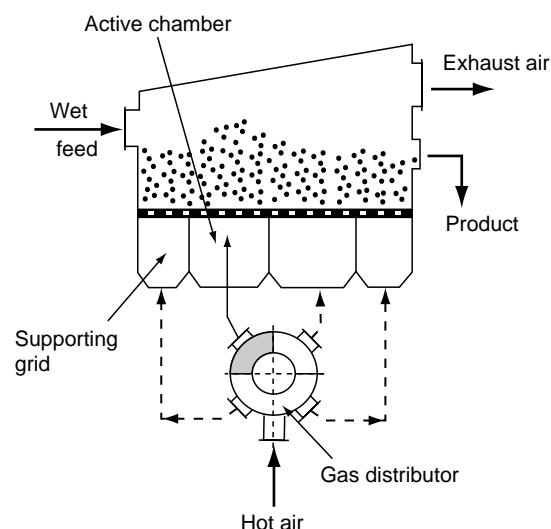


FIGURE 20.38 Principle of a pulsed fluid bed with periodically relocated gas stream.

compressibility and bed inertia, the entire bed is fairly fluidized with vigorous fluidization in the active zone. The advantage of the rotary-type valves over the on-off ones is that the sinusoidal variation of the gas flow makes relocation of the active bed more smooth. A modification of the PFB with relocated gas stream is a spouted bed configuration in which the gas stream from a rotary valve-distributor activates periodically the bed of particles in a tapered chamber with rectangular cross-sectional area [57].

As compared with conventional fluid bed dryers, the PFB with periodically relocated gas stream offers the following advantages [57,58]:

- Good fluidization of even large and anisotropic particles (e.g., sliced vegetables 20–30 mm in diameter and 1.5–3.5 mm thick)
- Reduced pressure drop (by 7–12%) due to the lower minimum pulsed fluidization velocity and larger opening area of the supporting grid
- Lower minimum fluidization velocity (by 8–25%)
- Improved bed structure (no channeling, better particle mixing)
- Operation with shallow beds
- Energy savings up to 50%

The principal operating parameters of a PFB dryer are:

Bed height	0.1–0.4 m
Gas velocity	0.3–1.8 m/s (depending on the particle characteristics)
Pressure drop	300–1800 Pa
Frequency of gas pulsation	4–16 Hz

TABLE 20.14

Performance Characteristics of an Impinging Stream Spray Dryer and Classical Spray Dryer

Parameter	VSD-100 ^a	ZT-100 ^b
Evaporation capacity, kg H ₂ O/h	96	80
Air consumption, N m ³ /h	3500	3000
Air temperature, °C		
Inlet	148	150
Outlet	74	70
Moisture content, % wb		
Initial	94	98.5
Final	7.0	11.8
Dryer dimensions, m		
Diameter	1.2	3.3
Length	4.0	3.5
Volumetric evaporation capacity, kg/(m ³ h)	20	2.62
Volumetric heat transfer coefficient, W/(m ³ K)	232	33.6

^aVSD-100, ITMO, Minsk, Belorussia.

^bZT-100, Germany.

TABLE 20.15**Comparison of Some Industrial Band Dryers for Vegetables with the Pulsed Fluid Bed Dryer**

Index ^a	SPOMASZ (Poland)	SP-66 (Yugoslavia)	41-A (Bulgaria)	PFB (Poland)
Unit steam consumption, kg/kg	23.33	10.0	13.33	9.33
Unit consumption of electric energy, kWh/kg	0.56	0.55	0.49	0.83
Volumetric throughput, kg/(m ³ h)	0.149	0.167	0.115	0.232
Unit steel consumption, kg/(kg h)	500	500	670	56
Quality index ^b	0.5	0.5	0.0	0.8

^aPer kg of dry product.^bFraction of the 1st class product.Source: From Glaser, R., *Proceedings of the VII Drying Symposium*, Lodz, Poland, 1991, pp. 147–154 (in Polish).

PFB dryers have been successfully used for drying of grains and seeds (e.g., peas, beans), sliced and diced vegetables, sugar-infused cranberries, as well as for drying of crystalline and powdery materials such as

sugar, thiohexame, calcium gluconicum, pentaerythritol, and so on [57,59–64]. Comparison of a PFB dryer with the conventional band dryers for sliced vegetables is given in Table 20.15, while Table 20.16

TABLE 20.16**Comparison of Selected Industrial Dryer-Coolers for Sugar (Throughput 20.8 t/h)**

Parameter	Rotary ^a	VFB I ^b	VFB II ^c	PFB ^d
Volumetric throughput, kg/m ³ h	265	495	862	584
Air velocity, m/s				
Dryer	0.22	0.83	1.17	1.1
Cooler	0.17	0.82	1.17	0.5
Inlet air temperature, °C				
Dryer	90	70	105	80
Cooler	—	—	10	15
Inlet moisture content, % wb				
Dryer	1.5	2.0	1.5	1.0
Cooler	0.03	0.04	0.04	0.04
Final sugar temperature, °C	25–38	(<i>T</i> + 15) ^e	34–46	34–38
Unit heat consumption, kJ/kg				
Of dry sugar	122.7	111.9	145.1	88.2
Of water evaporated	12,650	11,548	14,950	9,080
Unit air consumption, N m ³ /kg sugar	2.32	2.28	2.20	1.52
Unit steam consumption, kg/kg sugar	0.058	0.053	0.068	0.041
Attrition index ^f	0.25	0.06	0.04–0.06	0.06
Relative production cost	1.5	1.3	1.4	1.0
Power consumption, kW	132	206.5	162.8	86.3
Grid cross-sectional area, m ²	85.0 ^g	19.55	17.7	13.55
Dryer-cooler volume, m ³	78.7	42.0	24.2	35.4
Total mass, kg	36,000	6,000	6,000	10,200

^aLicense of Roto-Louvre.^bVibro-fluidized bed dryer (IIC PL).^cVibro-fluidized bed dryer (Swidnica).^dPulsed fluid bed dryer (AE Wroclaw).^eAmbient temperature.^f $1 - d_g/d_{e0}$.^gDrum surface area defined by louvers' peripherals.Source: From Glaser, R., *Scientific Papers of Wroclaw Academy of Economy*, 490, 81, 1989 (in Polish).

compares the PFB dryer for crystalline sugar with the rotary and VFB dryers.

20.3.2 MECHANICALLY FLUIDIZED BED DRYERS

Pasty solids that do not fluidize well can be dried when mechanically dispersed in a hot gas stream. In such a dryer, the air pressure drop is much lower than in a conventional fluidized bed. A well-known commercial dryer of this type is the U-Max dryer (Figure 20.39), which expands the uses of the mechanically induced pulsed or fluidized bed beyond mixing to include drying that may follow mixing, reaction, milling, and agglomeration. The fluid bed is generated mainly by the rotation of a shaft within the confines of a horizontal cylinder. A permanently affixed arm with plow-shaped agitating elements extend from this shaft. Rotation of the shaft causes the material to be lifted by the mixing elements and thrust into a free space in a pulsing or fluidized bed action with simultaneous axial and radial motion within the cylinder. The dryer is equipped with a labyrinth-design jacket welded to the outside of a dryer shell. The baffles in the jacket assure turbulent flow of the heating medium for maximum heat transfer. Intense mixing results in continuous exposure of the material to the heating surface. Contact time is long enough to allow the required heat to be transferred but short enough to prevent material from thermal damage. Because of continuous mechanical agitation by aerodynamically shaped mixing elements, there is no segregation of products, with bulk densities and shapes varying dur-

ing drying. Compact design and lack of airflow permit operation under vacuum, hence lowering the process temperature. Further, chopping mills can be added to the unit, if required, to break agglomerates and control particle-size growth, thereby reducing the drying time (Figure 20.40). The U-Max dryer finds applications in processing stearates, pharmaceutical intermediates, pigments, hot malt adhesives, ceramic and semiconductor materials, and more [65].

20.3.3 JETZONE DRYER

Referring to Figure 20.41, the JetZone[®] (Wolverine, Proctor & Schwartz, LLC, Merrimac, MA) fluid bed dryer uses a series of air nozzles to direct hot air onto the surface of a nonperforated belt conveyor or vibrating solid pan. Hot air passes from the pressure plenum chamber through a set of nozzles onto an oscillating solid conveyor surface, thus creating a “bed of air” under and around particles. The air jets also fluidize gently the particles suspended on the bed of “reflected” air as they progress through the dryer. Air then rises vertically from the conveyor around the nozzles and exits into a cyclone in which airborne particles are removed. Precise control over the air velocity allows setting of exact fluidization in any section of a dryer, therefore ensuring even exposure of all the particles to the drying medium. The dryer can also be separated into multiple zones with different air temperatures for process control. The principal advantages of a JetZone dryer are:

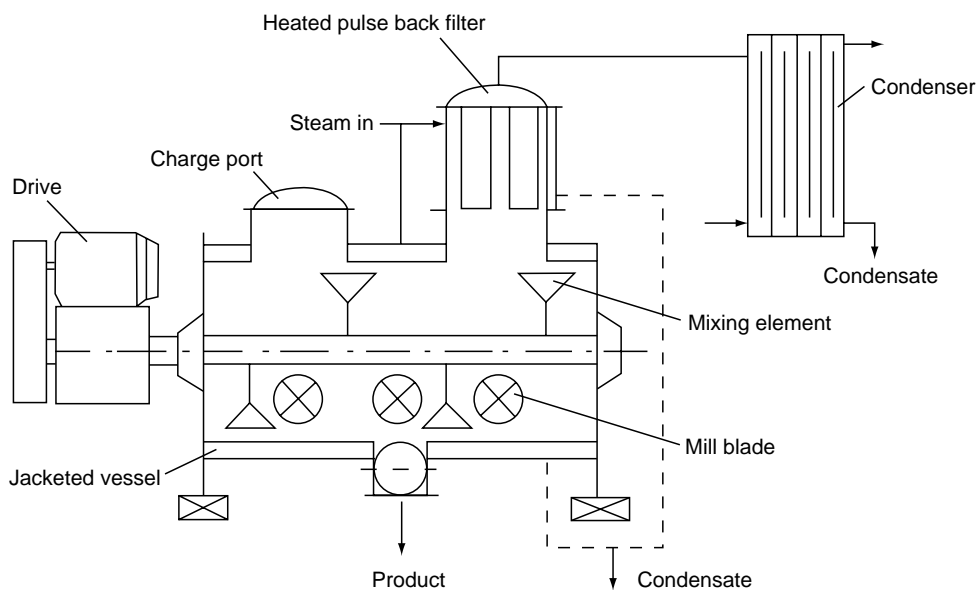


FIGURE 20.39 Schematics of the U-Max vacuum dryer. (Courtesy of PROCESSALL, Inc., Cincinnati, OH. With permission.)

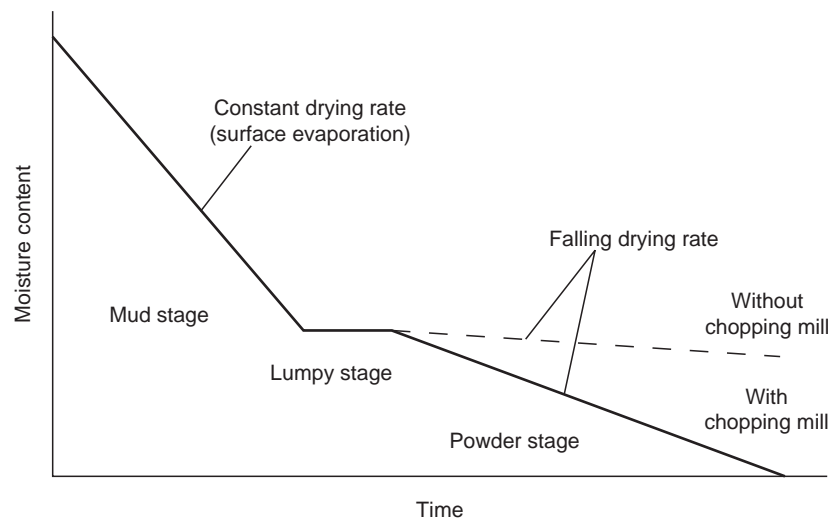


FIGURE 20.40 Typical drying curves in a U-Max dryer with and without chopping mill. (Courtesy of PROCESSALL, Inc., Cincinnati, OH. With permission.)

Uniform and controllable drying
 Good cleanability (no perforations to be clogged with fines)
 Fewer moving parts
 Quick product changeover

The JetZone dryer is capable of processing materials that vary in size, shape, and density. These include abrasive particles, cut leaf forms, diatomaceous earth, fibrous and pelletized foods, sawdust, pet foods, wood chips, and so on [66]. The operating limits of the dryers are 400°C for gas temperature, 0.3 m/s for conveyor speed, 70 m/s for air jet velocity, and 2 m/s for air

velocity through the bed of particles. The production capacity for a single dryer ranges from 90 to 41,000 kg/h.

20.3.4 VORTEX DRYERS

Vortex dryers are characterized by a spiral flow of a particle-gas suspension due to tangential entry of the gas stream into the dryer chamber. Wet material is fed directly into the drying chamber by a screw feeder or nozzle atomizer, or dispersed with the drying medium (air, inert gas, superheated steam, etc.). After performing several spiral trajectories, the particulate

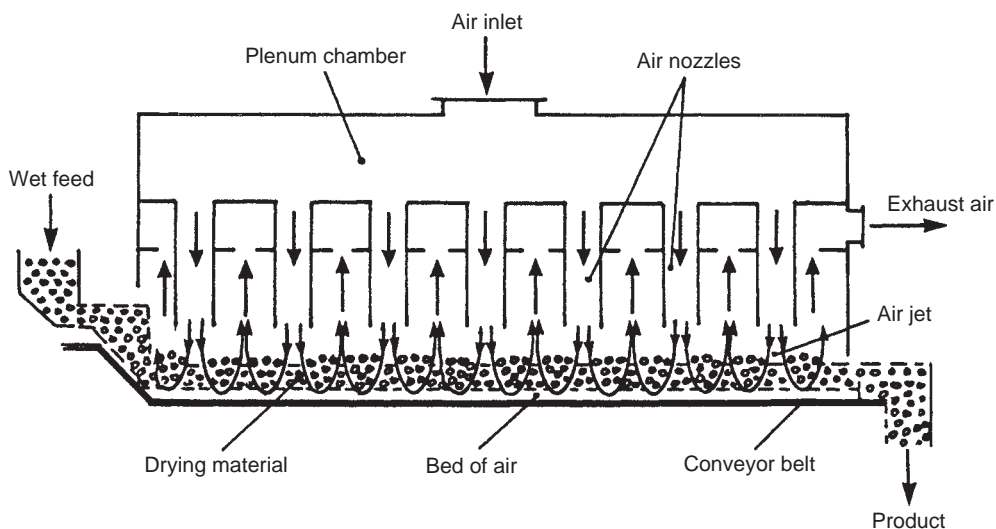


FIGURE 20.41 JetZone fluidized bed dryer. (Courtesy of Wolverine, Proctor & Schwartz, LLC Merrimac, MA. With permission.)

material leaves the dryer with an exhaust gas. Alternatively, it may be separated from the gas within the dryer and further discharged via a rotary lock.

In a vortex dryer, particles are subjected to the coupled action of centrifugal and drag forces, which causes penetration of particles to the dryer wall in a spiral trajectory. Near the dryer wall, the particles lose a part of their kinetic energy due to friction and nonelastic collisions. This results in increase of the relative gas-particle velocity, which enhances the external heat and mass transfer. The transfer processes are intensified additionally owing to disturbances in the boundary layer on the particle surface as a result of its unsteady motion in a highly turbulent flow and rotation in the near-wall region due to wall friction and interparticle collisions. The coupled action of drag and centrifugal forces also causes particle segregation, which promotes drying uniformity because the heavier (wet and large) particles reach the near-wall region with highly intensive drying conditions well before the smaller ones, which dry almost instantaneously and are taken away from the central core zone of the dryer.

Vortex dryers can be configured vertically or horizontally (Figure 20.42). The vertical configuration is widely known as a convex dryer. The horizontal vortex dryers are less popular, but they can easily be used in a multistage arrangement that allows for extended residence times, controlled temperature regimes, and intermittent operation, if necessary.

Multistage horizontal vortex dryers (Figure 20.43) have been developed extensively in Russia and used for drying chemicals, vegetables, and pharmaceuticals.

Typical dryers are 0.2 to 1.0 m in diameter, which may be configured in aggregates of 1 to 6 stages. Table 20.17 summarizes operating data for a variety of materials that have been processed in such dryers. An industrial four-stage vortex dryer for processing of polyvinyl alcohol, polyvinyl acetate, and pressed sea buckthorn is presented in Figure 20.44. The advantage of possible operation at different temperature levels was taken in the case of drying sea buckthorn, which contains a large amount of bound water. The desired moisture content of 4 to 6% wb at a residence time of 4 to 6 min (required by drying kinetics and hydrodynamic conditions) was achieved in a periodic temperature regime via alternate supply of hot air at 150 and 60°C [67].

20.3.5 AGITATED FLASH DRYER

As shown in Figure 20.45, an agitated flash dryer consists of a vertical cylinder with an inverted conical bottom, an annular air inlet, and an axially mounted rotor. Tangential air inlet with the action of the rotor causes a turbulent whirling gas flow in the drying chamber. The wet material, typically filter cake, is extruded off the screw feeder into the drying chamber just above the agitator, thus it becomes coated by a dry powder flowing with the swirling air [68]. Alternatively, a cavity pump can be used for a dilatant fluid feed. The powder-coated lumps then fall into the fluid bed and are kept in motion by the rotor. As they dry, the friable surface material is abraded by a combination of attrition in the bed and the mechanical action of the rotor. Thus, a balanced fluidized

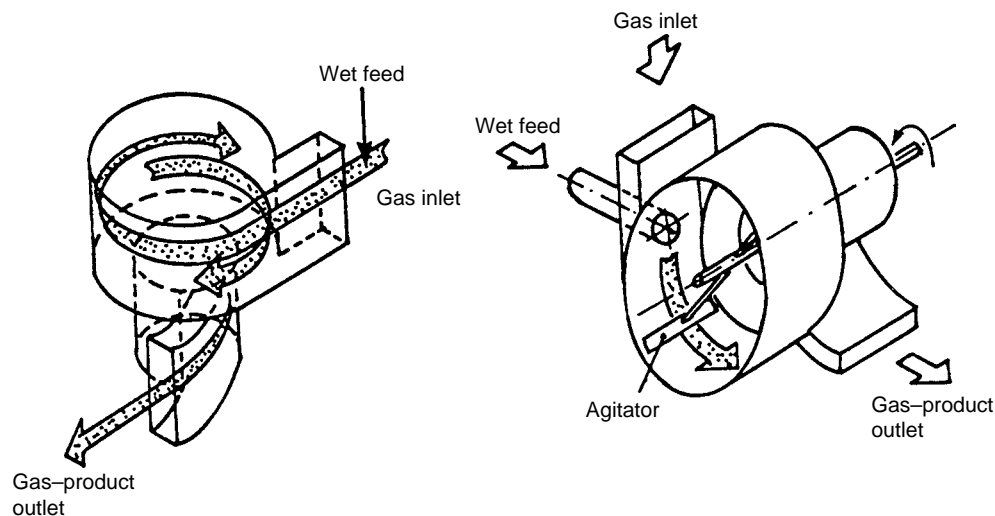


FIGURE 20.42 Horizontal (left) and vertical (right) configurations of a vortex dryer.

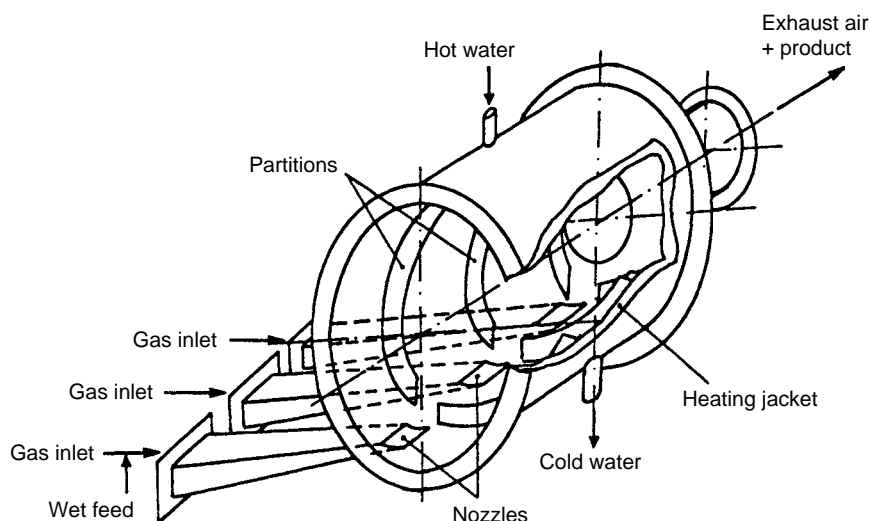


FIGURE 20.43 Multistage horizontal vortex dryer. (From Sokolovskii, A.A. and Khudnyeev, I.K., Russian Patent 421,870, 1974.)

bed is formed that contains all intermediate phases between the raw material and the finished product. The dry (therefore small and light) particles become airborne and rise up the wall of the drying chamber to the filter bag, providing continuous backmixing in the feed zone. The largest lumps fall back into the fluid bed to resume drying.

Typical operating parameters for a spin-flash dryer are [69]:

Feed material	Dilatant liquids, cohesive pastes, filter cakes, moist granules
Drying medium	Air, inert gas, low-humidity waste gases
Inlet gas temperature	Up to 980°C
Product throughput	Up to 10,000 kg/h
Residence time	5–500 s
Rotor speed	50–500 rpm
Mean particle size	3–70 μm

TABLE 20.17
Characteristics of Multistage Horizontal Vortex Dryers

Material	Dryer ^a	T_{gl} , °C	T_{go} , °C	T_m , °C ^b	X_0 , %	X_f , %	W_m , kg/h	W_{ev} , kg/h	t_r , min
Thiamine bromide	VD-800/1	130	70	80	16	1.2	200–250	40–50	1.5
Phthalic acid	VD-800/1	160	80	130	18	0.5	250–300	45–55	1.0
DKGK hydrate ^c	VD-800/1	90	60	70	16	7.0	150–250	20–25	0.8
Pyridine	VD-800/1	160	90	130	20	0.5	250–300	70–75	1.0
Polyvinyl alcohol	VD-1000/6	148	90	132	60	2.0	650–700	1000–1100	6.5
Polyvinyl acetate	VD-800/3	150	90	136	50	2.0	250–300	250–300	4.5
Polystyrene	BD-800/1	100	60	70	40	1.0	300–350	200–220	1.2
Threonine sulfate	VD-800/1	160	80	132	18	0.5	250–300	65–70	1.0
Sea buckthorn	VD-1000/4	T_g^d	70	100	50	5.5	100–150	100–150	7.5
Sesame seeds	VD-600/2	150	110	120	30	1.0	100–150	50–65	4.0

^aDryer code (slash separates the dryer diameter from number of stages).

^bTemperature of product degradation.

^c2-Keto-L-gluconic acid.

^dOscillatory regime (150/60°C).

Source: From Sokolovskii, A.A., Kondrateva, N.M., and Chlenov, V.A., *Khimiko-Pharmaceuticheski Zhurnal*, 17, 1097, 1983 (in Russian).

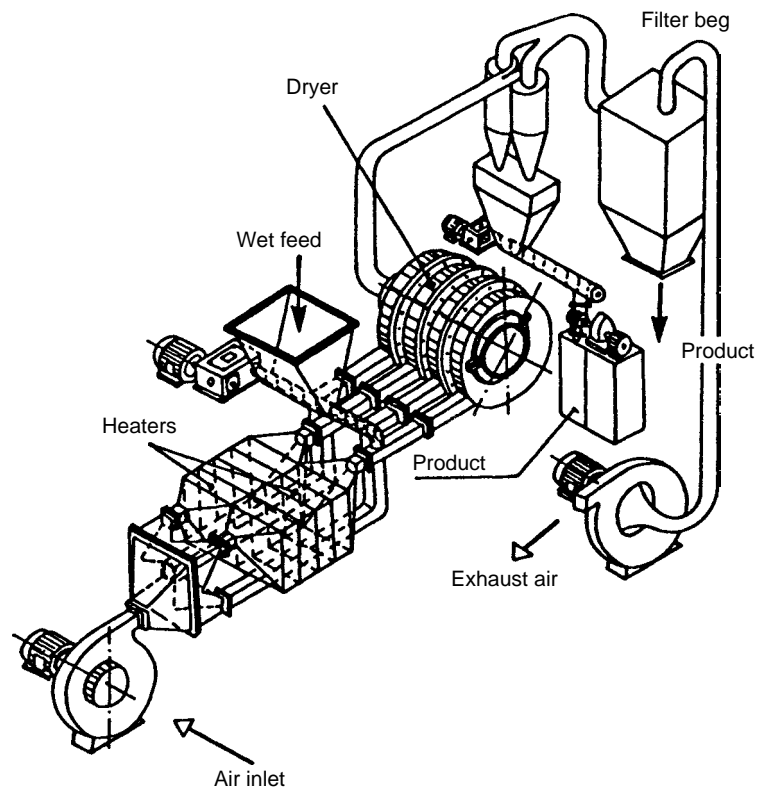


FIGURE 20.44 Four-stage vortex dryer. (From Sokolovskii, A.A., Kondrateva, N.M., and Chlenov, V.A., *Khimiko-Pharmaceuticheski Zhurnal*, 17, 1097, 1983 (in Russian).)

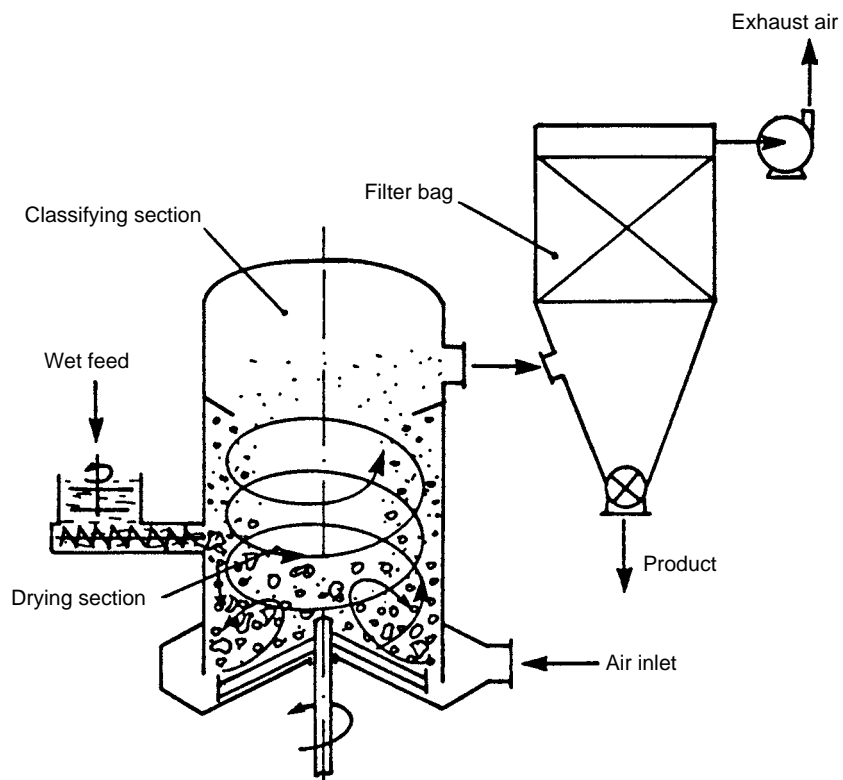


FIGURE 20.45 Agitated flash dryer.

In comparison with a spray dryer, the agitated flash configuration has a much shorter residence time and consequently is considerably smaller and requires less building space. Moreover, its ability to dry to an even lower moisture content than a spray dryer results in significant operating and capital cost savings (Table 20.18).

Despite their obvious size and cost advantages, the agitated flash dryer cannot always replace a spray dryer. Such cases occur when free-flowing spherical particles of a defined size range have to be obtained or when agglomeration is needed. Also, the agitated flash dryer cannot handle fibrous materials such as fruit by-products or cellulose because they tend to form pellets in the dryer chamber.

20.3.6 SPIN-STREAM DRYER

A modification of the vortex dryer in the vertical arrangement that improves the performance of the well-known spin-flash dryers is the spin-stream dryer. Designed basically for drying liquids, slurries, and suspensions, the spin-stream dryer has a multijet air distributor with curved and movable vanes (Figure 20.46) [70,71]. Changes in the position of the vanes cause change in the angle at which air is supplied to the drying chamber as well as a change in the linear velocity of the air jet. This allows variation of the jet-to-axial air velocities ratio, thus granting control of the particle residence time independently of the

gas flow rate. Moreover, profiled flights at the periphery of the rotor deflect periodically the local air jet toward the dryer center, which increases the active drying zone [72]. This unique design of a dryer permits processing of hard-to-dry materials such as lignopol (sodium lignosulfonate)—a highly viscous liquid that becomes a sticky paste when dried and thus tends to form a gummy load inside a dryer. The performance of the spin-stream dryer for lignopol is as follows:

Dryer diameter	0.62 m
Dryer height	2.0 m
Gas distributor diameter	1.1 m
Feed rate	0.2 kg/s
Initial moisture content	50% wb
Final moisture content	15% wb
Inlet air temperature	150°C
Outlet air temperature	100°C
Airflow rate	2015 Nm ³ /h
Evaporation rate	0.0082 kg/s

Selected dryer indices for other materials with distinct properties are given in Table 20.19.

Spin-flash and spin-stream dryers are examples of dryers operating in an “active hydrodynamic regime” [73]. Comparison of several such dryers that may be useful for dryer selection is given in Table 20.20 [74].

20.3.7 ROTACURRENT DRYER

Particle residence time in conventional (straight-tube) flash dryers is determined by the length of the tube, air velocity, and material properties. Figure 20.47 presents the rotacurrent flash dryer, which offers increased residence time (by a factor of up to 5 with no backmixing) in a much shorter tube since the geometrical height is not directly tied to the retention time. Wet feed enters the rotacurrent dryer with the main hot airstream generated by a conventional blower and heater system. Because the secondary hot air is tangentially injected into the main airstream, the wet feed follows a spiral path through the annular space between an internal cylinder and a dryer shell. The trap at the bottom of the dryer is counterweighted so that the extra heavy particles (lumps, stones, stray metal, etc.) are automatically discharged. Hot air is added to the main flow through adjustable louvers spaced along the internal cylinder. This secondary airflow produces an aerodynamic effect to reinforce the initial spiral flow and a thermal effect to compensate for the loss of drying potential as the main airstream cools due to moisture evaporation.

The rotacurrent dryer can handle particulate and nonsticky products such as aluminum silicate,

TABLE 20.18
Performance, Size, and Cost Comparison of Spray Dryer with Spin Flash Dryer for Drying Yellow Iron Oxide

Parameter	Spray Dryer	Spin Flash Dryer
Product throughput, kg/h	400	400
Moisture content, % wb		
Initial	30	30
Final	0.4	0.4
Feed rate, kg/h	1360	885
Evaporation rate, kg H ₂ O/h	560	486
Gas consumption, m ³ /h	125	62
Power consumption, kWh	40	30
Dryer diameter, m	4.3	0.8
Floor area, m ²	60	30
Building volume, m ³	700	150
Investment cost ratio	1	0.676
Operation cost, \$/h	36.88	26.57
Cost per kg of product, \$/kg	0.092	0.066

Source: From *Drying Handbook*, APV Crepaco, Inc., Attleboro Falls, MA, 1993.

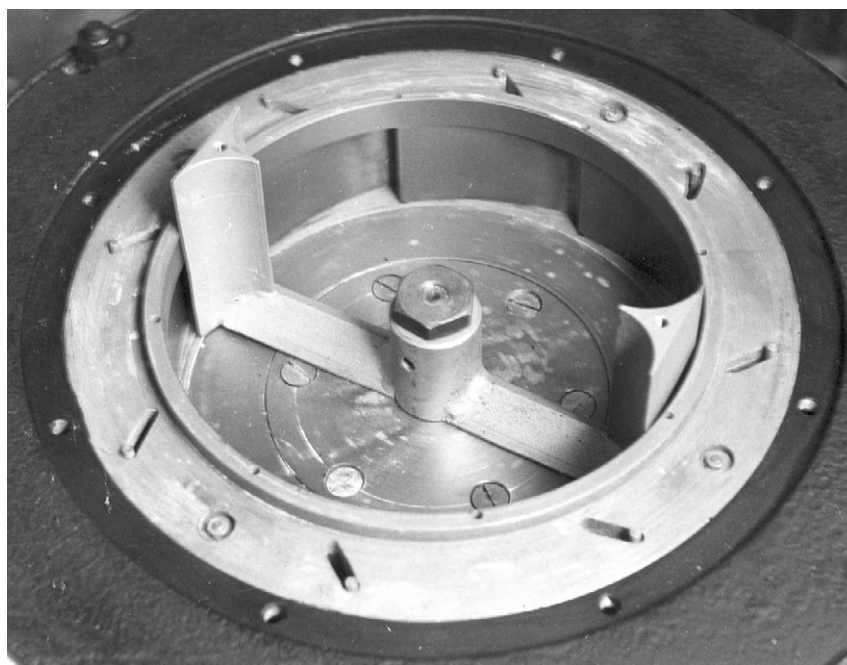


FIGURE 20.46 Gas distributor and agitator in a spin-stream dryer.

copper powder, pyrite ore, charcoal, granulated rubber, chopped straw, detergents, and the like [75]. Performance data on the rotacurrent dryer are given in Table 20.21.

20.3.8 VENTURIJET DRYER

Figure 20.48 shows an interesting concept of a dryer, the Venturijet, considered as an energy transformer. The system receives a reduced mass of hot air with relatively high dynamic pressure, the partial transformation of which ensures product and air recycling without mechanical devices. The primary airstream

from a standard heater is injected into a vertical venturi to create suction, which allows feed dispersion as well as mixing of the primary hot air and recycled air to produce a larger mass of air at a lower temperature. A secondary hot airstream is introduced tangentially to generate a spiral trajectory of heavier recycled particles inside the drying chamber. Simultaneously, the lighter particles are conveyed into the cyclone and discharged. An adjustable selector at the dryer outlet ensures selective ejection of the dry particles. The principal features of the Venturijet dryer are [75]:

Lower air consumption by approximately 50% than that in conventional dryers

TABLE 20.19
Drying Indices for Spin-Stream Dryer

Parameter	Imidazole	CaCO ₃ ^a	CaCO ₃ ^b	H-acid ^c	Lignopol ^d
Heat consumption, MJ/kg H ₂ O	3.5–4.2	3.5–5.5	8.0–13.0	5.6–7.7	12.0–20.0
Air consumption, N m ³ /kg H ₂ O	23–50	30–48	48–84	38–53	72–144
Thermal efficiency	0.6–0.7	0.4–0.6	0.2–0.5	0.5–0.6	0.2–0.3
Evaporation capacity, kg/m ³ h	80–200	120–320	60–80	90–170	35–70

^aReagent grade.

^bCosmetic chalk.

^cC₁₀H₂O₇ S₂NNa.

^dSodium lignosulfonate.

Source: From Kudra, T., Pallai, E., Bartczak, Z., and Peter, M., *Drying Technology*, 7, 583, 1989.

TABLE 20.20
Characteristics of Various Dryers for Particulate or Pasty Materials

Dryer Type	K_1	K_2	K_3	n	Operating Parameters		Electricity	
					u , m/s	ΔP , kPa	ϵ	Static electricity kV
Fluid bed	1	1	2	1–3	0.1–1.5	1.5–5.0	0.6–0.75	5.0
Fast fluid bed	4	6	10	5–10	1.5–15	0.2–0.5	0.7–0.85	1.0
Spouted bed	6	3	3	3–5	8–60	3–30	0.65–0.8	3.0–5.0
Jet-spouted bed	5	6	4	5–10	10–40	0.1–0.3	0.9–0.99	1.0
Vibro-fluidized bed	5	10	6	40–100	0.05–0.8	0.1–0.2	0.6–0.7	0.5–2.0
Vortex bed	6	6	8	5–10	10–80	1.5–2.5	0.65–0.8	1.0–3.0
Pneumatic transport	6	8	9	20–30	10–50	0.2–1.0	0.8–0.99	0.5–1.0
Swirling streams	8	6	10	5–10	15–40	0.6–2.0	0.85–0.95	1.0
Impinging swirling streams	10	6	10	5–10	10–50	0.5–2.5	0.8–0.99	1.0

n is a number of theoretical mixing stages while K 's are the following criteria based on Sazhin's classification, which assigns the grade from 1 to 10 (maximum): K_1 , hydrodynamic stability (10 is stable); K_2 , mixing behavior (10 is perfect mixing); K_3 , intensity of hydrodynamic turbulence under normal operating conditions (10 is highly intensive).

Source: From Sazhin, B.S. and Sazhin, V.B., *Scientific Principles of Drying Technique*, Nauka, Moscow, 1997 (in Russian).

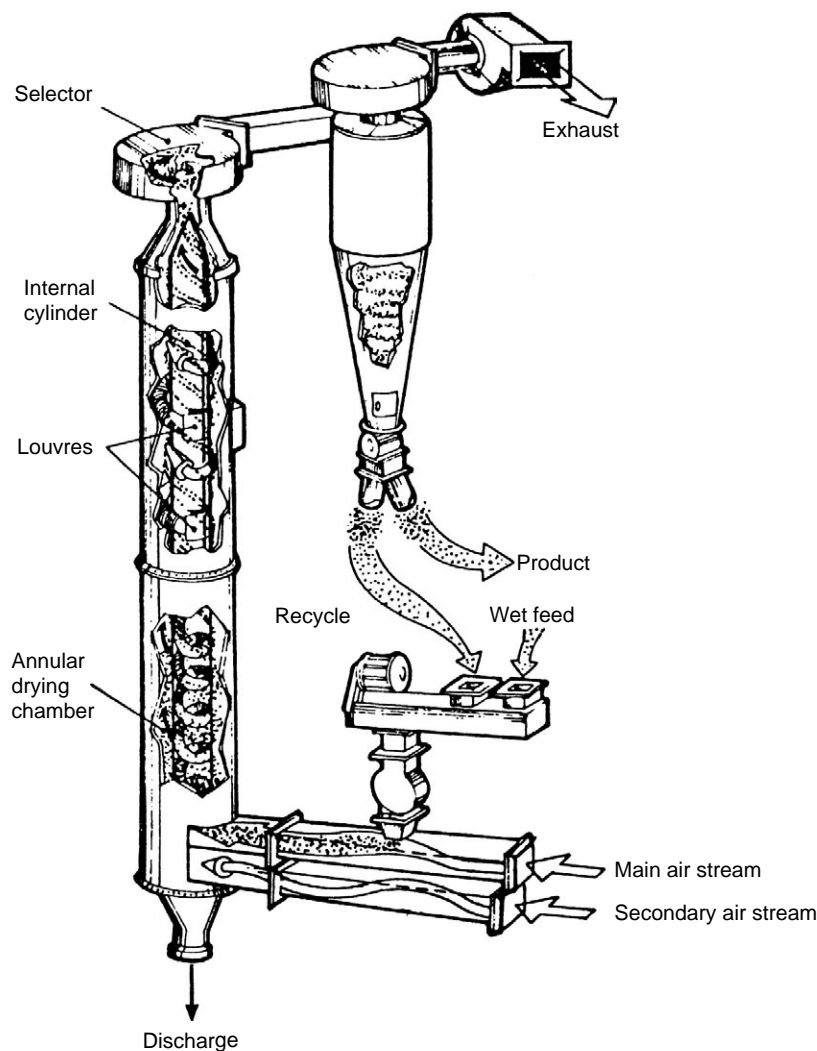


FIGURE 20.47 Rotacurrent dryer. (Courtesy of Heyl & Patterson, Inc., Renneburg Division, Pittsburgh, PA. With permission.)

TABLE 20.21
Performance Data on Rotacurrent Dryers

Parameter	Tricalcium Phosphate	Polyethylene Resin	Sodium Bicarbonate	Distillers Dark Grain
Production rate (feed), kg/h	1,515	4,773	9,318	22,211
Moisture content, % wb				
Initial	55	10	6	55
Final	10	0.5	0.1	10
Gas temperature, °C				
Inlet	538	177	165	450
Outlet	90	77	82	82
Air consumption, N m ³ /h	6,382	17,244	17,906	114,860
Evaporation rate, kg H ₂ O/h	757	455	550	13,210
Fuel consumption, kg/h ^a	62	56	68	978
Dryer diameter, m	0.6	0.86	1.02	2.7
Dryer height, m	6.5	8.13	9.1	15.3

^aBased on natural gas at 50,777 kJ/kg.

Source: From Heyl & Patterson, Inc. Technical brochure, Heyl & Patterson, Inc., Renneburg Division, Pittsburgh, PA.

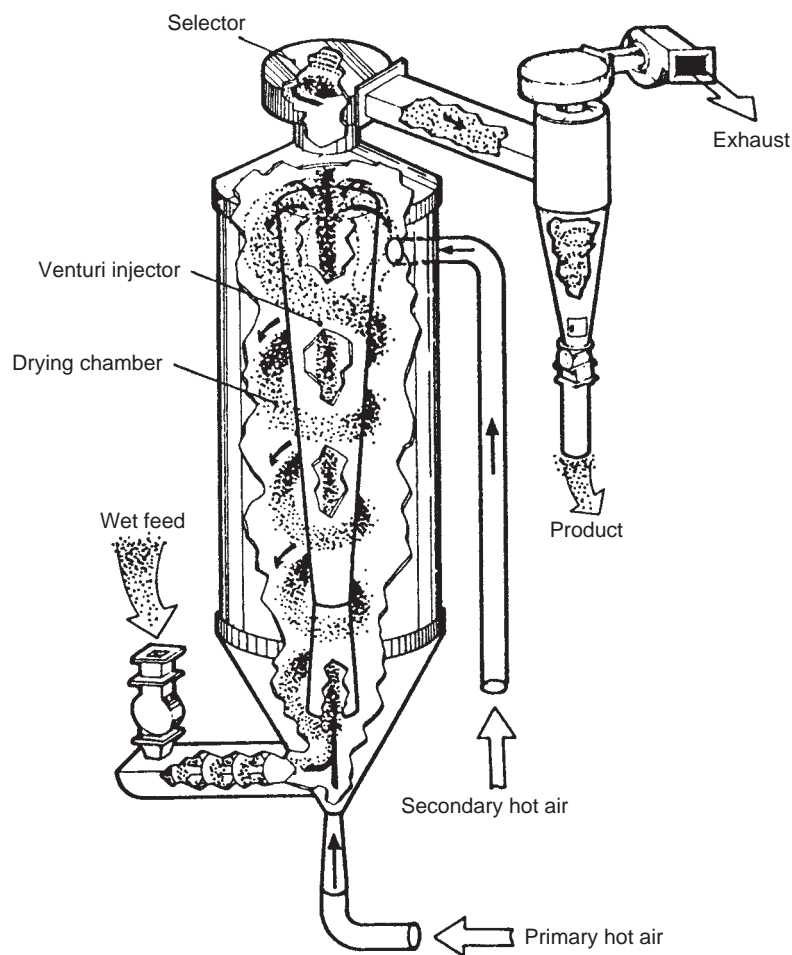


FIGURE 20.48 Venturijet dryer. (Courtesy of Heyl & Patterson, Inc., Renneburg Division, Pittsburgh, PA. With permission.)

Lower stack heat losses due to internal air recycling
Improved feed dispersion at the entrance to the system, thus providing the possibility for operation with sticky materials

Increased residence time for heavier (wetter) particles

Typical products dried in the Venturijet dryer include insecticides, fine chemicals, polymer agents, and the like. Performance data on the Venturijet dryer for these products are given in Table 20.22.

20.3.9 AGITATED CONTACT DRYERS

Agitated contact dryers are normally used for high-volume drying of bulky and granular materials via contact heat transfer from a heated rotor, a heated jacket, or a combination of these. In a typical design, single or double intermeshing screw augers have shaft and hollow flights heated by a circulating thermal fluid. Due to rotation of the auger, the material is thoroughly mixed and transported along the drying chamber. The evaporated moisture is entrained by a small amount of air passing over the material. Alternatively, the dryer may operate under vacuum. Figure 20.49 shows a paddle dryer with a twin agitator consisting of a number of spaced blades that mix the material with essentially no pumping effect [76]. The material flow through the dryers is due to mild

fluidization caused by the mechanical action of a rotor. Typical applications are drying of fine and moderately wet materials such as gypsum, pigments, and dyestuffs.

Figure 20.50 shows the operating principle of the SOLIDAIRE[®] (Bepex International, Minneapolis, MN) dryer, which combines the convective and contact heat transfer to dry solids, slurries, gels, and wet filter cakes. It consists of a mechanical agitator rotating within a cylindrical housing that is jacketed for indirect heating. The agitator is equipped with a large number of narrow, flat, adjustable-pitch paddles that sweep close to the inner surface of the housing. The high speed of the paddle tips (10 to 20 m/s) creates turbulence in the compact material layer, thus breaking up agglomerates and continuously renewing the material surface exposed to heat, which finally provides the heat transfer coefficients in the order of 110 to 570 W/(m² K). Residence time can be varied from seconds to approximately 20 min by adjusting the paddle pitch or by changing the rotor speed. The SOLIDAIRE holdup volume is up to 5.6 m³ and the heat transfer area up to 100 m². The dryer can operate at temperatures up to 430°C and pressures up to 1000 kPa [77].

20.3.10 REMAFLAM DRYER

A textile fabric typically goes through several drying processes before it is in the finished stage. Conventionally impinging jets, contact, infrared, or through dryers are used for drying textiles. With electric infrared drying and convection dryers, the thermal efficiencies are in the range of 70%. In the Remaflam[®] drying process [78], the fabric is soaked in an aqueous solution of methanol (35% v/v), which is then burned in a carefully controlled combustion chamber. Other solvents (which must be miscible with water and burn without soot) can also be used, but methanol has significant advantages over ethanol or isopropanol (Table 20.23). Acetone has a very low flash point (−20°C) and hence is excluded. Both methanol and ethanol have flash points under 21°C. With added water, the flash point increases to 35°C. Sometimes isopropanol is added to improve wetting of the methanol–water solution to the fabric. Most dyes are not affected by the process.

Figure 20.51 shows a schematic of the Remaflam dryer. An infrared section at the entrance to the dryer ignites the liquid on the fabric surface (both sides). The fabric temperature is 45 to 70°C, although the radiator wall is at 800°C and the air in the upper section of the dryer is at about 600°C. The fabric speed depends on the fabric weight (g/m²), liquor applied, methanol concentration, and the

TABLE 20.22
Performance Data for Venturijet Dryers

Parameter	Insecticide	Polymer Agent	Manganese Carbonate
Production rate (feed), kg/h	2272	455	3508
Moisture content, % wb			
Initial	15	19	30
Final	0.5	0.5	5
Gas temperature, °C			
Inlet	315	200	815
Outlet	77	65	130
Air consumption, N m ³ /h	3924	1728	6948
Evaporation rate, kg H ₂ O/h	331	85	923
Fuel consumption, kg/h ^a	29	104	90
Dryer diameter, m	1.3	0.94	1.3
Dryer height, m	6.9	5.52	6.9

^aBased on natural gas at 50,777 kJ/kg.

Source: From Heyl & Patterson, Inc. Technical brochure, Heyl & Patterson, Inc., Renneburg Division, Pittsburgh, PA.

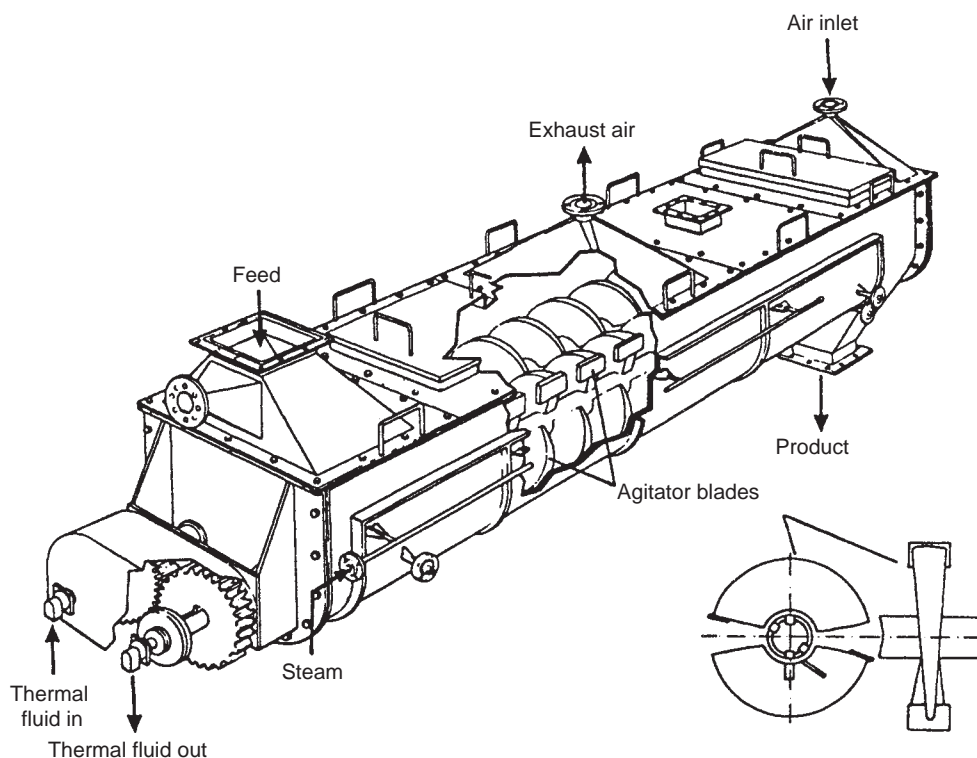


FIGURE 20.49 Paddle dryer with twin, disk-type agitator. (Courtesy of Komline-Sanderson Engineering Corporation, Peapack, NJ. With permission.)

residual moisture content. For a polyester-cotton blended fabric with a weight of 220 g/m^2 and liquor absorption by fabric of 43% by volume, the fabric speed is 42 m/min for a 36% v/v methanol liquor [78]. The drying zone is about 1.6-m long. The process has been successfully applied for drying of cotton,

polyester, polyamide fabrics, and blends. Fabrics with textures that do not permit good absorption of the alcohol-water liquor may create problems of uneven drying. Pilot testing of at least 30 min run on the Rameflam machine is recommended for new fabrics, blends, and the like.

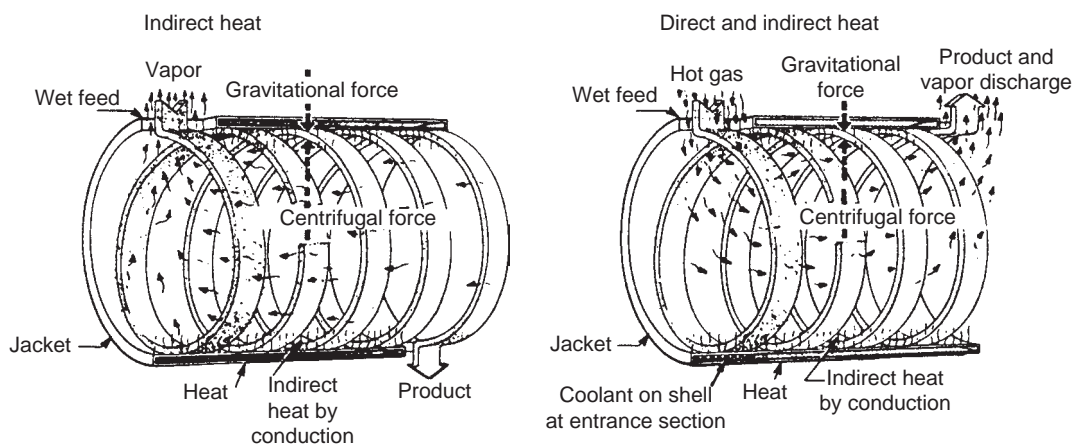


FIGURE 20.50 Typical processing modes of the SOLIDAIRE dryer. (Courtesy of Bepex International, Minneapolis, MN. With permission.)

TABLE 20.23
Characteristics of Solvents Used in the Remaflam Process

Solvent	TLV, ^a ppm	Flash Point, °C	Heating Value, kJ/kg	Oxygen Demand, mol/mol	Relative Cost
Methanol	200	11	19,929	1.5	1.00
Ethanol	1,400	12	26,837	3.0	2.16
Isopropanol	400	13	30,472	4.5	1.24
Dioxane	50	11	24,587	5.0	7.40

^aTLV, threshold limit value (lower toxic limits).

20.3.11 MICROGAS DRYER

Figure 20.52 presents a drying system that combines microwave and convective heating in an integrated, natural-gas-fired unit. In this system, a natural-gas-fired prime mover drives a generator that provides the electricity to power the microwave generator and all ancillary equipment (feeders, conveyor belts, air blowers, water pumps, etc.), whereas waste heat recovered from the engine is used for convective heating (cogeneration). Of the natural gas fed to the prime

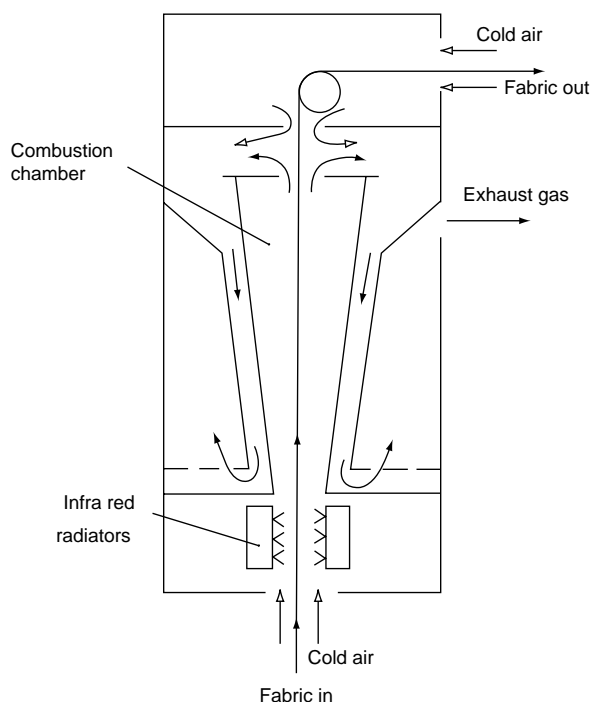


FIGURE 20.51 Airflow in the Remaflam dryer. (From von der Eltz, H.-V., Petersohn, G., and Schön, F., *Int. Textile Bull.*, 2, 1, 1981.)

mover, approximately 30% is converted to electricity and 50% is recovered as thermal energy. The overall efficiency of the MicroGas system for delivery of energy to the production process is claimed to be approximately 80% [79]. The system is recommended in applications for which:

- There is appreciable value in significantly decreasing the processing time
- A 20 to 50% reduction in energy consumption leads to large cost savings
- Product unit value is high
- Large decrease in equipment space is crucial
- Transformation of a batch process into a continuous one results in significant productivity increases
- Noncontact heat-exchange aspects of microwaves can produce unique product results

20.3.12 HELIX AND RING-TO-RING DRYERS

An innovative variant of the vibrated bed dryer is the so-called Helix (or ring-to-ring) dryer [80,81]. This dryer consists of a cylindrical chamber with a centrally located hollow column supporting a perforated helical tray (Figure 20.53). Two horizontal plates split this chamber into the upper gas plenum, the central drying chamber, and the base chamber. The drying medium (e.g., hot air or flue gas) entering the gas plenum passes through the openings in the head of the central column and then flows down the column. Since the column is closed at the bottom, the gas stream moves through a series of holes into the triangular plenum under the helical tray. The floor of the tray is made of overlapping radial slats, which form louvers to force gas in the same direction as the material movement along the helical tray. The gas stream then flows up through the material layer into the annular space between the helical tray and the chamber wall and then to the exhaust duct. The drying material enters and leaves the tray through rotary feeders that also act as gas seals. The column-tray assembly is resiliently supported by four springs rigidly connected to the base of the chamber and to the lateral surface of the column. To allow vibration the column is connected to the gas plenum and the base chamber by flexible seals.

Vibration causes the pseudofluidized particulate material to move downwards along the tray. At the same time the material is dried by the hot gas stream passing through the vibrated bed at a superficial velocity determined by heat and mass transfer requirements. This superficial velocity is usually lower than the fluidization velocity and varies from 0.7 to 3 m/s, depending on the material characteristics.

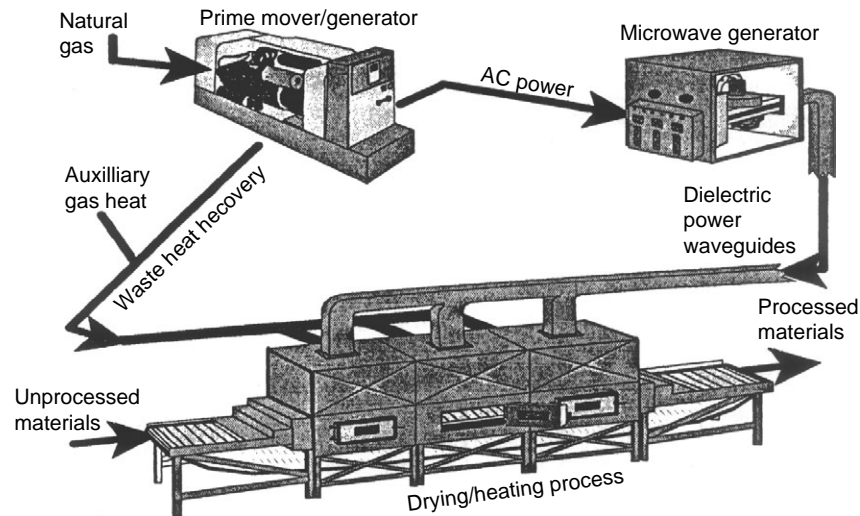


FIGURE 20.52 MicroGas configuration.

In its basic design, a motor-driven variable eccentric vibrator located at the base of the dryer vibrates the column with a nominal amplitude of 0.9 mm at a frequency of 16.6 Hz (both of these parameters could be changed). Because the material flow rate and thus the residence time in the dryer can be varied by increasing or decreasing the amplitude and by changing the direction of the amplitude vector, the newer model of the dryer is equipped with a synchronized vibrator of special design. This vibrator provides separate control over the vertical and horizontal

components of the vibration vector and thus allows nearly independent control of the degree of fluidization and the material residence time [80]. Amplitude and frequency of vibration can range from 0.25 mm and 30 Hz to 1 mm and 15 Hz.

Because of the presence of the horizontal component of vibration and the louvered tray ensuring a horizontal air velocity component, the material can be transported at desired rate even on a horizontal tray. Therefore, an alternative model of the dryer termed the ring-to-ring dryer [82] is designed as a series of

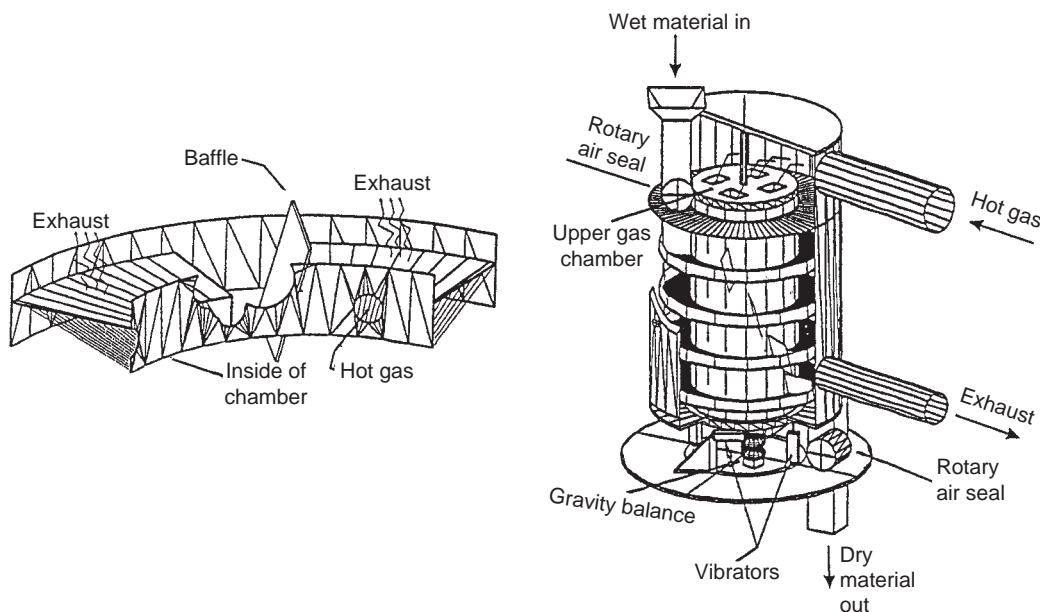


FIGURE 20.53 Ring-to-ring dryer. (Courtesy of Carlyle Consulting Limited, Vancouver, BC, Canada. With permission.)

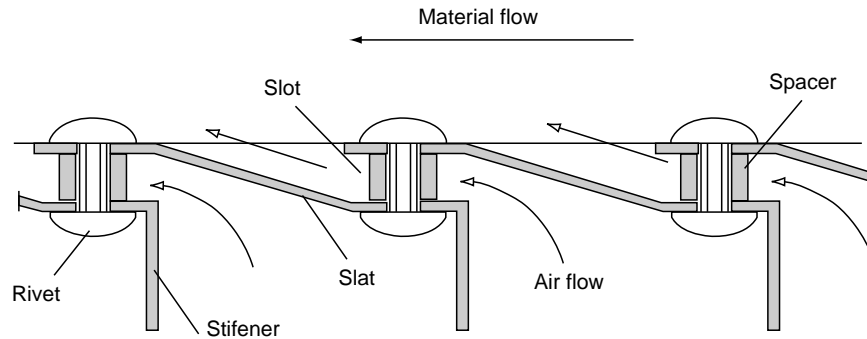


FIGURE 20.54 Jet ventilated conveyor tray. (From Carlyle, A.M., U.S. Patent 6,035,543, 2000.)

annular trays equally spaced along the central column. Each tray has an opening to allow the material to tumble from an upper to a lower tray progressively until the bottom tray exit is reached. Each tray opening is offset in alignment, in a manner that requires the material to travel the full circumference of each tray before it tumbles to the tray below. A baffle mechanism is provided in each opening to turn the material trough 180° as it falls from tray to tray.

Either design of the dryer is equipped with the “jet ventilated” conveyor tray [83]. The tray shown in Figure 20.54 comprises a plurality of slats arranged to overlap in a manner to result in a slot along the leading edge of the top slat. Each slot forces gas flow in the same direction as the flow of particulate material. The slats overlap in normally one-quarter of the narrowest width of the slat. Because the gas flow through the slot is a function of the slot length and the vertical height of the slot, the slots are sized to ensure required hydrodynamics for a given particulate material. Further, the slat width is determined by the number of slots and the conveyor length.

An important feature of the Helix dryer is its compact size. The dryer with an evaporative capacity of about 8400 kg of water/h at air temperature 315 and 82°C at the dryer inlet and outlet, respectively, is about 3.5 m in diameter and 5 m high, and contains about 75 m^2 of tray area. A conventional vibrated fluid bed dryer for a similar application would be about 1.2 m wide and 48 m long [84]. Cost and energy efficiency comparisons, however, are not available.

20.3.13 DRY-REX DRYER

One of the options for low-cost drying is the use of air at ambient temperature as a drying medium using moisture concentration gradient as the primary driving force. The Dry-RexTM system uses unsaturated ambient air drawn through a moving bed of wet

particulate material or palletized paste [85,86]. Here, drying is carried out in two stages (Figure 20.55) in the first stage large-size material (such as sugarcane or bark) is chopped to enhance the heat and mass transfer area as well as to reduce resistance to moisture diffusion. Pasty or semisolid feeds may be mechanically pressed to form a granular product of required dimensions. In the second stage, drying occurs as ambient air passes through the stationary bed placed on a multideck perforated belt conveyor.

The ambient air dryer can be competitive with other dryer types because its simple design reduces both investment and operation costs. Also, no high-grade thermal energy is required for drying. These advantages are partially offset by the need for high-capacity fans to provide air velocity high enough to compensate for the lower heat transfer rates due to the use of lower air temperature. Also, the ambient air should be sufficiently dry for this method to work efficiently. An alternative model of the Dry-Rex system includes a heat exchanger to preheat ambient air utilizing low-grade heat such as waste process water, steam, or exhaust air from other thermal processes.

The ambient air dryer as shown in Figure 20.56 has been tested by a manufacturer for a variety of materials including solids (bark, wood chips, and sugarcane residues), pastes (wood pulp, pulp and paper mill sludge), and powders (ash and fly ash). Table 20.24 presents representative data on ambient air drying of various wastes from paper mills [85].

20.3.14 DRYER-GRANULATOR

Two techniques are commonly used to distribute a liquid binder in a fluidized bed of the material to be granulated: (i) spraying of the liquid over the top surface of the fluid bed and (ii) injecting the liquid spray into the bed of fluidized particles. The major disadvantages of both techniques are: nonuniform

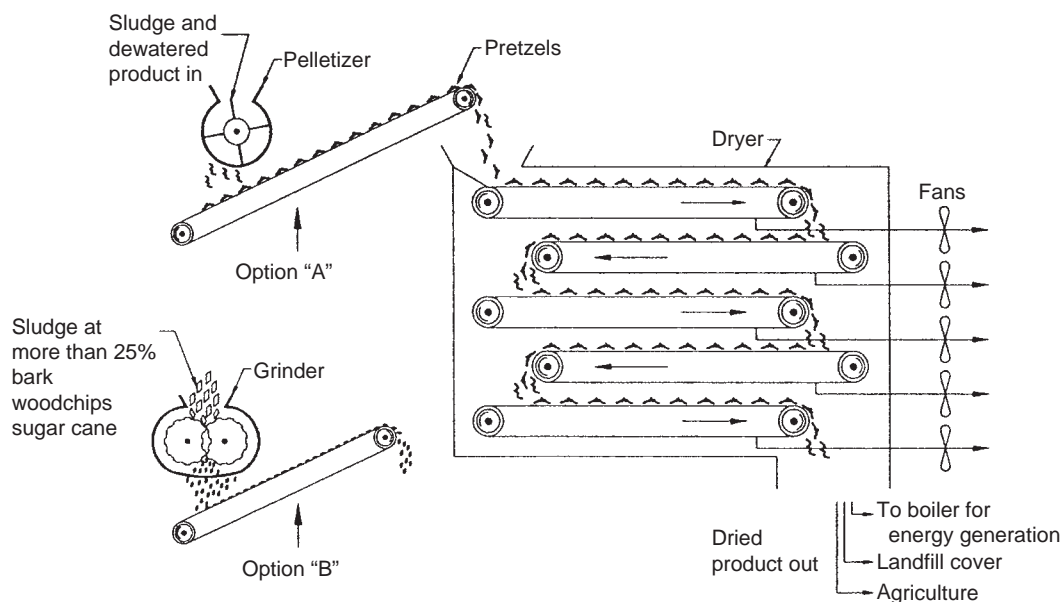


FIGURE 20.55 Basic configuration of Dry-Rex dryer.

coating of the particles by the liquid binder, need for multiple nozzles, clogging of nozzles, and high-energy consumption for liquid dispersion.

Marchevsky [87] has presented an innovative technique of feeding liquid into a fluidized bed of particles to be granulated (Figure 20.57). Essentially, this

dryer-granulator consists of a slowly rotating disk (from 400 to 500 rpm) immersed in the circular bed, which distributes the liquid binder flowing out of a hollow shaft over the disk surface. This liquid film coats a fraction of particles in contact with the disk while the centrifugal force moves them away from the

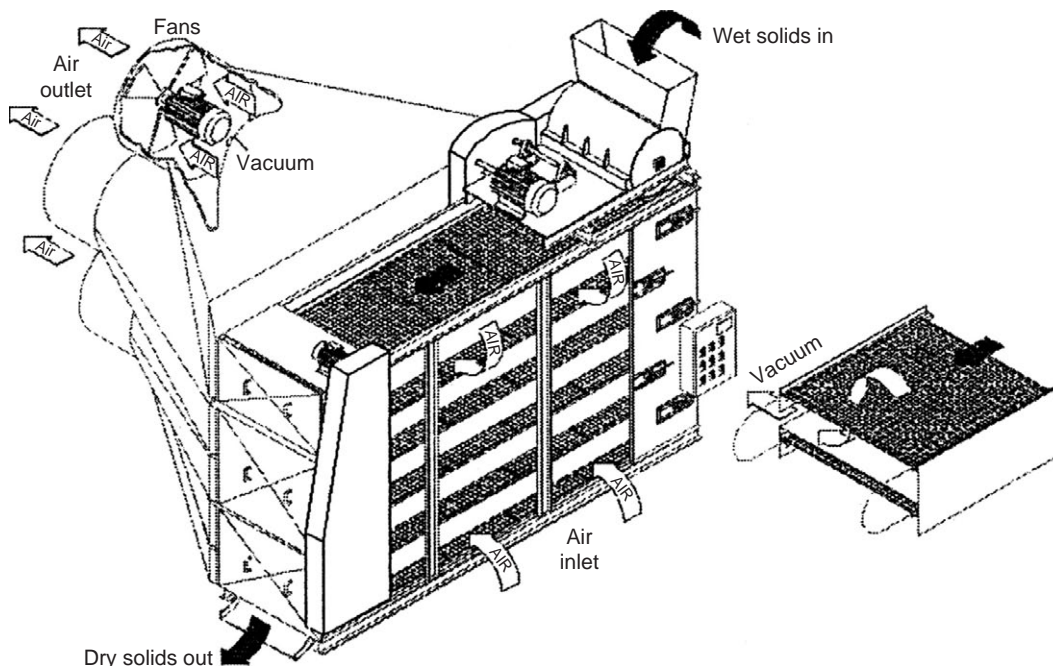


FIGURE 20.56 General view of the Dry-Rex dryer. (Courtesy of Mabarex, St. Laurent, Canada. With permission.)

TABLE 20.24
Ambient Air Drying of Residues from Paper Mills

Material	Air Temperature, °C	Air Relative Humidity, %	Moisture Content, % wb	
			Initial	Final
Primary–secondary sludge (% mass)				
10–90	17	60	87.4	30.7 ^a
66–33	17	60	85.4	32.0 ^a
64–36	5.5	30	70.2	26.8 ^a
35–65	18	62	84.4	22.0 ^a
30–70	15	40	76.0	17.1
60–40	17	62	81.0	22.0 ^a
Deinking sludge –50% clay	16	92	61.7	7.9
Frozen bark	20	62	60.0	30.0 ^a
Bark-sludge (50–50%)	20	29	60.0	20.5 ^a

^aLower values can be obtained but these were determined by the end-users as sufficient for firing in the boiler.

Source: From Courtesy of Mabarex, St. Laurent, Canada. With permission.

rotor. These particles agglomerate with others and dry in a hot airstream in the fluid bed. Because of particle recirculation in the bed, the cycle of coating, agglomeration and drying repeats itself, which results in a progressive and thus uniform drying. The fine particles are separated from the product granules in a pneumatic classifier and returned to the bed. The separated

granules are further cooled in a rectangular fluidized bed attached to the cylindrical dryer-granulator. Oversize granules are disintegrated in a side grinder, classified pneumatically and returned to the bed. Table 20.25 highlights the key operating parameters for 2.5 m in diameter granulator with the bed height about 0.3 m and the disk rotated at 450 rpm [87].

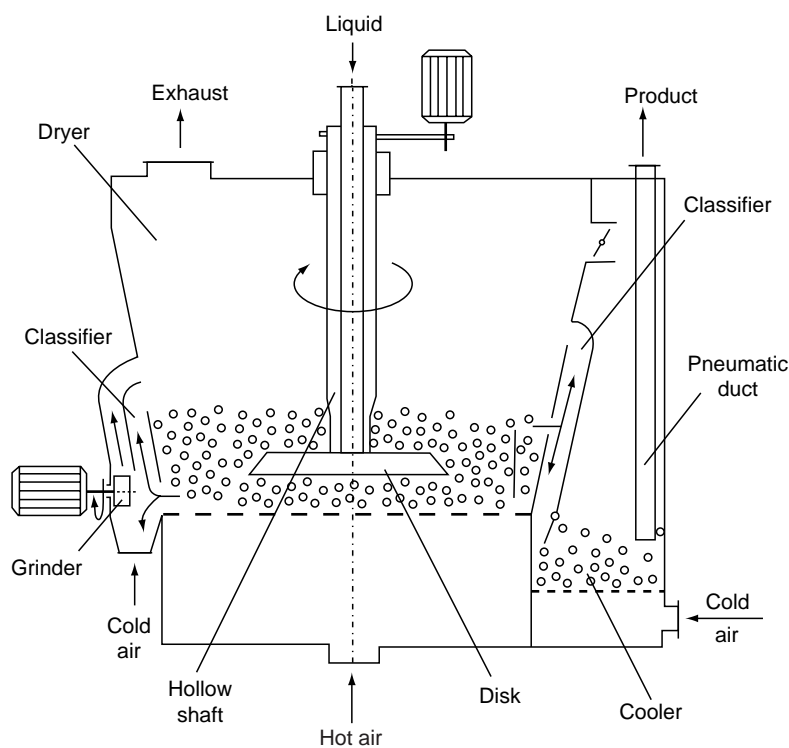


FIGURE 20.57 Dryer-granulator.

TABLE 20.25
Characteristics of Dryer-Granulator

Parameter	Skim Milk	Albumin
<i>Production rate, kg/s</i>		
Evaporated moisture	0.139	0.139
Dry product	0.1 ± 0.01	0.0334
<i>Moisture content, % w/w</i>		
Initial	40 ± 5	15
Final	4.0	8.5
Inlet air temperature, °C	115 ± 15	115 ± 5
Material temperature, °C	57 ± 3	50 ± 5
Electricity consumption, kJ/kg of water	612	552.2
Steam consumption, kg/kg of water	2.0	1.7

Source: From Marchevsky, V.N., in *Proceedings of the 12th International Drying Symposium IDS2000* (P.J.A.M. Kerkhof, W.J. Coumans, and G.D. Mooiweer, Eds.), 2000, Paper 195.

20.3.15 IR-MW FREEZE-DRYER

Freeze drying is one of the most energy-intensive, time-consuming and hence expensive drying technologies. Direct supply of thermal energy into the sublimation zone during freeze drying of capillary-porous materials appears to be the most effective method of reducing energy consumption as well as the drying time. Figure 20.58 presents a schematic diagram of a continuous vacuum freeze dryer for drying of liquid feeds such as black currant juice [88]. In contrast to the conventional batch freeze dryers with heating plates and IR radiators, a combined mode of heat supply is used here, namely IR irradiation in the freezing and partial sublimation zone (the upper chamber), and microwave heating for finish drying (the lower chamber). The liquid feed from a storage tank is pumped to the upper chamber and dispersed by an ultrasound nozzle. Because of abrupt pressure drop to about 70 Pa, the

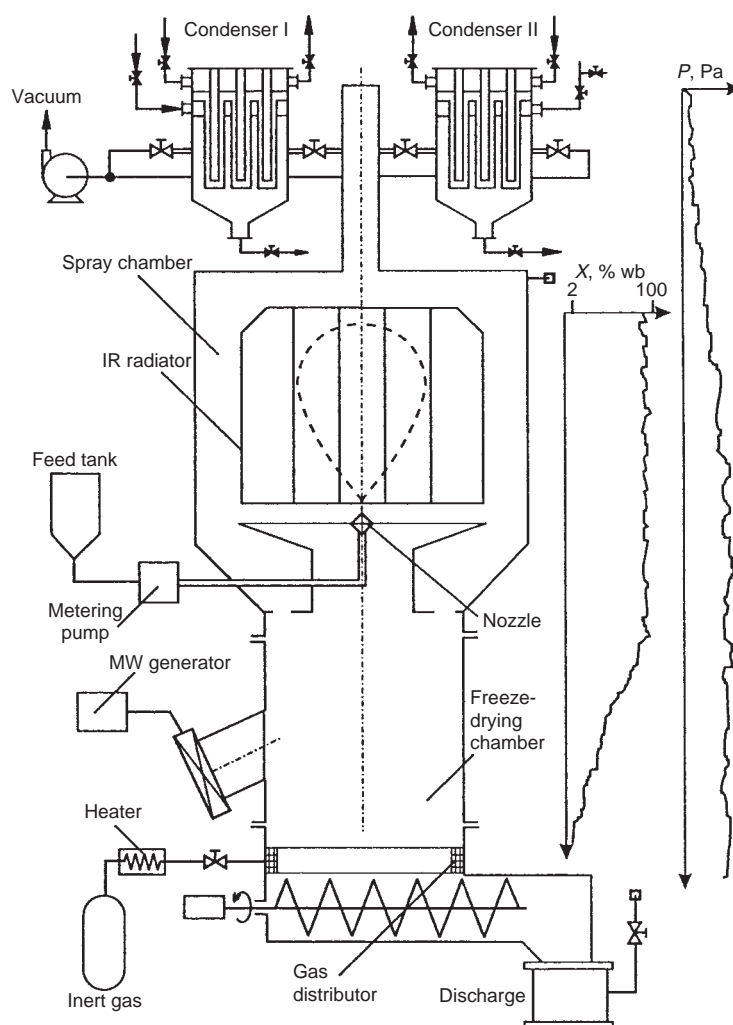


FIGURE 20.58 Continuous IR-MW-convective freeze-dryer.

dispersed material freezes rapidly and then agglomerates due to contact with a cloud of the dispersed liquid. At the same time, the granules are partially dried because of the infrared irradiation.

Frozen and surface-dry granules fall through a conical collector into the lower chamber in which a bed of particles is maintained throughout the process. Because of MW heating, finish drying by sublimation of the frozen water proceeds as the bed moves down by gravity (c.f., pressure and moisture content distribution along the dryer height in Figure 20.58). A screw conveyor is used to transport the dry particles into a vacuum lock allowing periodical discharge of the product from a continuously operated dryer.

To facilitate removal of water vapor, a small amount of inert gas (e.g., nitrogen, argon, or helium) heated electrically to 20 to 30°C is injected into the bed of dried granules. The gas stream is fed to the lower chamber through a porous metal-ceramic ring embedded in the chamber wall just above the screw conveyor. The evaporated moisture is removed from the inert gas in two alternately operated condensers [88]. Although the basic model of the dryer is designed for liquid feeds, semisolid materials such as whole or sliced berries can also be dried in such a dryer if the spray nozzle is replaced with a screw feeder. Table 20.26 compares the performance indices for several dryers for black currant [89].

20.3.16 DRYER WITH A SPRING-TYPE ROTOR

Figure 20.59 presents a schematic of the contact dryer where thermal energy supplied by conduction from the dryer wall is complemented by mechanical energy delivered by a spring-type rotor that disintegrates the wet particulate material, or breaks agglomerates when drying pasty feeds [90]. The rotor is made as a coil spring bent at 90° with ends seated in the rotating

pans. The spring is driven with a controllable speed from 500 to 1500 rpm, depending on the material characteristics and drying kinetics. The spring is typically made from a steel wire 3 to 10 mm in diameter, and the mean coil diameter varies from 20 to 130 mm depending on the dryer volume. By selecting the pitch of the spring it is possible to grind the material to the required size up to 15 mm. When fine powder should be obtained, or pasty materials are to be dried, the drying chamber can be filled with inert solids such as corundum, ceramic, or steel balls. To avoid choking of the rotor the ball diameter should be larger than the maximum distance between two adjacent coils of the bent spring. For example, in an industrial dryer 0.05 m³ with a spring made of 6-mm wire, the 6-mm steel balls are sufficient as the maximum gap between two adjacent coils is 4 mm. The minimum static bed height of inert balls securing proper circulation within the dryer chamber is typically equal to one fourth of the coil diameter. The material-to-balls mass ratio may vary from 0.5 to 1.5 kg/kg. Performance characteristics of the dryer-grinder are given in Table 20.27, taking red beet as a sample material.

The dryer can be heated with steam, hot water, or a thermal fluid. In batch operation, the cooling medium can be passed in the jacket if cooling of the dry product is required. When drying heat-sensitive materials, the dryer can also be operated under vacuum down to about 60 Pa. The dryer-grinder can also be operated in a continuous mode as well as with hot gas as the drying medium. Detailed information on various configurations of the dryer with a spring-type rotor can be found in the monograph by Shulyak [91].

20.3.17 GRINDER-DRYER

The KDS Micronex grinder-dryer [92] is essentially a grinder, which has the beneficial and originally

TABLE 20.26
Comparative Performance of Infrared, Convective and IR-MW-Vacuum Freeze Dryer for Black Currant

Parameter	Atmospheric Drying		Freeze-Drying	
	IR Band Dryer	Convective Band Dryer	Contact Heating	MW + Inert Gas
Drying time (average), h	0.5–2.5	4–6	20–30	0.5–2.5
Energy consumption, kWh/kg of water	1.0	1.85	4.5	1.5
Surface area, m ² /kg water	0.04	0.07	0.26	0.06
Initial moisture content, % wb	98	98	98	98
Final moisture content, % wb	2–3	8.0	3.5	4.0
Temperature, °C	30–60	50–110	–35 ± 40	–10 ± 120
Loss of vitamin C, %	58	91	7	4

Source: From Lebedev, D.P., Unpublished report, 2002.

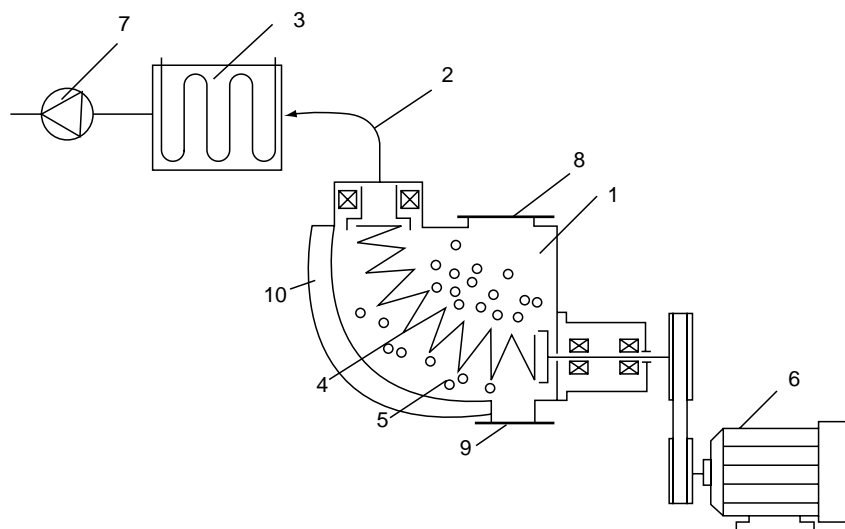


FIGURE 20.59 Schematic diagram of the dryer-grinder: 1, drying chamber; 2, vapor exhaust; 3, condenser; 4, rotor; 5, dried material–inert particles; 6, drive; 7, vacuum pump; 8, feed; 9, discharge; 10, heating–cooling jacket.

unintended serendipitous side effect of drying anything that it grinds. No supplementary heat input is required, yet it has been demonstrated to dry, for example, deinking (paper) sludge from an initial moisture content of 53% wb to a final moisture content of 22% wb, at the same time grinding it to reduce its particle size.

Figure 20.60 shows the schematic diagram of the KDS Micronex machine and the isometric sectional

view of the torus. The KDS chamber encloses a set of eight spinning chains and a stationary torus above it. The chains are spun around in a horizontal plane by a motor-driven hub. The velocity at the chain tips is 200 m/s. The top of the torus is concave and its bottom is flat. Eight radially disposed baffle plates are welded at 120° to the bottom surface of the torus. The blades provide a surface for the particles to impinge on,

TABLE 20.27
Performance Characteristics of the Dryer-Grinder for Red Beet

<i>Raw material</i>	
Size	Cubes 10 mm
Moisture content	75% wb
<i>Product</i>	
Size	0.02–0.12 mm
Moisture content	5% wb
Drying time	15 min
Inert particles	Steel balls 6 mm
Spring	54 mm in diameter, 6-mm wire
Temperature of the heating medium	75–90°C
Temperature in the condenser	(–10) – (–15)°C
Pressure	1–2 kPa
Wall-to-material heat transfer coefficient	300–700 W/(m ² K)
Evaporation capacity per heat transfer area	16–18 kg H ₂ O/m ²
Fraction of heat from mechanical energy	10%

Source: From Shulyak, V.A. and Berezyuk, D.I., in *Proceedings of the International Conference on Energy-Saving Technologies for Drying and Hygro-Thermal Processing*, Moscow, 2002.

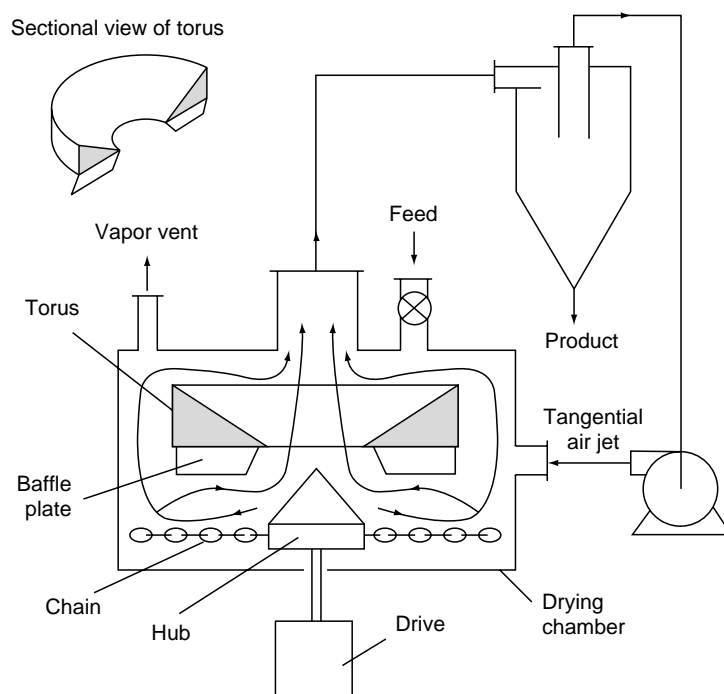


FIGURE 20.60 Schematic diagram of the KDS Micronex grinder-dryer. (Courtesy of First American Scientific Company, Delta, BC, Canada. With permission.)

hence pulverize the product. In addition, the blades direct the peripheral air to the flow through the central hole of the torus.

The raw material is fed into the KDS chamber through the inlet star valve. It slides down the concave sides of the torus and falls on top of the spinning hub. From there on, three different forces act on the material:

1. A centripetal acceleration of 4800 gees flings the material (and air) against the sides of the KDS chamber. The impact pulverizes the material. Some comminution also occurs because the centripetal force literally pulls the material apart.
2. Since the particles are accelerated along a radial path, the Coriolis force also plays a part in the break-up of the material.
3. The particles also collide against the baffle plates and the chains and with each other and get comminuted.

The kinetic energy of impact causes heating of the particles, so the moisture in them is flashed into steam. The steam escapes from the particles and then immediately recondenses into a fine mist, since the temperatures inside KDS Micronex machine never exceed 70°C. Thus, water is removed from the

material. The fine mist leaves the KDS Micronex machine through the vapor vent.

The direction of air movement in the KDS chamber is indicated by arrows. The airflow caused by the chains doubles back on itself and goes out through the cyclone duct carrying most of the material with it. In addition to the air set in motion by the spinning chains, air from the recirculating blower issues out as a jet which is tangential to the torus. This jet splits into two streams, one of which lifts up the material between the torus periphery and the KDS chamber and carries it above the torus and then on to the cyclone. The other stream assists in the evacuation of the particles through the cyclone duct, leading to the cyclone. A vapor vent acts as the exit point for the water mist, as mentioned earlier.

The dry particles are carried by the air in the cyclone duct to a conventional cyclone where the air and particles are separated. The dry particles come out through an outlet star valve at the cyclone bottom and the cleaned air leaves through the cyclone outlet on the top only to be pumped back into the KDS chamber by the recirculating blower.

Some performance data on the KDS Micronex dryer 1.4 m in diameter are shown in Table 20.28. The moisture content reported below is on wet basis.

The numbers in Table 20.28 were obtained when operating the grinder-dryer rather conservatively.

TABLE 20.28
Performance Data of the KDS Micronex Grinder-Dryer

Material	Feed Rate,kg/h	Moisture Content, % wb		Electricity Consumption, kW	Product Particle Diameter, mm
		Feed	Product		
Deinking (paper) sludge	1000	50	15	90	Fibers
Primary + secondary paper sludge (50/50 mass)	489	80	28	130	Fibers
Chicken (broiler) manure	2095	15	12	150	0.60
Wood chips	980	14	7	130	0.71
Limestone	3825	<10	<5	150	0.075

Source: From Courtesy of the First American Scientific Corporation, Delta, BC, Canada. With permission.

Most of the power consumption is due to the aerodynamic drag on the chains. Hence, reducing the power consumption is quite possible [93].

20.3.18 MW AND RF-ASSISTED DRYERS

Though dielectric drying as presented elsewhere in this handbook is a mature technology, a continuous progress is noted in RF- and MW-assisted drying as it greatly eliminates thermal lag due to heat diffusion and provides some unique features such as [3]:

- Enhanced diffusion of liquid and vapor moisture
- Coincidental temperature and mass concentration gradients
- Internal pressure gradient as an additional mass transfer driving force
- Stabilized material temperature at or below the liquid boiling point

Because of these features, a total drying cost for green lumber dried in a combination dryer with RF heating and hot air drying can be reduced even by half as compared to conventional hot air kiln [94].

An advanced technique in the RF band of an electromagnetic energy, which provides continuous tuning of the generator to the load, is the so-called 50-ohm technology [95–97]. In contrast to conventional RF generators with variable impedance, the output impedance of the quartz-driven RF generator used in this technology is fixed at 50 Ω . It permits the use of flexible coaxial cables (50 Ω) so the dryer is not physically constrained to the location of the power source. In addition, precise control over the incident and reflected power is possible since the impedance of the load can easily be measured using a spectrum

analyzer, for example. The exact value of the impedance then permits the calculation of a fitted matching unit with variable vacuum capacitors. In addition, the use of a phase and amplitude discriminator allows the system to be automated. Quartz-driven generator guarantees frequency stability and hence conforms to the authorized frequency band regulations.

A typical RF system operated with the 50-ohm technology consists of an RF generator with 50 Ω output impedance, the matching unit, and the applicator (a dryer), all linked with 50 Ω coaxial cables (Figure 20.61). Power specifications for the RF generator are defined according to the product that is heated and permissible frequency of 13.56 or 27.12 MHz. The output power of the generator can be adjusted automatically or manually by a potentiometer with the incident and reflected power that is displayed on the front panel of the generator. Because the reflected energy can destroy the RF generator, the system is equipped with a limiter, which adjusts the incident power in order to restrict the reflected power at 10% of the maximum power.

The matching unit permits the maximum of the RF energy from the generator to be transferred to the product. Such an optimum power transfer (reflected power is zero) is only possible when the input impedance of the applicator is equal to the output impedance of the generator. To match the impedance, two parameters have to be adjusted: one to adjust the phase and another to adjust the amplitude. The matching unit can be automatic or operated manually. In most of the cases, the characteristics of the material vary during the process. This requires continuous adjustments on the matching device; in such a case, the manual method is not advised. In order to have an automatic matching unit, it is necessary to measure the phase and the amplitude of the system by means of a discriminator.

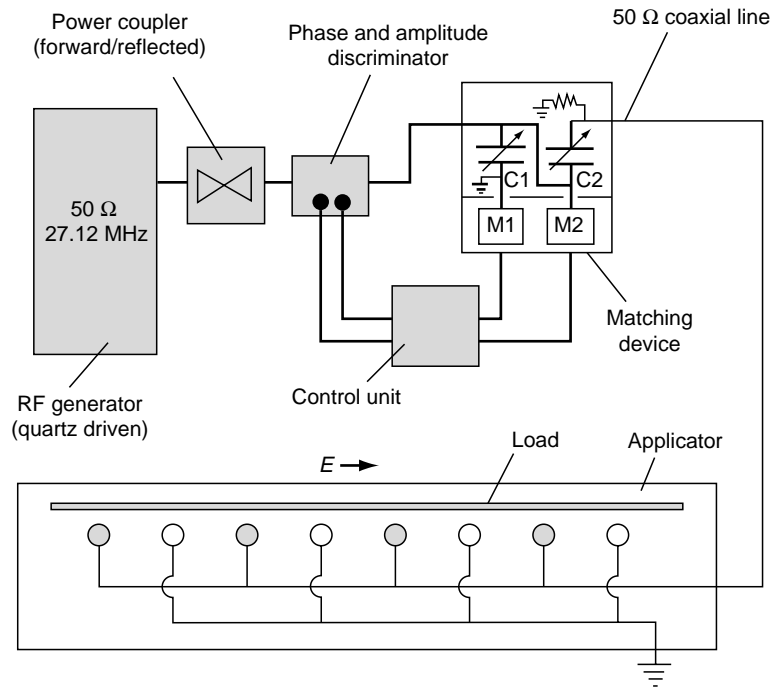


FIGURE 20.61 Basic configuration of the RF 50 Ω technology for paper drying. (Courtesy of SAIREM, France).

The matching unit has to be as close as possible to the applicator to minimize the losses.

The applicator is defined according to the material to be treated (particulate materials, sheets or bands, bulk solids), the process to be used (e.g., drying, gluing), treatment to be made (curing, surface heating, hardening), complementary energy source (hot air, infrared, heating plates), or special applications such as vapor extraction.

Essentially, the drying characteristics of the RF 50-ohm technology are the same as that for conventional RF drying. The main feature, however, is better utilization of the electromagnetic energy, which alters the drying rate and allows more uniform drying to be obtained.

The energy and drying performance of the 50-ohm technology was tested with a $190 \times 190 \times 14.3$ -mm board made of mineral and cellulose components that form the structure of foamed particles stuck in a fibrous network with pore size ranging from 1 to 100 μm [98]. The board with initial moisture content of about 1.6 kg/kg was placed between two vertical 200×200 -mm electrodes spaced by 40 (or 60) mm in a $477 \times 350 \times 340$ -mm drying cavity. The electromagnetic energy from the 1.2-kW RF solid-state generator operated at 27.12 MHz was transmitted to the applicator by an automatic matching device, yielding a constant 50-ohm impedance load. Drying experiments were performed for five levels of electrode voltage (0.8, 1.2, 1.6, 2.0, and 2.4 kV). Approximately

0.05 m^3/s of ambient air was blown through the cavity to remove the evaporated water.

Three different methods were used to quantify the RF energy dissipated in the product: thermal energy balance based on the product mass and temperature measurements; electrical losses in the matching device (insertion losses) and inductance losses calculated from the flow rate and temperature rise of the cooling water.

Figure 20.62 presents a comparison of the energy transferred to the product determined with these three methods (for the case of worst discrepancy). These estimates have all about the same shape, which indicates that any one of these methods can be used for the determination of the transmitted power. Figure 20.63 presents the normalized RF power and drying rate plotted against the product moisture content for several values of the electrode voltage. The normalized values are here the ratios of the actual value to the maximum value obtained in the experiment. Characteristically, these curves are of the same shape, which indicates a close relation between the RF power and the drying rate and therefore confirms favorable energy performance of the 50-ohm technology, as far as transmission and matching issues are concerned.

The main advantages of this technology are claimed to be as follows [96]:

- The use of the quartz-driven generators yields high stability of frequency

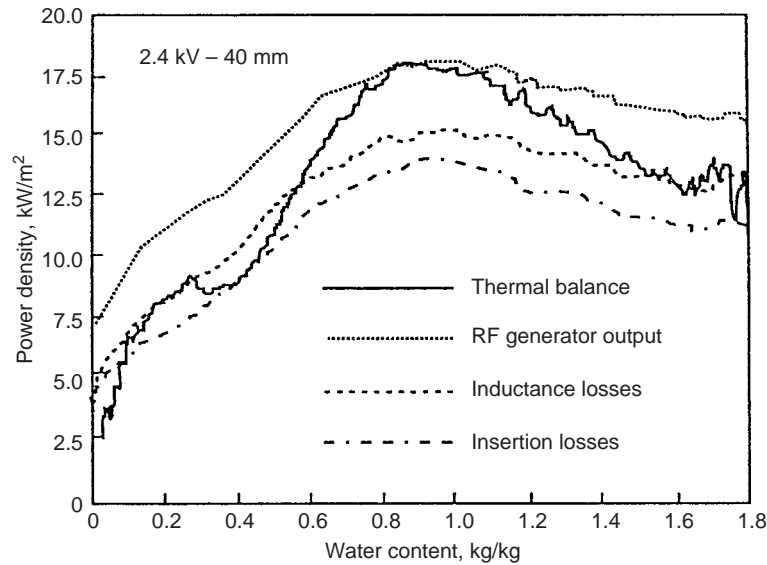


FIGURE 20.62 RF heating capacity as a function of the moisture content. (From Dostie, M. and Navarri, P., *Drying'94. Proceedings of the IDS'94*. (V. Rudolph and R.B. Keey, Eds.), 1994, pp. 607–614.)

- Reduction of RF radiation emitted by the applicator due to permanent tuning
- Limitation of flash risk
- Reduced power consumption as compared to the traditional RF equipment
- Improved operation (simplified control, reduced adjusting time, complete automation, versatile and remote command of the system, etc.)

The RF technology appears to be well suited as a heating source for industrial applications that require

high power and short process times, such as paper drying. Flexibility of the 50-ohm technology enables one generator to be used in sequential mode with two or three applicators. In terms of the costs of the RF drying systems, the investments are clearly very high, but the payback is relatively short.

Except for some special applications like drying with sol-gel transformations [99], dielectric drying is especially advantageous when combined with other drying methods. Examples are hybrid technologies such as RF-assisted heat pump drying, MW-convective

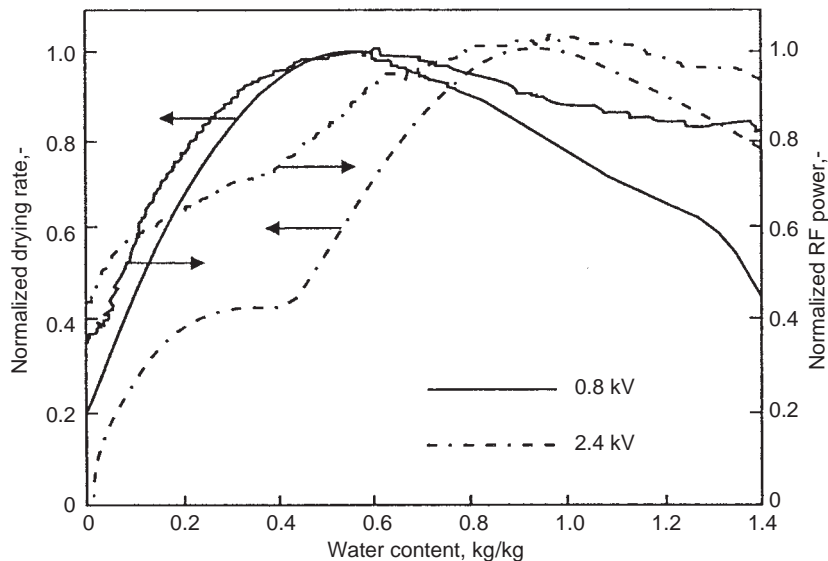


FIGURE 20.63 Normalized RF power and drying rate as a function of the moisture content. (From Dostie, M. and Navarri, P., *Drying'94. Proceedings of the IDS'94*. (V. Rudolph and R.B. Keey, Eds.), 1994, pp. 607–614.)

TABLE 20.29
Characteristics of the RF-Vacuum Dryer

Operation	Batch (Wood); Continuous (Food Products)
Chamber capacity, m ³	5–75
Throughput (wood), m ³ /year	Up to 70,000
Frequency, MHz	6.78 (wood); 40.68 (food products)
RF power output, kW	20–300 kW
RF power density, kW/m ³	Up to 5
Vacuum	Down to 3300 Pa
Energy consumption, ^a kWh/kg H ₂ O	1.3 (hygroscopic materials) 1.0 (nonhygroscopic materials)

^a0.7 kWh/kg free water.

Source: Courtesy of HeatWave USA. McMinnville, OR. With permission.

drying, MW-vacuum drying, and MW-superheated steam drying that have been verified in laboratory and pilot tests [3].

One of the commercially successful hybrid dryers built on the 50-ohm technology is the RF-vacuum dryer for bulk solids such as sawn wood [100]. At lower drying temperatures, resulting from reduced boiling point under vacuum, wood retains mechanical strength so it is less susceptible to drying defects such as checking. Because moisture is removed from the kiln as water vapor, there is a potential for increased energy efficiency as heat is not lost with the vented hot and humid air. An added feature of RF heating for vacuum drying of wood is a possibility to stack lumber in a solid pile (without stickers). This not only reduces the load volume but also allows the use of the load-restraining system [101], which prevents wood from any movement during drying and thus greatly reducing twist, bow, cap, crook, warp, and the like defects. Characteristically shaped electrodes provide the most uniform *E*-field throughout the bulk of a drying material [102]. The end-point detection system

integrated in the control circuit allows the average moisture content and the time to complete drying to be monitored over the entire drying cycle. Provided with an access to Internet or Intranet, the control system secures remote supervising of the kiln operation. Once the target moisture content is reached, the kiln automatically stops and unloads the kiln charge.

As applied to drying of solid wood, the RFV technology has the following production advantages [100]:

- The drying rate is up to ten times higher than that in a conventional kiln. An added benefit is no stickering of the lumber packages.
- Accelerated drying for smaller kiln charges. The ability to dry at the same time different species and dimensions allows for greater production and scheduling flexibility. The RFV kilns can also be used for pasteurization and fixation of preservatives.
- Improved product quality; the RFV-dried wood retains its natural color with the appearance as of freshly sawn wood. In addition, there is no heat discoloration or brown staining during drying. Chemical oxidation due to contact with drying air is practically eliminated.
- Stress-free drying at low temperature substantially reduces shrinkage and this allows a reduction in green target sizes and improved lumber yields.

Table 20.29 provides performance characteristics of the RF-vacuum dryer whereas Table 20.30 compares drying time and electricity consumption for several wood species dried in the radio frequency vacuum wood kiln.

20.4 CONCLUSIONS

This chapter focused on novel drying technologies based on one or more concepts not found in corresponding conventional drying technologies.

TABLE 20.30
Performance of the RFV Kiln for Lumber

Parameter	Air-Dry Lumber (2" Hemfir) 22–12%	Green Lumber		
		2" Hemfir 55–10%	4" × 4" White Pine 40–10%	1" and 2" Birch 505–508%
Drying time, h	5	36	36	60
Total electricity, kWh/Mfbm	64	546	356	863

1 m³ = 0.424 Mfbm.

Source: Courtesy of HeatWave USA. McMinnville, OR. With permission.

All techniques mentioned here have found industrial-scale applications or have been field-tested. Most of the feedstocks considered are in the form of liquids, pastes, sludges, or granular solids, except for the fluid bed dryer with active solids (for drying of leather) and the Remaflam dryer for textiles. New technologies for drying paper are covered in the chapter on pulp and paper drying. This chapter effectively updates several chapters in this handbook.

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NOMENCLATURE

b	width, m
d	particle diameter, m
D	diameter, m
f	frequency, 1/s
H	bed height, m
H	distance between accelerating ducts, m
L	dimensionless distance
n	number of stages,—
n	rotational speed, 1/s
N	number of blades
P	pressure, Pa
ΔP	pressure drop, Pa
s	pitch, m
t	time, h
T	temperature, K (°C)
u	velocity, m/s
W	mass flow rate, kg/s
x	coordinate, m
X	moisture content (dry basis), kg/kg or %
X'	moisture content (wet basis), %

SUBSCRIPTS

e	equivalent
ev	evaporation
f	final

g	gas
i	inlet
m	material
o	outlet
r	residence
0	initial/flow duct
p	particle (droplet)
1	inlet (initial)
2	outlet (final)

SUPERSCRIPTS

*	reference parameter
---	---------------------

GREEK LETTERS

β	angle, °
ε	bed voidage
η	efficiency, %

ABBREVIATIONS

db	dry basis
dB	decibel
E	electric field
dm	dry matter
ppm	parts per million
rpm	revolutions per minute
v/v	volume by volume
wb	wet basis
w/w	weight by weight

LIST OF ACRONYMS

C–G	Carvar–Greenfield (process)
DCCA	drying control chemical additive
FB	fluid bed
IS	impinging stream
ISD	impinging stream dryer
MVR	mechanical vapor recompression
PCD	pulsed combustion dryer
PFB	pulsed fluid bed
PTFE	polytetrafluoro ethylene (Teflon)
TLV	threshold limit value
VFB	vibro fluid bed
VRB	vibro-rotational bed
VSD	vortex spray dryer

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Part III

Drying in Various Industrial Sectors

21 Drying of Foodstuffs

Shahab Sokhansanj and Digvir S. Jayas

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21.1 INTRODUCTION

The removal of moisture from solids is an integral part of food processing. Almost every food product is dried at least once at one point of its preparation. The main objectives of dehydration are summarized as follows [32].

21.1.1 EXTENDED STORAGE LIFE

A dry food product is less susceptible to spoilage caused by the growth of bacteria, molds, and insects. The activity of many microorganisms and insects is inhibited in an environment in which the equilibrium relative humidity is below 70%. Likewise, the risk of unfavorable oxidative and enzymatic reactions that shorten the shelf life of food is reduced.

21.1.2 QUALITY ENHANCEMENT

Many favorable qualities and nutritional values of food or feed products may be enhanced by drying. Palatability is improved, and likewise digestibility and metabolic conversions are increased. Drying also changes color, flavor, and often the appearance of a food item. The acceptance to that change varies by the end user.

21.1.3 EASE OF HANDLING

Packaging, handling, and transportation of a dry product are easier and cheaper because the weight and the volume of a product are less in its dried form. A dry product flows easier than a wet product; thus gravity forces can be utilized for loading and unloading and short-distance hauling.

21.1.4 FURTHER PROCESSING

Food products are dried for improved milling, mixing, or segregation. A dry product takes far less energy than a wet product to be milled. A dry product mixes with other materials more uniformly and is less sticky compared with a wet product.

21.1.5 SANITATION

Drying has also been used as a means of food sanitation. Insects and other microorganisms are destroyed during the application of heat and moisture diffusion. The sanitation aspect of drying is a time-temperature phenomena [25]. The temperature should be at least 60°C for a short duration of 3 to 5 min. Lower temperatures, to a minimum of 48°C, can be used for disinfestation but the treatment duration should last at least for 24 h or longer. Disinfestation in high temperature rotary and tunnel dryers has been studied and results have been published by Sokhansanj et al. [26] and Sokhansanj and Wood [24].

21.2 MOISTURE IN FOODS

The volatile part of a food item can be termed moisture. Moisture, in the form of water molecules, is bonded to various parts of the product in varying ways as follows: (a) ionic groups, such as carboxyl and amino acids and (b) hydrogen groups, such as hydroxyl and amides. In high-moisture foods in which moisture contents are more than 50% wet basis, unbound free water exists in the interstitial pores and in intercellular spaces. The descending

order of difficulty to remove water from the product also follows the above order.

A food product is in equilibrium with its surroundings when its internal vapor pressure is in equilibrium with the outside vapor pressure. The moisture content of the product at this stage is called the equilibrium moisture content (EMC). The corresponding vapor pressure in surrounding air at the same temperature is called the equilibrium vapor pressure. The ratio of the equilibrium vapor pressure to the saturation vapor pressure is known as the equilibrium relative humidity (ERH), or water activity.

A plot of EMC versus ERH usually has a sigmoid (S) shape. The reason is that the affinity of the solid to moisture and the ease of moisture adsorption or desorption depend upon the way moisture bonds to the solid [19]. Starting from a bone-dry solid exposed to a humid environment, in a range of 5–10% moisture content, a single layer of water molecules is formed. The next stage is the formation of a multilayer of water molecules, during which the slope of the isotherm curve is gradual. The last stage is filling of the capillary pores by condensed water.

The EMC varies with temperature. Several plots of EMC versus ERH are drawn for various temperatures. Each plot is often called an isotherm. Samples of isotherms of various food items are given in Figure 21.1.

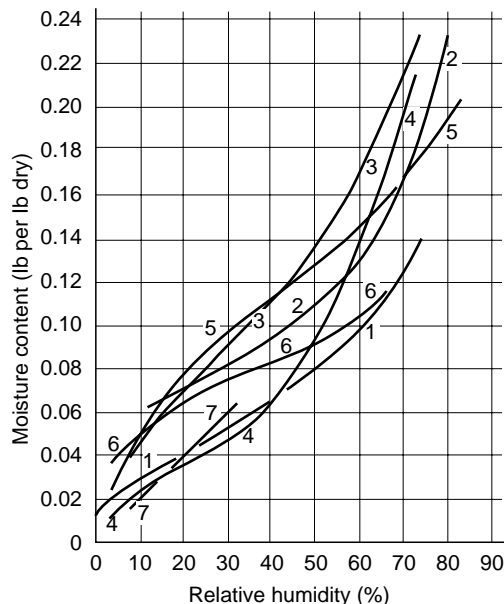


FIGURE 21.1 Selected water sorption isotherms (1. egg solids, 10°C; 2. beef, 10°C; 3. codfish, 30°C; 4. coffee, 10°C; 5. starch gel, 25°C; 6. potato, 28°C; 7. orange juice). (From Van Arsdel, W.B. and Copley, M.J., *Food Dehydration*, Vol. 1, AVI Publ. Co., Inc., Westport, CT, 1973. With permission.)

Food demonstrates a hysteresis phenomenon during adsorption and desorption processes. Irreversible physical and chemical changes are responsible for this behavior. Often the EMC of a food product in a given condition is varied by 2% points, depending on whether EMC is obtained by removing or adding moisture.

EMCs of foods generally are found experimentally. The following equation has been found to fit the experimental data [20]:

$$\frac{P_v}{M(P_s - P_v)} = \frac{1}{M_s C} + \frac{C - 1}{M_s C} \frac{P_v}{P_s} \quad (21.1)$$

where P_v and P_s are water vapor pressure and saturated vapor pressure, respectively. C is called the energy constant and it is temperature-dependent [31]. M and M_s are the moisture and the EMC, respectively. These values can be found by performing drying experiments on a thin layer of sample of the product and plotting $P_v/M(P_s - P_v)$ versus P_v/P_s . The order of magnitude of C and M_s are 13.0 and 0.10, respectively. M_s in Equation 21.1 is considered constant at which a single layer water molecules coat the surface of the solid particles.

An equation that has been widely used in the dehydration of cereal grains is of the form

$$M_e = E - F \ln[-R(T + C) \ln(RH)] \quad (21.2)$$

Values of the constants E , F , and C are given in Table 21.1. R is the universal gas constant and its value is 1.987 in SI units.

The role of moisture in food drying and storage is expressed in terms of water activity. Water activity in

TABLE 21.1
Coefficients for Use in EMCE RH (Equation 21.2)

Product	C	E	F
Barley	91.323	0.368149	0.402787
Beans	120.098	0.480920	0.066826
Corn	30.205	0.379212	0.058970
Rice	35.703	0.325535	0.046015
Sorghum	102.849	0.391444	0.050970
Soybean	24.576	0.375314	0.066816
Wheat, durum	112.350	0.415593	0.055318
Wheat, hard	50.998	0.395155	0.056788
Wheat, soft	35.662	0.308163	0.042360

Source: From ASAE, *Standards 1984*, American Society of Agricultural Engineers, St. Joseph, Michigan, 1984. With permission.

a moist food is defined in a similar manner as the relative humidity is defined in moist air, that is, the ratio of vapor pressure to the saturated vapor pressure at the same temperature.

Water activity relates to the chemical activity of moisture in the food during drying and storage. Oxidation activity is only possible at water activities higher than 0.4 and the rate of inactivation of other organisms requires a water activity of 0.7 or lower. Some enzymatic activities may continue at low levels of water activities of 0.1–0.3 but their reaction rate decreases at low water activities.

Water activity in foods is related to moisture content using Guggenheim–Anderson–de Boer (GAB) equation:

$$M = \frac{M_s C K a_w}{(1 - K a_w)(1 - K a_w + C K a_w)}$$

where C , K , and M_s are constants. Their value can be estimated from experimental data of M versus a_w as follows:

$$\frac{1}{M} = \beta_0 + \beta_1 \frac{1}{a_w} + \beta_2 \frac{1}{a_w^2}$$

$$\beta_0 = -\frac{1}{C M_s}, \quad \beta_1 = \frac{1}{C K M_s} - \frac{2}{C^2 K M_s}, \quad \beta_2 = \frac{1}{M_s (C K)^2}$$

21.3 AIR–VAPOR RELATIONSHIP

Dry air and water vapor exert a certain pressure upon each other when they are mixed. These pressures are called partial pressures. The difference in partial pressure of water vapor in the air and the pressure of the moisture in the product is the driving force for drying.

The interrelationships between air and water vapor are called psychrometric properties [3]. Changes in these properties are shown by a psychrometric chart. A sample of the chart is given in Figure 21.2. The psychrometric terms given in the chart are defined as follows.

21.3.1 DRY BULB TEMPERATURE

Dry bulb temperature is the air temperature indicated by an ordinary thermometer. It is given on the horizontal axis of the psychrometric chart.

21.3.2 HUMIDITY RATIO

Humidity ratio or absolute humidity is the ratio of the weight of water vapor to the weight of dry air. Humidity ratios are on the vertical axis of the psychrometric chart.

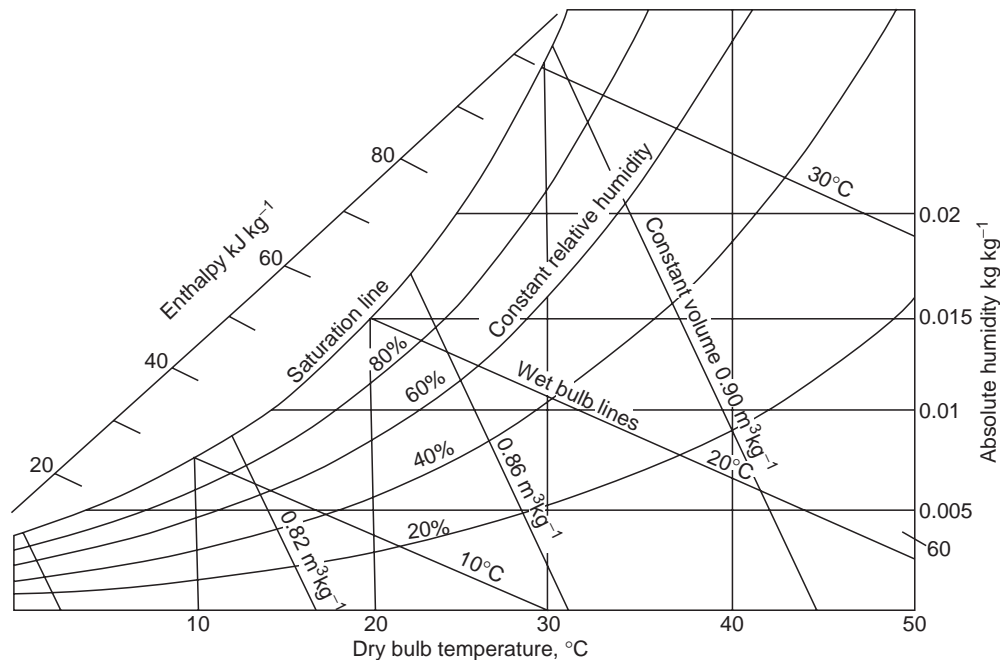


FIGURE 21.2 Sample of the psychrometric chart.

21.3.3 RELATIVE HUMIDITY

The ratio of vapor pressure to saturation vapor pressure is called relative humidity. Relative humidity is an indication of the maximum moisture that the moist air can hold at a given temperature. The 100% relative humidity line constitutes the extreme left-side boundary of the psychrometric chart.

21.3.4 DEW POINT

Dew point temperature is the temperature at which condensation occurs when the air is cooled at a constant humidity ratio and at constant pressure. Dew point is read on the dry bulb axis corresponding to the saturation relative humidity curve.

21.3.5 WET BULB TEMPERATURE

Wet bulb temperature is the temperature shown by a thermometer with a liquid container that is wrapped by a dampened cloth and exposed to the moving air. Air must blow at a speed of 5 m/s (18 km/h) over the moistened cloth in order to obtain a correct wet bulb temperature. Wet bulb temperatures are the slanted lines on the psychrometric chart.

21.3.6 ENTHALPY

Enthalpy is the heat content of the moist air. Heat content is based upon a convenient 0°C. Enthalpy lines are almost parallel to the wet bulb lines on the psychrometric chart.

21.3.7 SPECIFIC VOLUME

Specific volume is the volume of each unit weight of the moist air. The utility of a psychrometric chart in an adiabatic dehydration process is shown in Figure 21.3. Point 1 is the ambient air condition. Point 2 is the air after heating to a certain temperature, T_2 . Point 3 is the exit condition of drying air after it has passed through the dryer. Note that in the drying process, air has been cooled while taking up moisture from the product. Often the line between points 2 and 3 is a straight line and is parallel to the wet bulb temperature.

A psychrometric chart is quite useful when used within its limitations. For instance, the chart in Figure 21.2 is valid only for atmospheric pressure. It also does not handle heats of formation and crystallization or temperature of the food. When vapors other than water are involved, a separate chart must be used.

21.3.8 PSYCHROMETRIC CALCULATIONS

Equations used to compute properties of moist air are as follows:

Saturation vapor pressure P_s as a function of absolute temperature T can be given as

$$\ln(P_s) = 31.9602 - \frac{6270.3605}{T} - 0.46057 \ln(T) \times 255.83 \\ \leq T \leq 273.16$$

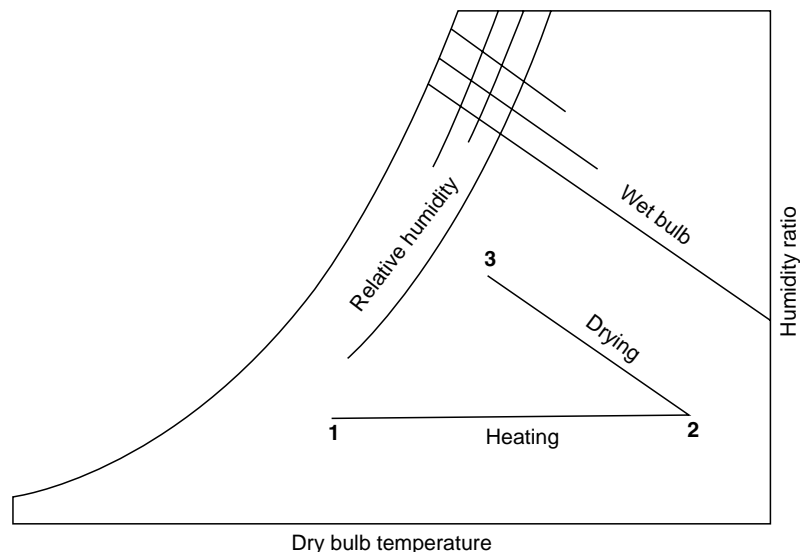


FIGURE 21.3 Adiabatic drying on a psychrometric chart.

$$\ln(P_s/R) = \frac{A + BT + CT^2 + DT^3 + ET^4}{FT - GT^2}$$

$$\times 273.16 \leq T \leq 533.16$$

$$R = 22105649.25, \quad A = -27405.526, \quad B = 97.5413$$

$$C = -0.146244, \quad D = 0.12558 \times 10^{-3}$$

$$E = -0.48502 \times 10^{-7}, \quad F = 4.34903$$

$$G = 0.39381 \times 10^{-2}$$

Humidity ratio, H

$$H = \frac{0.6219P_v}{P_{\text{atm}} - P_v}, \quad 255.38 \leq T \leq 533.16, \quad P_{\text{atm}} = 101,330 \text{ Pa}$$

Specific volume

$$V_{\text{sa}} = \frac{287T}{P_{\text{atm}} - P_v}, \quad 255.38 \leq T \leq 533.16, \quad P_{\text{atm}} = 101,330 \text{ Pa}$$

Relative humidity, RH

$$\text{RH} = \frac{P_v}{P_s}$$

The units of T are kelvin and pressures are in pascal.

One usually has the initial air conditions specified in terms of the relative humidity and temperature. Drying calculations are carried out in terms of the humidity ratio. Atypical use of psychrometric equations may follow the following sequences: use T to calculate P_s , use RH and P_s to calculate P_v , and use P_v and P_{atm} to calculate the humidity ratio H . For more equations relating other properties of moist air, see American Society of Agricultural Engineers (ASAE) standards [25].

21.4 DRYING MODELING AND CALCULATIONS

21.4.1 GENERAL

A mathematical model with which the dryer and process parameters can be studied is of extreme utility to the industry. A reliable model often will prevent or minimize costly mistakes in prototype development. The model also can be utilized for the control of process, specifically in adaptive and feed-forward control strategies. With the proliferation of low cost and powerful computers, the simulation models are useful. In this chapter, the emphasis is on the presentation of formulas and equations rather than the conventional charts and graphs in drying calculations.

The following is a list of significant parameters that influence the performance of a dryer [2]:

1. Process air variables
 - (a) Airflow rate
 - (b) Drying air temperature
 - (c) Drying air humidity ratio
2. Product variables
 - (a) Product throughput
 - (b) Initial and final moisture contents
 - (c) Material sizes and size distribution
3. Dimensional variables
 - (a) Width, height, or diameter of the dryer
 - (b) Length of the dryer and the number of passes
 - (c) Dryer configuration

These variables can be controlled or adjusted. Many variables, such as aerodynamic properties of the drying materials, exposed surface area of the product, drying rate characteristics, and the EMC, are specific and cannot be controlled.

A comprehensive drying model must include all the aforementioned variables. It must include the psychrometric and thermal properties of air, vapor, and liquid water. The development of a drying simulation model is described in the following. Simplifying assumptions and limitations inherent in the model are as follows:

1. A uniform feed rate, with a uniform moisture content
2. A uniform size product
3. No heat loss or heat gain through the dryer walls
4. No external heat and mass fluxes other than those between the air and the material in the dryer

21.4.2 HEAT BALANCE EQUATION

Heat balance over a layer of drying particles over the time interval dt , with 0°C as the reference temperature for enthalpy, is written as [29]:

Energy lost by air = energy gained by material

$$AG \, dt(i - i') = Ad \, dz[(C_p + MC_w)(T'_g - T_g) + dMC_w T'_g] \quad (21.3)$$

Refer to the Nomenclature for definitions.

The enthalpy i can be expressed as

$$i = 1005T_a + X_a(1820T_a + 2,501,000) \quad (21.4)$$

Let

$$F = \frac{d \, dz}{G \, dt} \quad (21.5)$$

Then

$$\begin{aligned} T'_a(1005 + 1820X'_a) &= T_a(1005 + 1820X_a) \\ &+ T'_g[F(C_p + M'C_w)] \\ &+ T_g\{-F[C_p + (M' - dM)C_w]\} \\ &+ 2,501,000(X_a - X'_a) \end{aligned} \quad (21.6)$$

21.4.3 HEAT TRANSFER EQUATION

The heat transfer equation describes what happens to the heat that is transferred from drying air to material. This heat raises the temperature of solids and evaporates moisture from the product:

$$\begin{aligned} Hdt \frac{(T_a + T'_a) - (T_g + T'_g)}{2} \\ = d[(C_p + M'C_w)(T'_g - T_g) + (-dM)(i'_v - C_wT_g)] \end{aligned} \quad (21.7)$$

where the enthalpy of the water vapor (i'_v) at T'_a is given by

$$i'_v = 1820T'_a + 2,501,000 \quad (21.8)$$

21.4.4 MOISTURE BALANCE

Taking a moisture balance over a time interval dt for the layer dz yields

$$A \, dz \, d(M - M') = AG \, dt(X'_a - X_a) \quad (21.9)$$

Equation 21.6 through Equation 21.9 must be solved simultaneously for a particular configuration of dryer. The configuration may involve stationary batch, concurrent flow, or counterflow, or it may be a crossflow dryer. In each instance, the values of dz and dt are chosen such that the model predicts the physical setup. Equation 21.6, Equation 21.7, and Equation 21.9 describe the air temperature T_a , air humidity X_a , grain moisture content M , and grain temperature T'_g . An additional relationship is needed to describe the drying rate of material. The drying rate will be described in the following section.

21.4.5 DRYING RATES

Moisture is transferred from inside a moist material to the outside surface, where it evaporates. High-moisture foods with a moisture content of more than 50% wet basis may demonstrate two distinct drying rates: (a) a constant drying rate and (b) a falling drying rate [21]. The curve in Figure 21.4 depicts these two periods. The methods of estimating drying rates for each period are different. The moisture content at which the food material demonstrates a change from constant rate drying to falling rate drying is called the critical moisture content. There may be several critical moisture contents for a particular food item.

21.4.6 CONSTANT RATE DRYING PERIOD

During constant rate drying, moisture is always available at the surface such that resistance to moisture removal is only the rate at which the moisture can evaporate. The constant rate drying period is described by the relationship

$$\frac{dM}{dt} = \frac{HA(T_a - T_w)}{L} = K_m A(X_w - X_a) \quad (21.10)$$

where H and K_m are coefficients that describe heat transfer and moisture transfer conditions at the surface. Heat transfer and moisture transfer are physical phenomena that are similar in their mathematical representations. Therefore, the relationships similar to those

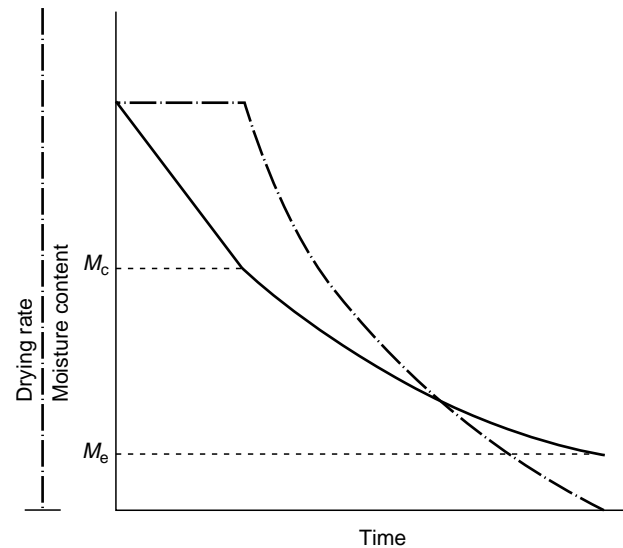


FIGURE 21.4 Drying curve for high-moisture foodstuff. (Sweeney, J.P. and Marsh, A.C., *J. Am. Diet. Assoc.* 59, 238, 1971. With permission.)

estimating H values can be used to find the K_m values. Coefficients H and K_m are related to each other by Lewis number Le as

$$Le = \frac{H}{K_m C_p} \quad (21.11)$$

Under conditions in which pressures are in the range of atmospheric (about 100 kPa) and temperatures are under 100°C, Le is nearly equal to 1. Hence, K_m can be found from H values as

$$K_m = \frac{H}{C_p} \quad (21.12)$$

The heat transfer H can be estimated using equations published in standard textbooks [30] (see Section 21.4.8).

Several empirical formulas for drying rate during a constant rate period have been developed with experimental data. For example, the following experimental relationships are given in high-temperature alfalfa drying. When $T_a > 200^\circ\text{C}$,

$$\frac{dM}{dt} = -K_0 = 0.03066 + 0.0004113T_a \quad (21.13)$$

Equation 21.13 signifies the fact that at temperatures beyond 200°C, the rate of drying is constant as long as the drying temperature remains constant.

21.4.7 FALLING RATE DRYING PERIOD

After all the water at the surface of the material has been exhausted, the moisture is diffused from the internal parts of the product to the surface. The amount of water at the surface becomes progressively scarce. As a result the drying rate will be slower as time progresses. The following relationship can be used to describe the rate of drying:

$$\frac{dM}{dt} = -KA \frac{dP}{dx} \quad (21.14)$$

where K is a moisture transfer (diffusion) coefficient and dP/dx is the driving force for the moisture movement in terms of water vapor pressure within the product. The moisture can be in the form of vapor or of liquid or a combination of these forms.

It is difficult to solve Equation 21.14 for many products. A modified form of this equation is presented as

$$\frac{dM}{dt} = \frac{\pi^2 D}{4L_x^2} (M - M_e) \quad (21.15)$$

Diffusion coefficient D is constant for the entire falling rate period. L_x is the thickness of the solid. A variation of Equation 21.15 is proposed when the ratio $(M - M_e)/(M_i - M_e)$ is greater than 0.6, as

$$\frac{dM}{dt} = \frac{H(T_a - T_w)}{2dL_x L} \frac{M - M_e}{M_i - M_e} \quad (21.16)$$

The applicability of Equation 21.16 is limited to slab or geometrically well-defined products. A much simpler expression for the average moisture content of food material is written as

$$\frac{dM}{dt} = -k(M - M_e) \quad (21.17)$$

In writing Equation 21.17, two assumptions have been made: (a) the distribution of moisture within the product is uniform and (b) the drying rate is much dependent on the drying constant k and the equilibrium moisture M_e . The value of k must be found experimentally. Usually a thin layer of product is fully exposed to a highly controlled hot air. Equation 21.17 is fitted to the experimental drying data.

A solution to Equation 21.17 takes the form of $M_r = \exp(-kt)$ or its variation $M_r = \exp(-kt^n)$. ASAE [28] publishes values of k and n for a number of foods (ASAE Standard S448 — Thinlayer Drying of Agricultural Crops). ASAE [28] also describes a standardized experimental technique for estimating the drying constants k and n .

When the assumption of “uniformity of moisture in the product” is no longer valid because of the product size, then the following procedure must be used to calculate drying rate and drying times. As it was mentioned earlier, moisture transfer is analogous to heat transfer. Therefore, the well-established methods of heat transfer calculations can be utilized for drying (moisture transfer) calculations. For example, the moisture gradients can be estimated using moisture transfer graphs similar to those of Heisler charts for heat transfer. The charts for finding moisture content at the center of an infinite slab, cylinder, and sphere are given in Figure 21.5. The use of these charts has been demonstrated by numerical examples in Ref. [5].

As an alternative to the charts, empirical equations have been developed. The data for the development of these equations are essentially those of the charts. The equations for use are as follows. For a slab-shaped food product,

$$M_c = R_p \exp(-S_p F_0) \quad (21.18)$$

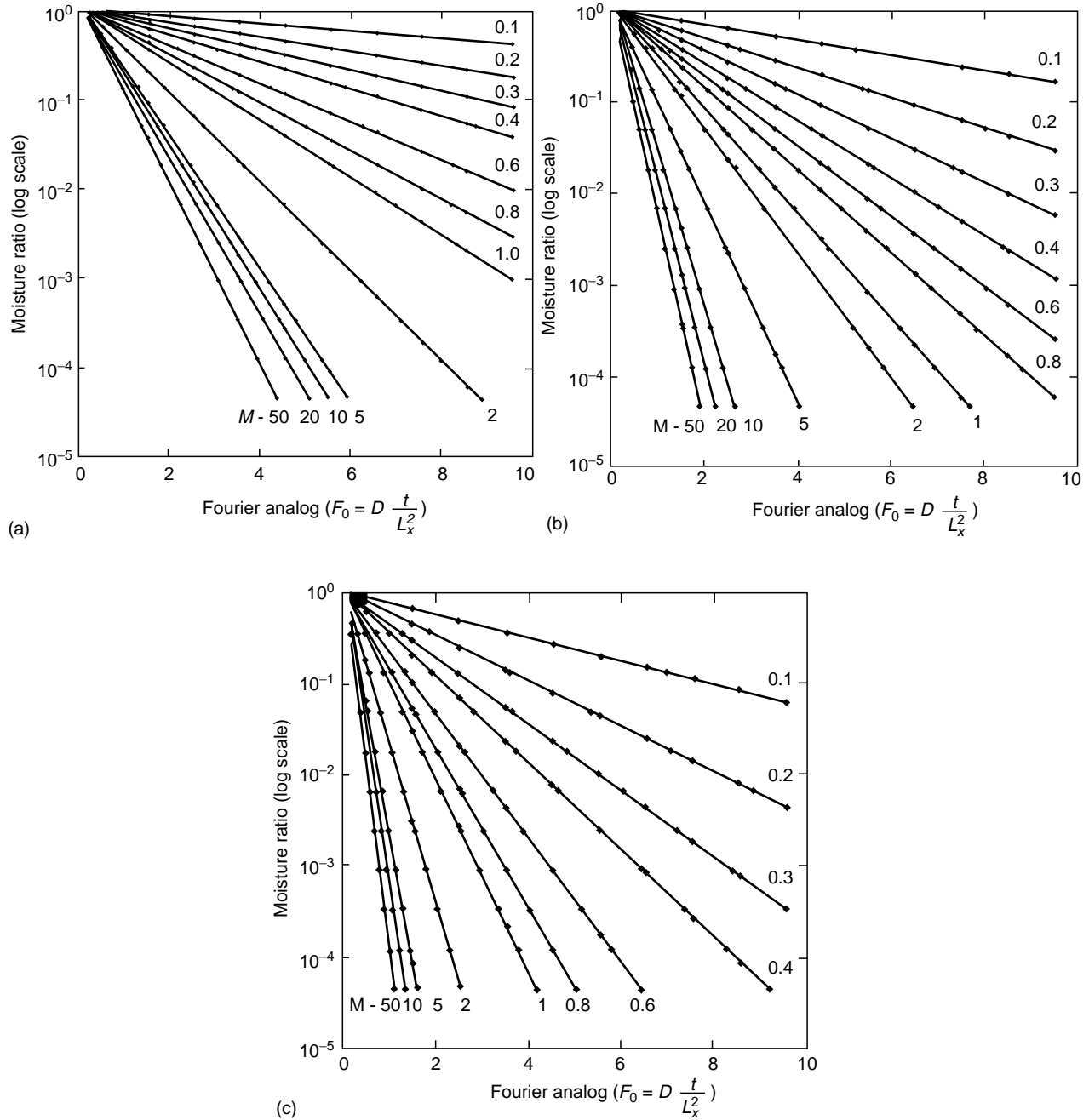


FIGURE 21.5 Moisture content distribution during dehydration, at the center of (a) an infinite slab; (b) a cylinder; and (c) a sphere. (From Ramaswamy, H.S. and Lo, K.V., *Simplified Relationships for Moisture Distribution during Drying of Regular Solids*. ASAE Paper No. 81-103. American Society of Agricultural Engineers, St. Joseph, Michigan, 1981. With permission.)

For a cylindrical food product,

$$M_c = R_c \exp(-S_c F_0) \quad (21.19)$$

For a spherical food product,

$$M_c = R_s \exp(-S_s F_0) \quad (21.20)$$

where M_c is the moisture content dry basis at the center of the slab, cylinder, or sphere. Other constants are defined as

$$R_p = 0.2884A(M) + 0.007927A(M^2) - 0.05287A^2(M^{0.5}) + 0.01696A^2(M/3) - 0.09581[A(M^2)]^{0.5} + 1.0018 \quad (21.21)$$

$$S_p = 0.4608A(M) - 0.08319A(M^2) + 0.03752A(M^3) + 0.8153A(M/3) + 0.2439A^2(M^{0.5}) + 0.001745 \quad (21.22)$$

$$R_c = 0.7398A(M/3) - 0.01723A(M^3) - 0.2151A^2(M/3) + 1.004 \quad (21.23)$$

$$S_c = 0.6517A(M) + 2.3186A(M/3) + 0.5124A^2(M^{0.5}) + 0.004211 \quad (21.24)$$

$$R_s = 1.0847A(M/3) - 0.2204A^2(M^{0.5}) - 0.1222A^2(M/3) + 0.1041[A(M)]^{0.5} + 0.9806 \quad (21.25)$$

$$S_s = 0.1970A(M^2) + 51,772A(M/3) + 0.6585A^2(M^{0.5}) + 0.3960A(M)A(1/M) + 0.003257 \quad (21.26)$$

where A is arctangent function, F_0 is $Dt/(L_x)^2$, and M is mass transfer resistance ratio defined as $(Dd/K_m L_x)$ and it is similar to (K/HL_x) in heat transfer. Equation 21.18 through Equation 21.26 are valid for $0.1 < M_c < 40$. For a full description, see Ref. [4].

21.4.8 HEAT TRANSFER COEFFICIENT

The heat transfer coefficient describes the rate of heat transfer between a food product and the flowing gas. It is mainly dependent on the gas flow and solid configuration. For a laminar flow over a flat food product, Nusselt number, Nu , can be estimated by

$$Nu = 0.664(Re)^{0.5}(Pr)^{0.33} \quad (21.27)$$

Heat transfer coefficient H can be found from

$$H = \frac{K_f Nu}{L_x} \quad (21.28)$$

For natural or free convection, in which the temperature differential is the main cause of heat transfer from the surface of a solid to air, the following relationships can be used:

$$Nu = k(GrPr)^a \quad (21.29)$$

The constants for vertical flat surfaces and vertical cylinders are as follows: for $10^4 < Gr Pr < 10^9$, $k = 0.59$ and $a = 0.25$, and for $10^9 > Gr Pr$, $k = 0.021$ and $a = 0.4$; for a horizontal cylinder, $k = 0.525$ and $a = 0.25$, when $Pr > 0.5$ and $10^3 < Gr < 10^9$. Gr and

Pr are Grashof and Prandtl numbers, respectively. These numbers are defined in the Nomenclature.

The heat transfer equation for a granular product in which airflows through the mass can be estimated by the following equation. For $Re > 350$,

$$H = 1.06Re^{-0.41} \quad (21.30)$$

and for $Re \leq 350$,

$$H = 1.95Re^{-0.51} \quad (21.31)$$

The units of H in Equation 21.30 and Equation 21.31 are Btu/(h ft² °F). 1 Btu/(h ft² °F) is equal to 5.678 W/m² K.

21.5 PHYSICAL AND THERMAL PROPERTIES

Drying calculations are based on air and material properties and conditions. The air properties were discussed in Section 21.3. The material properties will be reviewed in this section. These properties are divided into two groups: thermal properties and physical properties. Physical properties are important in the selection of the right drying system for the food material.

21.5.1 SPECIFIC HEAT

Specific heat is the amount of heat energy needed to raise the temperature of a unit weight of material by one degree. The specific heat of meat is given by

$$C_p = 0.4 + 0.006M \quad (21.32)$$

where M is the percentage moisture content, dry basis. This formula has also been used for juices.

A more general equation to calculate the specific heat of food from the specific heat of its constituents is given by

$$C_p = 0.34X_c + 0.37X_p + 0.4X_f + 0.2X_a + 1.0X_m \quad (21.33)$$

where X_c , X_p , X_f , X_a , and X_m are mass fractions (decimal) of carbohydrates, protein, fat, ash, and moisture content, respectively. A simplified formula is of the form

$$C_p = 0.5X_f + 0.3X_s + 1.0X_m \quad (21.34)$$

where X_s , X_m , and X_f are mass fractions of solids, moisture, and fats, respectively.

21.5.2 THERMAL CONDUCTIVITY

Thermal conductivity is a material property indicating the ease and speed with which heat can be transferred through the food item. Thermal conductivity changes with moisture content. The following equation has the widest use:

$$K = K_l \left[\frac{1 - (1 - aK_s/K_l)^b}{1 + (a - 1)b} \right] \quad (21.35)$$

where

$$a = \frac{3K_l}{2K_l + K_s} \quad (21.36)$$

$$b = \frac{X_s}{X_s + X_l} \quad (21.37)$$

where K_l is the thermal conductivity of the liquid component of the product, K_s is the thermal conductivity of the solids, X_l is the mass fraction of the liquid, and X_s is the mass fraction of the solids.

A more specific equation, which has been applied to fruit juices and sugar solutions, is

$$K = (307 + 0.645T - 0.001T^2)[0.46 + 0.054(\%M)]10^{-3} \quad (21.38)$$

where T is temperature in °F and K is in Btu/(h·ft·°F).

Thermal conductivity of a bulk of material filled with air can be estimated by the equation

$$K = K_g v_f + K_s(1 - v_f) \quad (21.39)$$

where v_f is the void fraction, K_g is the thermal conductivity of air, and K_s is the thermal conductivity of solids.

Other physical properties required in dryer calculations are dry density dd , wet density wd , and density d , at any moisture content:

$$d = dd + (wd)M_d \quad (21.40)$$

where M_d is the dry basis moisture content.

21.5.3 PARTICLE SIZE

The size of food particles and its variation play a very important role in the design of a drying system. Particles can be defined as fine or coarse. The size of a particle influences the movement and residence time in the dryer. The size also controls the rate and amount of moisture removed or held by the material.

A powder with an average size of 50 μm holds more moisture than a larger granular product with an average size of 5 mm.

Particle size and its variations are measured by passing the material through standard sieves. The mass fraction of material kept on each sieve is plotted against the sieve size. This plot usually takes the form of a normal plot from which a mean and standard deviation for the particles can be calculated (see Standard S319.2 in Ref. [25]). Since particle size usually has a wide range, a logarithmic scale is used to express the size of particles.

Another important size factor in drying is the specific surface area that is defined as the ratio of surface area of particles in a unit volume or a unit of mass. One may estimate the surface area of a particle from πD^2 , where D is the diameter of sphere of the same volume as that of the particle. The surface area per unit volume can be estimated from Ref. [26].

$$A = \frac{6\lambda w}{\rho_p D_p}$$

where λ is a shape factor that is about 1.75 for most irregular shaped particles, w is the mass of particulate material (kg), ρ_p is the particle density (kg/m^3), and D_p is the particle size (m) determined by sieving technique.

21.5.4 DENSITY

The density of food material is expressed in two ways, solid density and bulk density. Solid density is the mass of the solid over the volume of the solid excluding the air voids:

$$\rho_s = \frac{m}{V_s}$$

Bulk density is the mass of solids over the bulk volume of the solids including inter- and intraparticle air voids:

$$\rho_b = \frac{m}{V}$$

Solid density is measured by a method by which air voids can be excluded from measurements. Air comparison pycnometry uses differential air pressure for volume measurements.

Bulk density is measured by filling a container of known volume with particles. The mass of the particles in the container is then measured. Care must be paid to a uniform packing method to fill the container.

The bulk density and solid density values are used in the following equation to estimate the bulk porosity of the granular food materials:

$$\varepsilon = 1 - \frac{\rho_b}{\rho_s}$$

Bulk density of moist food material ranges from 0.3 to 0.5.

21.6 METHODS OF DRYING

Drying methods have been evolved around every product's specific requirement. The process takes many forms and uses many different kinds of equipment. In general, drying is performed by two basic methods: (a) adiabatic processes and (b) nonadiabatic processes. In adiabatic processes, the heat of vaporization is supplied by the sensible heat of air in contact with the material to be dried. In nonadiabatic processes, the heat of evaporation is supplied by radiant heat or by heat transferred through walls in contact with the material to be dried. Dehydration may also be accomplished by mechanical dewatering. However, in this chapter, dehydration due only to adiabatic or nonadiabatic as defined above will be described.

In all drying methods, the product must be brought in contact with a medium, which is often air, in order to remove the moisture from the product surface and its surroundings.

21.7 DRYERS

21.7.1 INTRODUCTION

The diversity of food products has introduced many types of dryers to the food industry. Some of the products to be dehydrated and possible dryer types are listed in Table 21.2.

Methods of supplying heat and transporting the moisture and the drying product are the basic variations among different types of dryers. In general, the types of dryers and their operational design characteristics are reviewed.

21.7.2 SUN DRYING

Throughout the world, sun drying is a popular drying method. The sun's radiant heat evaporates moisture. The traditional drying method consists of spreading a thin layer of the product on a smooth pad. The product is stirred and turned over at intervals. Drying proceeds well in warm, dry weather. The temperature

TABLE 21.2
Food and Feed Products and the Most Suitable Dryer Types

Product	Dryer Type
Vegetables, confectionery, fruits	Compartment and tunnel
Grass, grain, vegetables, fruits, nuts, breakfast cereals	Conveyor band
Grass, grain, apple, lactose, poultry manure, peat, starch	Rotary
Coffee, milk, tea, fruit purees	Spray
Milk, starch, predigested infant foods, soups, brewery, and distillery by-products	Film drum
Cereal grains	Moving or stationary packed beds
Starch, fruit pulp, distillery waste products, crops	Pneumatic
Coffee, essences, meat extracts, fruits, vegetables	Freeze and vacuum
Vegetables	Fluidized bed
Juices	Foam mat
Apples and some vegetables	Kiln

of the product during sun drying ranges from 5 to 15°C above ambient temperature, and drying time may extend up to 3–4 weeks, as for raisins and apricots. Color, shape, and the initial and desired final moisture conditions of the product influence product temperatures and drying time.

Sun drying is widely practiced in grain drying. Figure 21.6 shows the relative absorptivity of various grains compared with the absorptivity of an ideal black surface, which is 1 on the scale in Figure 21.6.

Solar energy is also used to dry fruits and grains indirectly. In this method, a solar collector captures solar energy and raises the air temperature. The warmed air is directed to a conventional batch or continuous dryer. In a flat-plate air-type solar collector, the temperature rise is usually about 5°C over a 24-h period for airflows used in the near-ambient temperature drying of granular products. The air may be heated further by conventional fuels if high-temperature drying is desired. Overdrying and susceptibility to spoilage are major problems in solar drying of deep beds of cereal and oil seeds.

21.7.3 CABINET DRYERS

A cabinet dryer can be a small batch tray dryer. Heat from the drying medium (hot air) to the food product is transferred by convection. The convection

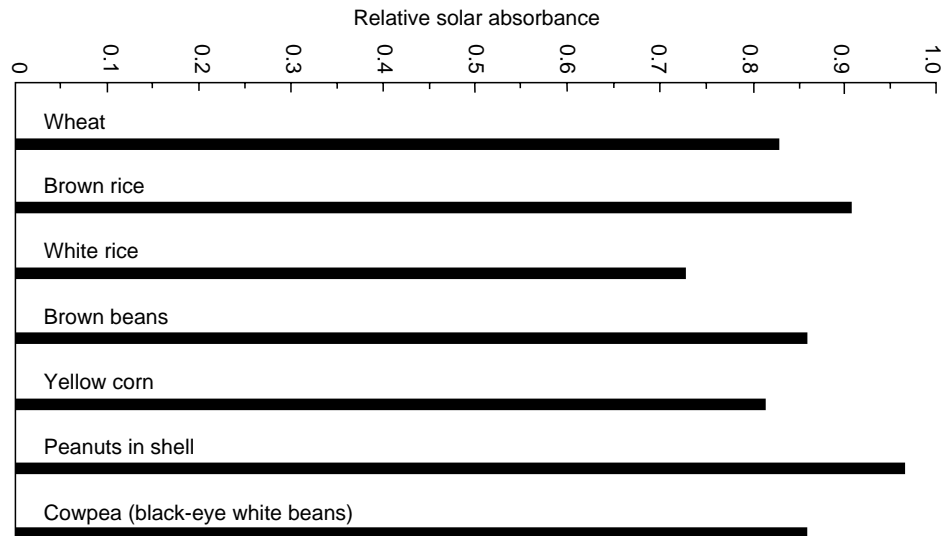


FIGURE 21.6 Relative solar absorbance of several grains. (From Arinze, E.A., Schoenau, G., and Bigsby, F.W., *Solar Energy Absorption Properties of Some Agricultural Products*, ASAE Paper No. 79-3071, American Society of Agricultural Engineers, St. Joseph, Michigan, 1979. With permission.)

current passes over the product, not through the product. It is suitable for dehydration of fruits, vegetables, and meat and its products.

The main feature of a cabinet dryer is its small size and versatility. The main problem with cabinet dryer is difficulty in even distribution of heated air over or through the drying material. The minimum airflow between the trays is maintained at 2.5 m/s [34] to ensure uniform drying. Cabinet dryers vary in capacity from a few to 20,000 kg/day of dried fruits and vegetables.

It is relatively easy to set and control the optimum drying conditions in cabinet dryers. For this reason, various heat-sensitive food materials can be dried in small batches. The heat source is usually steam batteries or steam coils. The air from a centrifugal fan is passed through the coils and then baffled across the trays loaded with the product. The trays may either be loaded onto trolleys in stacks of 10–12 or may be stacked individually into the slots of the cabinet. The movement of trays may be manual or mechanically assisted depending on the size and capacity of the dryer. The hot air in almost all cabinet dryers is introduced at the top, and provision is made to recycle the air so that its total drying potential is utilized by the time it discharges to the atmosphere. A schematic illustration of a cabinet dryer and its principle of operation are shown in Figure 21.7.

21.7.4 TUNNEL DRYERS

Tunnel dryers are of many different configurations in general having rectangular drying chambers. The

number of tunnels in a dryer is quite variable and can be as high as 100. Truckloads of the wet material are moved at intervals into one end of the tunnel.

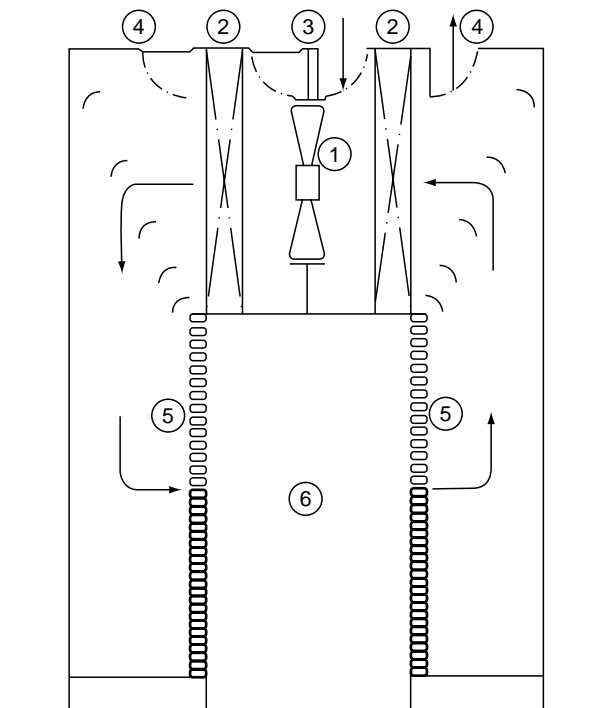


FIGURE 21.7 Schematic illustration of a cabinet dryer (1. circulating fan, fully reversible; 2. heater batteries; 3. vented air inlet ports; 4. vented air exhaust ports; 5. adjustable louver walls; 6. truck space). (From Heldman, D.R., *Food Process Engineering*, AVI Publ. Co., Inc., Westport, CT, 1975. With permission.)

The whole string of trucks is periodically advanced through the tunnel until these are removed at the other end of the tunnel. Air movement, circulation, and heating methods vary in tunnel dryers. Three different arrangements, namely, counterflow, parallel flow, and combined flow, are shown in Figure 21.8. These dryers are simple and versatile in comparison with other types of dryers. Food pieces of any shape and size can be handled. If solid trays are incorporated, fluids can also be dried.

In the three-tunnel dryer arrangement, wet loads are moved at intervals into outer wet tunnels. At the end of the wet tunnels, the trolleys are alternately loaded to the central dry tunnel. This in effect reduces the residence time in the dry tunnel to half the time spent in wet tunnels. Dry tunnels are normally operated at a lower temperature than wet tunnels to prevent long exposure of the product to high temperatures and to save energy. The range of wet tunnel temperatures is 99–104°C and of dry tunnel temperatures is 65–71°C for drying of root vegetables.

21.7.5 BELT DRYERS

The belt dryer is particularly suitable for cut vegetables. The main components of this dryer consist of finely woven wire mesh supported on rollers to form a belt trough, a duct supplying hot air at the bottom of the trough, and a rotary cleaning brush. The arrangement of these components is shown in Figure 21.9. The belt trough is inclined sideways at an angle of 15–20° to facilitate the gravity unloading of dried product. The airflow rate provides an air cushion but not so high as to fluidize the product. Soft materials can be dried without mashing or rounding their cut edges. Since the product is partly supported by the hot air, it requires relatively uniform size wet material. If material is not of uniform size, large particles may not be supported well and small particles may be blown out of the trough.

The drying air passes through material at the rate of 0.6–1.4 m/s. About 60 to 90% of air is recirculated in each section. Temperatures are usually limited by the heat sensitivity of the material but seldom exceed

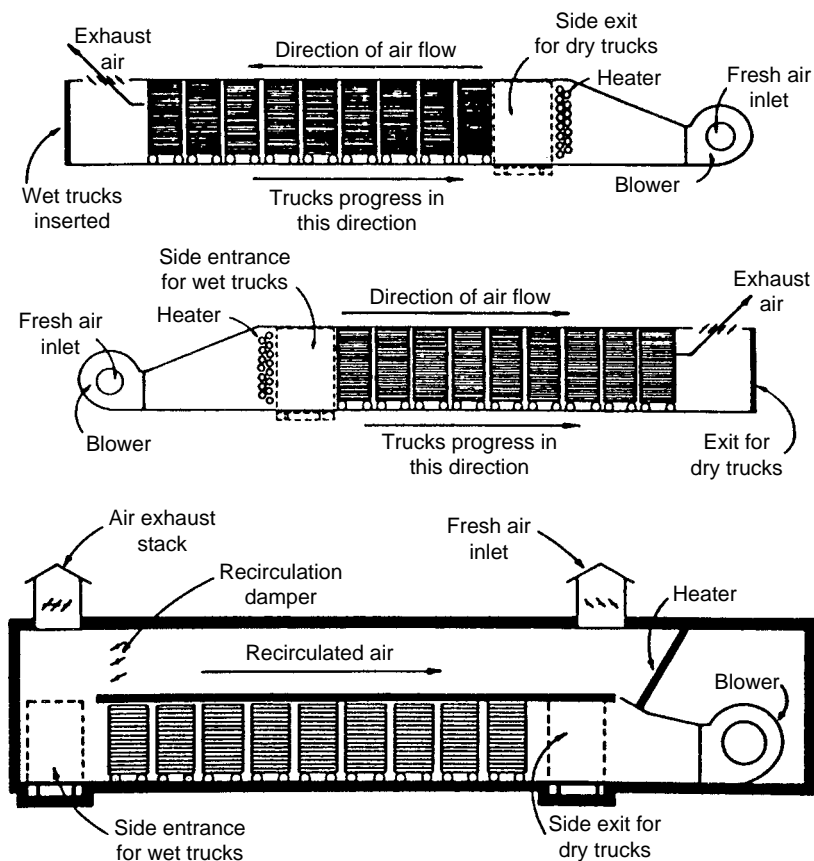


FIGURE 21.8 Schematic illustrations of tunnel dryers. (From Heldman, D.R., *Food Process Engineering*, AVI Publ. Co., Inc., Westport, CT, 1975. With permission.)

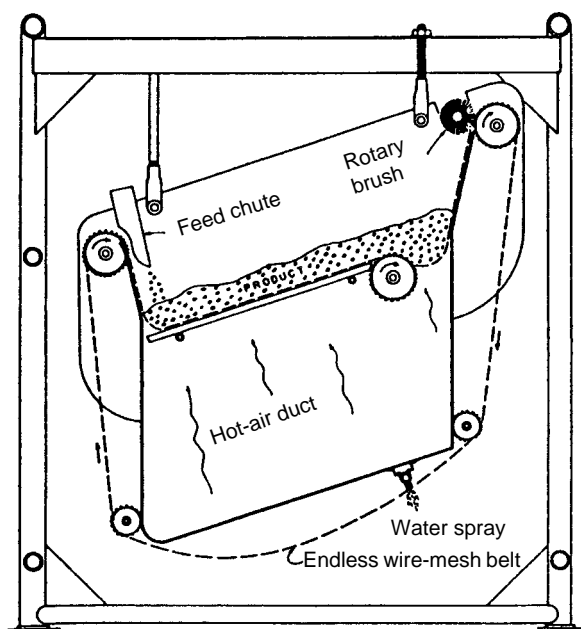


FIGURE 21.9 Schematic cross section of belt trough dryer. (From Van Arsdel, W.B. and Copley, M.J., *Food Dehydration*, Vol. 1, AVI Publ. Co., Inc., Westport, CT, 1973. With permission.)

300°C. A higher temperature may damage the lubrication of the conveyor's moving parts. The fan power is about 1 kW/m of the belt width.

The main problem in these dryers is that the sticky materials will lodge and stick to the chains and links of the band. The band must be washed frequently to prevent air blockage.

21.7.6 CONVEYOR BAND DRYERS

A conveyor band dryer consists of single or several perforated wire mesh aprons or conveyor bands as the main component. The wet material is fed evenly at the feed end and is conveyed along the length of the dryer. As in a belt trough dryer, hot air is forced through the bed of moving material. However, unlike the belt trough dryer air is not forced at high rates to support wet material. The dried product is continuously discharged at the end of the dryer. The air characteristics, such as temperature and relative humidity, may be adjusted throughout the passage to satisfy the drying characteristics of the product. The direction of movement of air through the permeable bed of the product may be either upward or downward. The thickness of the bed of the product is kept at 25–250 mm. This thin product layer and higher airflow rates result in the uniform drying of the product. Two configurations of band dryer (two stage and multideck) are shown in Figure 21.10.

21.7.7 SPRAY DRYERS

Spray dryers are used for dehydrating fluids. The fluid is introduced in the heated airstream in spray form. Dried product is separated from the airstream and is collected for further processing. The design of spray dryers ranges from very simple to very complex, depending upon the fluid. The main differences in the designs are the variations in atomizing devices, in airflow patterns, in air-heating systems, and in separating and collecting systems.

The main components of a spray dryer are a drying chamber, an atomization, and dispersion device to introduce fluid in small droplets into the drying chamber, an air-heating and blowing system, and a device to separate and collect the dried product from the air. Air heating may be done by direct firing of gas or liquid fuels in the airstream or by indirect heating through a heat exchanger. Direct firing may contaminate the product; the indirect system results in lower efficiency. Two configuration of spray dryers are shown in Figure 21.11.

Pressure spray heads, two-fluid nozzles, and centrifugal atomizers are used for dispersion of the fluid particles. Dried product is separated by different methods depending on the make of the dryer. Some dryer designs use cyclone separators; others employ settling chambers. In a cyclone separator, air is introduced tangentially at the top, swirls downward, and reverses the direction of flow to exhaust from the top. In the settling chamber, direction of airflow is changed to facilitate the separation of dried product under force of gravity.

DESIGN EQUATIONS FOR SPRAY DRYERS

21.7.7.1 Energy Requirements

Energy requirement in horsepower (hp) for three different types of the atomizers is given as follows.

1. Pressure nozzles

$$E = 7QP \times 10^{-4} \quad (21.41)$$

where Q is the liquid flow rate (gal/min) (U.S. gallon) and P is the pressure drop over the nozzle (psi).

2. Air blast atomizer

$$E = 0.136GT_a[0.5M_c^2 + 2.5(1 - P^{0.286})] \quad (21.42)$$

where G is the air mass flow rate (lb m/s), T_a is the temperature (R), M_c = Mach number, as defined in the Nomenclature.

3. Rotary atomizers

$$E = 42.5Q(DN')^2 \times 10^{-12} \quad (21.43)$$

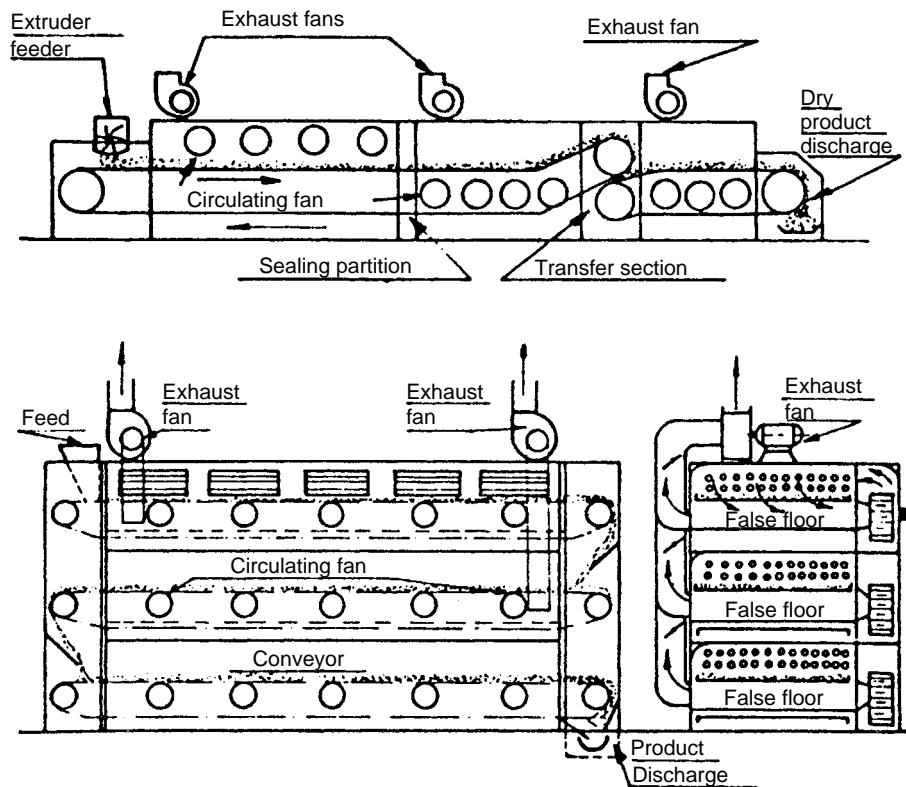


FIGURE 21.10 Schematic views of a two-stage (top) and multideck (bottom) conveyor band dryer. (From Gardner, A.W., *Industrial Drying*, Leonard Hill, London, 1971. With permission.)

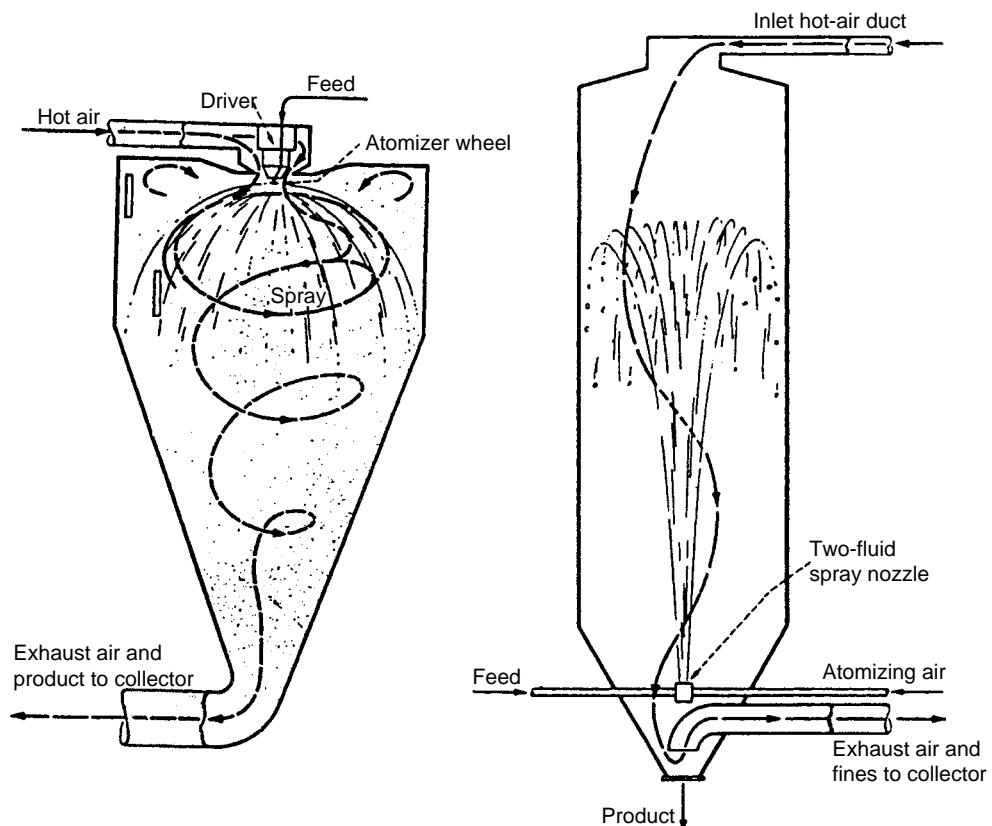


FIGURE 21.11 Two configurations of spray dryers.

where Q is the liquid flow rate (lb m/h), N is the rotation speed of the atomizer (rpm), and D is the diameter of the atomizer (ft).

21.7.7.2 Droplet Size

The droplet size obtained after the atomization process is given by

$$MD = \frac{B}{C} \frac{FN\gamma d}{\sin(T)P} (d_a)^{-1/6} \quad (21.44)$$

where MD is the mean droplet diameter (μm), B is the constant, varies from 26.3 to 43 and is found experimentally, C is the discharge coefficient for a nozzle size, ~ 0.4 , FN is the fluid number, a constant representing the flow rate divided by the square root of the pressure differential, ~ 0.8 , T is the angle of the spray, ~ 120 , P is the pressure gradient, psi, ~ 100 , d is the liquid density, lb m/ft³, ~ 60 , d_a is the air density, lb m/ft³, ~ 0.065 , and γ is the surface tension, dyn/cm, ~ 50 .

21.7.7.3 Heat and Mass Transfer Coefficients

Heat and mass transfer coefficients in spray drying can be found from the following formulas:

$$Nu = 2 + K_1 Re^{1/2} Pr^{1/3} \quad (21.45)$$

$$Sh = 2 + K_2 Re^{1/2} Sc^{1/3} \quad (21.46)$$

where Nu is the Nusselt number, Sh is the Sherwood number, Sc is the Schmidt number, and $K_1 = K_2 \sim 0.52 = 0.60$.

21.7.7.4 Drying Time

The drying time required for a droplet to vanish completely during drying is given by

$$t_c = \frac{d_l L D_0^2}{8K_f(T_a - T_w)} \quad (21.47)$$

where d_l is the liquid density, L is the heat of vaporization, D_0 is the initial droplet diameter, K_f is the thermal conductivity of air at droplet and air interface, T_a is the air temperature, and T_w is the wet bulb temperature.

The drying time for a droplet to reach from diameter D_1 to D_2 is given by

$$t_c = \frac{L}{8K_f(T_a - T_w)} (d_l D_1^2 - D_2 D_2^2) \quad (21.48)$$

where the diameter D_2 is the droplet diameter at the end of the constant drying rate period and d_1 and d_2 are densities of the product at the stages 1 and 2.

The drying time during the falling rate period is given by

$$t_f = \frac{d_p D_c L (M_c - M_e)}{6HdT} \quad (21.49)$$

where d_p is the particle density, D_c is the diameter at the critical moisture content M_c , M_c is the critical moisture content (db), M_e is the equilibrium moisture content (db), and dT is the time-averaged temperature difference between particle and air during the falling rate period.

21.7.8 FREEZE DRYERS

Upon heating of substances in the frozen state, liquefaction generally precedes vaporization. However, in a process called sublimation, the liquid state is bypassed and frozen particles are changed directly to the vapor state. Freeze dryers work on the principle of direct evaporation of the ice in the product to be dried. For direct sublimation, the temperature and the vapor partial pressure is the ice point below the triple point of water. The triple point of water is at approximately 0°C and 4.59 mm Hg absolute pressure.

The process of freeze drying is completed in two stages: (a) freezing and (b) high-vacuum drying. Both these stages are highly expensive; therefore, freeze drying is feasible only for highly valued foods such as fish, meat, chicken, and coffee. This method may also be a necessity for highly heat-sensitive food products. The drying is done at low temperatures; therefore, the heat degradation of the nutrient is minimal and the product obtained is of high quality. The dehydration of vegetables is possible with this method.

Design equations in freeze drying:

The drying time required to freeze dry the product can roughly be estimated by either of the following equations:

1. Formula based on heat transfer

$$t = \frac{LL_x^2}{4K(T_a - T_I)} \quad (21.50)$$

where L_x is the thickness of the product (ft), L is the latent heat of sublimation (Btu/lbm), K is the thermal conductivity of product (Btu/(°F·ft·h)), T_a is the air temperature (°F), and T_I is the ice temperature (°F).

2. Formula based on mass transfer

$$t = \frac{BL_x^2}{16K} \quad (21.51)$$

where B is the mass of water per unit volume of the product (lb/ft^3) and K is the thermal conductivity of the material.

21.7.9 DRUM DRYERS

In drum dryers, the surface of a pair of rotating hot drums is coated with the liquid or semiliquid form of the product. Drums are heated by steam or by direct firing inside the drum. The drums rotate slowly and over the course of about 300° of rotations the product is dried. The dried product is scraped off with doctor blades in flakes or sheets.

Drum dryers are classified on the basis of the number of drums as single-drum dryers and double-drum dryers. A typical feed arrangement for a drum dryer is given in Figure 21.12. The product is collected in the trough and is conveyed toward one end of the drum. By completely enclosing the system, these dryers may be operated under vacuum for heat-sensitive materials that may otherwise spoil at atmospheric pressure.

Foods containing high amounts of sugars may still be in the molten stage when they approach the doctor blades. High-velocity air flowing countercurrent to the drum rotation, a blast of chilled air directed on the product just preceding the doctor blades, and a supply of low-humidity air near the scraping and collecting area are special features to deal with viscous materials.

21.7.10 FOAM MAT DRYERS

Foam mat dryers can only be used to dehydrate liquid foods that are capable of forming stabilized foams. This requirement imposes a restriction on the wide applicability of the method in the food industry be-

cause only a few foods, such as whole milk, have a foam-stabilizing capability. Fruit juices have been dried successfully by adding foam stabilizers. Vegetable gum and soluble protein have successfully been used in the food industry as foam-stabilizing agents.

The liquid food is first foamed and then spread in mats on a perforated or solid support to be dried in hot airstreams. Foaming is accomplished by agitating the fluid product with the airstream. The foam structure is required to last during drying so that the dried product is easily ground to powder. The thickness of the foam layer is maintained at about 0.1–0.5 mm. The layer is dried in a few minutes at temperatures of about $65\text{--}70^\circ\text{C}$. The foamed product is spread on perforated floor craters as the airstream is forced through the bed. The cratering of foam exposes more surface to the airstream when air passes through the holes of the screen and through cratered product around the screen hole.

21.7.11 VACUUM DRYERS

In vacuum drying, the boiling point of water is lowered below 100°C by reducing the pressure. If the atmospheric pressure is reduced 100 times, then the boiling point will be around 0°C . The degree of vacuum and the temperature for drying depend on the sensitivity of the material to drying rate and temperatures.

Vacuum drying is one of the most expensive methods of drying. Its costs are comparable to freeze drying but are higher than other methods. Because they are expensive, vacuum dryers often serve as a secondary dryer. The moisture content of high-moisture food is reduced to 20–25% by a conventional method, such as hot air drying, and then vacuum is applied to bring the moisture down to 1–3%.

Because of the reduced pressures, transfer of heat depends on methods other than convection. Radiation and conduction are other modes; however, conduction may not be efficient because the drying materials shrink, thus reducing the contact area.

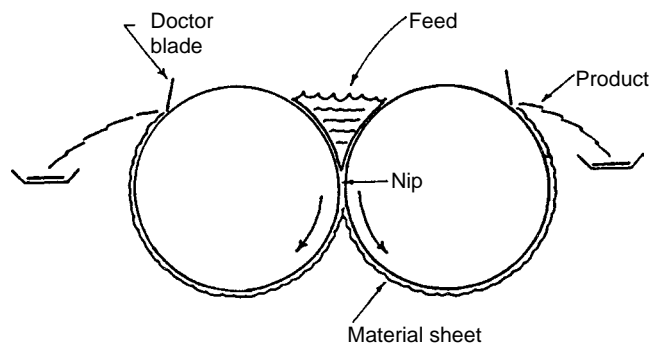


FIGURE 21.12 A typical feed arrangement of a drum dryer.

This method of drying is not very common in the food industry because of high costs. The method has been applied for dehydration of citrus juices, apple flakes, and various heat-sensitive products in which the ascorbic acid retention is important.

The product is dried under vacuum at low temperatures. Temperatures range between 35 and 60°C. Vacuum-dried products are quite hygroscopic. Special care is needed during packing of the materials to protect against absorption of moisture. This method could be applied to batch as well as to continuous systems. In batch systems, product in trays is dried in a compartment maintained at pressures reduced below atmospheric pressure.

In continuous systems, the product is spread over a stainless steel belt and passed through a vacuum vessel with the help of rollers. The dried product is collected at the other end. This method is also used in conjunction with foam mat drying, described previously.

The vacuum drying technique is also used for the extraction and concentration of essences and flavors.

21.7.12 FLUIDIZED BED DRYERS

The drying air is introduced at high velocities through the particulate material. The velocity is raised to the point that the particulate material is fully suspended in the hot airstream. All particles are completely exposed to drying air, resulting in high rates of heat transfer. The high rate of heat transfer results in instantaneous evaporation of moisture at the entry point. The dryer mainly consists of a chamber with a perforated bottom through which the air is forced at high velocities. If a glass window is made in the chamber, the start of fluidization can be observed visually. If this is not possible, then the flattening of the pressure drop curve with increasing air velocity indicates the onset of fluidization (Figure 21.13). This system is more suitable for batch drying but could also be applied to continuous drying of particulate materials.

Food materials that are suitable for fluidized bed drying have the following characteristics:

1. The average particle size should be between 10 μm and 20 mm. Very fine particles tend to lump together.
2. The particle size distribution should be reasonably narrow. A wide range of particle sizes makes the selection of a gas velocity nearly impossible. A vibrated fluid bed sometimes overcomes this problem when it is impossible to have a uniform size material.
3. The particle shape must be regular.
4. A lump of particles must disintegrate easily upon fluidization.

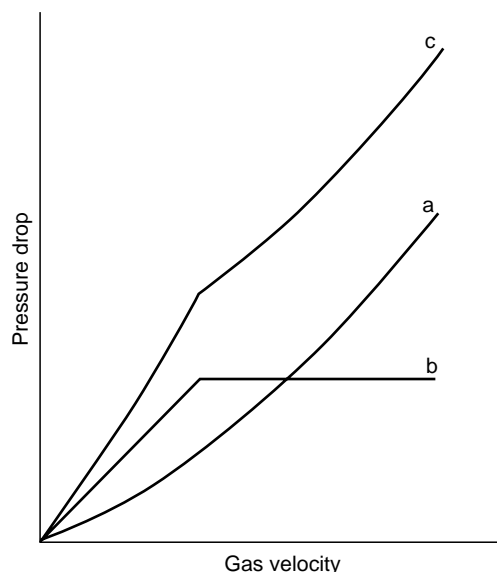


FIGURE 21.13 Pressure drop through fluidized beds: (a) pressure drop through perforated base plate; (b) pressure drop through granular bed; (c) combined pressure drop (a and b). (From Gardner, A.W., *Industrial Drying*, Leonard Hill, London, 1971. With permission.)

21.7.13 MICROWAVE DRYERS

In microwave drying, the product is exposed to very high-frequency electromagnetic waves. The transfer of these waves to the product is similar to the transfer of radiant heat. As a result of high-frequency waves, water molecules are polarized and tend to change orientation. In the process of orientation, sufficient heat to expel moisture from the product is generated.

Microwave generators designed for drying operate under two frequencies, 100 and 900–2500 MHz. As the frequency is increased, the heat generated by the system is decreased. Microwave applications have tended to be in tempering, blanching, and cooking of food. However, there are some new installations for drying purposes. The drying of pasta is perhaps the best known of these, in which microwave is used because normal drying must be very slow to avoid surface cracking of this diffusion-limited material. Since high-frequency heating evaporates water inside the material and causes convecting cooling on the surface, the outside surface remains wet until the end of the drying cycle; hence, cracking does not occur.

Another advantage of microwave is an increase in the rate of killing of bacteria due to the speedy temperature rise. Microwave drying under vacuum is becoming an established method of drying citrus fruit concentrates.

21.7.14 SUPERHEATED STEAM DRYING OF FOODS

Superheated steam drying has long been recognized as a drying method that leads to nonpolluting and safe drying at low energy consumption. The principle of the process is based on using superheated steam for drying coupled with a vapor recompression cycle to recover heat. The schematic of the system is shown in Figure 21.14. It consists of a heat treatment chamber, a compressor, a heat exchanger, and a number of blowers. The drying medium is superheated steam that runs in a closed loop picking up moisture from product in the heat treatment chamber and condensing the evaporated water in a heat exchanger. Part of the saturated steam is bled off from the main recycling line, compressed, and its heat is transferred to the recycled steam to the treatment chamber. The warm condensate can be used for heating purposes or discarded.

Superheated steam drying has been credited with the following advantages:

1. Improved drying efficiency, as much as 50% less than conventional systems
2. Pollution free, because it is a closed system no particulate or obnoxious gases are emitted to the environment
3. No product oxidation, because there is no direct contact of hot oxygen with the product
4. Phytosanitary; steam heat is superior to dry air in destroying all stages of insects, molds, and other microorganisms

5. Better degree of control of the dryer operation; by adjusting the quantity of steam bled into the compressor, the degree of dryness of the final product can be adjusted
6. Superior nutritional quality of feed; steam treatment and hydrolysis of product improve digestibility of feeds by animals

Several plants based on the vapor recompression cycle have been constructed and tested in Europe. One system uses a horizontal conveyor band in a tunnel drying setup (Bertin, France). Another manufacturer has developed a fluidized bed dryer using high pressure steam (NIRO, Denmark). The results of both systems have been encouraging.

21.7.15 CONTINUOUS PACKED BED DRYER

The mixed flow dryer is a continuous dryer shown in Figure 21.15. The dryer consists of one or two rectangular columns. A typical column is 762 mm thick. A series of inverted V-shaped ducts are staggered traversal within the column. One row of these ducts are open to the inside plenum of the dryer and the other group are open to the outside. Grain flows from top to the bottom around these ducts. In this design, grain is mixed somewhat as it flows downward.

The central plenum is divided into a heating plenum and a cooling plenum by a divider. Grain is dried in the top section of the column (about 3 m length). The grain is cooled in the bottom section (about 1.2 m

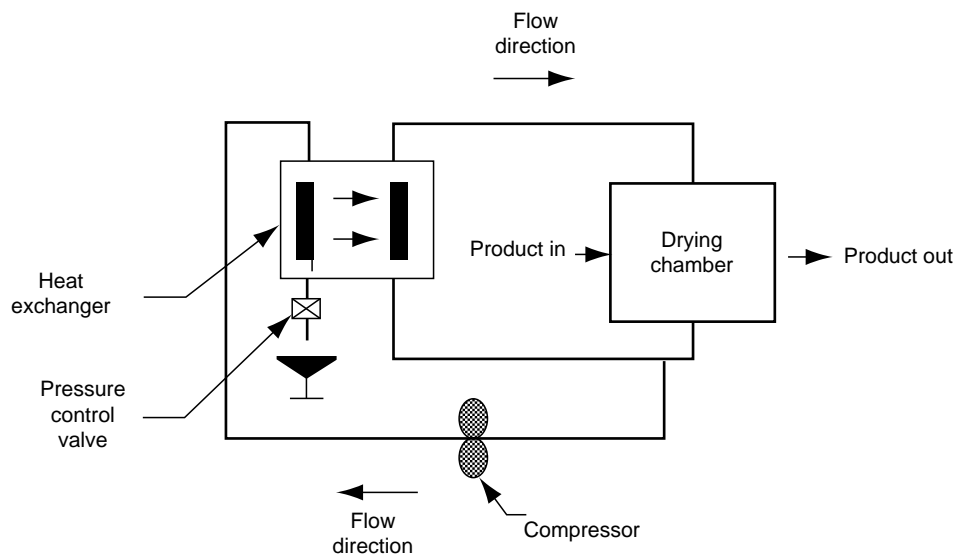


FIGURE 21.14 Schematic of superheated steam dryer.

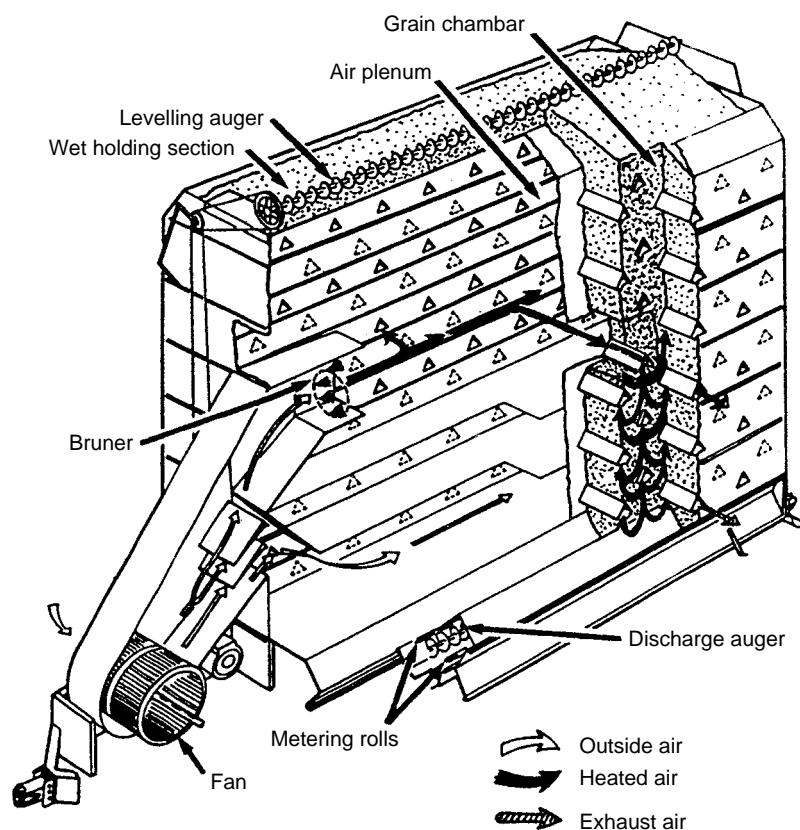


FIGURE 21.15 Mixed flow dryer. (Courtesy, PAMI, Humboldt, Saskatchewan, Canada.)

length). A single blower pushes the outside air through the grain columns. Air on the top is heated with a burner placed in the dryer plenum. Part of the air from the blower enters the lower section for cooling. A grain auger levels the grain on the top of the dryer for even grain flow. Table 21.3 lists typical conditions for a mixed flow packed bed dryer.

the product is generally above room temperature but well below sterilization temperature. The added heat and exposure time of the product at elevated temperatures affect the nutrient quality of the food products. The types of food degradation during drying are listed in Table 21.4 [27]. The influence of drying parameters on the food degradation will be discussed below.

21.8 NUTRIENT LOSSES DURING DEHYDRATION OF FOOD PRODUCTS

Many food products of nutritional importance, such as fruit juices, milk, eggs, and vegetables, are dehydrated in the food industry. During dehydration,

TABLE 21.3
Factors That Influence Food Quality during Drying

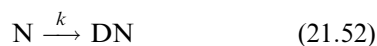
Chemical	Physical	Nutritional
Browning reaction	Rehydration	Vitamin loss
Lipid oxidation	Solubility	Protein loss
Color loss	Texture aroma loss	Microbial survival

TABLE 21.4
Operating Temperature, Energy Used to Evaporate 1 kg of Water, and Throughput Capacity (5% Point Moisture for Wheat, Barley, and Canola; 10% Point Moisture for Corn)

Grain	Drying Temp. (°C)	kJ/kg Water Removed	Throughput (t/h)
Wheat	90	4700	9.1
Barley	93	5100	8.1
Canola	72	4600	3.2
Corn	110	4400	5.3

Holding capacity of the dryer was 19.1 m³.

In terms of chemical kinetics, the loss of nutrient can be visualized as the decomposition of a particular chemical compound. This decomposition for a single monomolecular reaction may be expressed by



where N is the nutritive compound, DN is the compound formed after decomposition of N , and k is the reaction rate constant, dependent on temperature.

The dependence of k on temperature is often described by

$$k = k_0 e^{E/RT} \quad (21.53)$$

where k_0 is the constant, E is the activation energy of the reaction, R is the gas constant, and T is the absolute temperature.

Since Equation 21.52 represents the loss of nutrient N , the rate of loss of the nutrient can be described as

$$\frac{d[N]}{dt} = -k[N] \quad (21.54)$$

where $[N]$ is the concentration of compound N and $d[N]/dt$ is the rate of loss of nutrient N .

Now if it is assumed that the constant k is a true constant and the concentration of N changes only because of the reaction, then Equation 21.54 may be integrated with appropriate boundary conditions to give Equation 21.55:

$$-\ln \frac{[N]}{[N]_0} = kt \quad (21.55)$$

where $[N]_0$ is the initial concentration of nutrient N .

As the temperature of the product increases, the reaction rate constant is increased (Equation 21.53). The dependence of reaction constant on temperature implies that low-temperature dehydration would produce products with less nutrient disintegration. The vacuum and freeze dehydration processes thus would result in less nutrient degradation compared with other drying processes. A longer constant rate drying period also increases nutrient retention because, owing to evaporative cooling, product is at rather lower temperatures.

The moisture in the product can be held as bound or free water. Water acts as solvent for the chemicals of nutritional importance present in the product. As water is removed, the concentration of the chemicals increases. The loss of nutrient described in Equation 21.52 through Equation 21.55 is concentration dependent and thus would increase as dehydration progresses. On the other hand, some

of the water-soluble compounds may act as catalysts to the decomposition process. These catalytic effects are greatly reduced as the moisture is removed. The increased viscosity of the solution hinders the mobility of the catalyst.

Some nutrients, such as riboflavin, are not very soluble in water. During the dehydration process, these compounds form saturated solutions and may be precipitated. Precipitation minimizes the loss of nutrients because of the reduced concentration.

The scarcity of the kinetic data precludes predicting exactly the interaction of water and chemical degradation of the nutrients. A brief review of the qualitative aspects and nutrient retention during dehydration is presented next.

21.8.1 PROTEINS

Not much information is available to quantitatively describe the losses of proteins during the dehydration process. The effects of heat treatment on the composition and nutritive value of herring meal are presented in Table 21.5. Because of scattered conditions used in these experiments no set pattern could be established, but it could be seen that the available lysine and pepsin digestibility is substantially decreased as the moisture content of the meal during heating was increased or the heating time was prolonged. The data also indicate that the high-temperature-high-moisture contents do not affect the nutritional value of the proteins significantly. Studies based on rat net protein utilization (NPU) have shown no significant difference between freeze drying and rapid drying cod meal at 110–115°C.

21.8.2 MILK LYSINE LOSS

The loss of milk lysine in spray drying ranges from 3 to 10%, whereas in drum drying the loss ranges from 5 to 40%. During spray drying, product particles are small and are dried rapidly. Even though the air temperatures may be high, the milk particles are not heated to high temperatures. In drum drying, the milk particles at medium to low moistures are in contact with the hot drum, and therefore the loss of lysine is greater.

21.8.3 WATER-SOLUBLE VITAMINS

Unstable water-soluble vitamins, such as ascorbic acid, are sensitive to drying. The loss of ascorbic acid is dependent on the presence and type of heavy metals, such as copper and iron, light, water activity level in the product, dissolved oxygen, and the temperature of

TABLE 21.5
Effect of Heat and Moisture on Fish Meal

Conditions of Treatment			Percentage of Freeze-Dried Control			
(°C)	Moisture (%)	Time (min)	Available Lysine (%)	Pepsin Digestibility (%)	NPU	
					Rats (%)	Chicks (%)
96	7.7	30	94	88	—	98.6
	8.8	60	96	84	—	102.0
	10.8	120	87	76	—	98.1
	36.0	60	87	71	97.7	98.6
116	6.4	120	94	78.1	95.3	96.6
	7.5	60	100	78.2	97.0	98.8
	8.4	30	96.0	80.0	97.4	99.7
132	2.5	120	97	58.4	91.8	97.1

Source: From Sweeney, J.P. and Marsh, A.C., *J. Am. Diet. Assoc.* 59, 238, 1971.

drying. The losses of ascorbic acid during drying have been between 10 and 50%.

A significant difference has been found in the loss of ascorbic acid when food product is dried on metal trays versus wood trays. The difference was attributed mainly to the faster heat transfer rate and thus faster drying through the metal tray. However, the losses were less with the metal grid compared with the wood grid under similar drying conditions.

Many studies of the losses of thiamine have been reported. Depending on the moisture content of the product and the exposure temperatures and time, losses up to 89% have been reported. At 63°C, 20 h are required for a 50% loss of thiamine in dried pork. It has also been shown that for samples held at 49°C at moisture contents of 0, 2, 4, 6, and 9%, the losses of thiamine were 9, 40, 80, 90, and 89%, respectively. Thiamine losses during drying from different sources have been summarized in Table 21.6.

The loss of other water-soluble vitamins is also reported to vary widely from one study to another. Results from some of these studies are summarized in Table 21.7.

In general, it can be said that losses of water-soluble vitamins during conventional drying are less than 20%. For vegetable dehydration, the losses are less than 5% for all vitamins except ascorbic acid.

21.8.4 FAT-SOLUBLE VITAMINS

Fat-soluble vitamins are expected to degrade by the oxidation mechanism. The degradation is superimposed by direct thermal reactions during drying of the food products. This is evidenced by comparing air drying and freeze drying of carrots (Table 21.8).

For freeze-dried orange juice, losses of β -carotene of the order of 4% have been reported. Losses of other fat-soluble vitamins, such as A, D, and E, during drying have been negligible. Losses of vitamins up to 5% have been reported in oilseed drying. It could be concluded that the vitamin losses during drying are mainly related to water-soluble vitamins.

In cereal drying, not much work has been done on the quality aspects of drying. Most of the literature is based on germination loss due to heat treatment. The recommendations for safe drying temperatures are varied over a wide range among different countries.

TABLE 21.6
Thiamine Losses in Drying

Product	Conditions	Loss (%)	Ref.
Freeze-dried pork	−40°C	5	11
Freeze-dried chicken	1000 μ m Hg	5	11
Beans ^a	Air dried	5	13
Cabbage	Air dried	9	13
Corn	Air dried	5	13
Peas	Air dried	3	13
Snap beans	Air dried	5	14
Beets	Air dried	5	14
Corn	Air dried	5	14
Peas	Air dried	5	14
Rutabagas	Air dried	5	14
Carrots	Air dried	29	14
Potato	Air dried	25	14
Bean powders	Double-drum dried at 93°C for 30 s	20	15

^aThe loss reported for vegetables does not include blanching loss.

TABLE 21.7
Losses of Vitamins during Drying

Vitamin	Product	Loss (%)	Ref.
Vitamin B ₆	Freeze-dried fish	0–30	16
Pantothenic acid	Freeze-dried fish	20–30	16
Riboflavin, niacin, and pantothenic acid	Vegetables	<10	14
Riboflavin	Freeze-dried chicken	4–8	10
Pyridoxine, niacin, and folic acid	Double-drum dried at 93°C for 30 s; bean powder	20	15

In England, for seed or malting grains, it is recommended that the inlet hot air temperature should not exceed 40 and 43°C for moisture contents of less than and greater than 24% wet basis, respectively.

21.8.5 NONENZYMATIC BROWNING

A model describing the rate of nonenzymatic browning in skim milk for the range of temperature from 35 to 130°C and 3 to 5% db moisture content. The equation takes the following form:

$$\frac{dB}{dt} = k_0 \exp\left(-\frac{E}{RT}\right)$$

with

$$k_0 = \exp\left(38.53 + \frac{15.83}{m}\right), \quad \frac{E}{R} = 13157.19 + \frac{90816.51}{m^3}$$

where dB/dt is the rate of nonenzymatic browning, k_0 is the Arrhenius constant, E is the activation energy (kcal/mol), R is the gas constant, T is the absolute temperature (K), and m is the moisture content (% db).

TABLE 21.8
Losses of Carotene in Drying Carrots

Drying Process	Average loss (%)		Ref.
	Total β -Carotene	<i>trans</i> - β -Carotene	
Tray air drying	26	40	17
Freeze drying	10	20	18

NOMENCLATURE

a	constant
a	local velocity of sound, m/s
A	cross-sectional area of the dryer, m ²
A	arctangent
B	mass of water per unit volume of product, kg/m ³
C	energy constant
C	constant
C	discharge coefficient of nozzle
C_w	specific heat of water, J/(kg °C)
C_p	specific heat of dry product, J/(kg °C)
d	density of dry product, kg/m ³
d_a	density of air, kg/m ³
d_l	density of liquid, kg/m ³
dd	dry density, kg/m ³
dM	moisture removed, kg/kg
dP/dx	vapor pressure gradient, Pa/m
dt	time step, s
dT	temperature increment, °C
dz	small segment along dryer, m
D	diffusion coefficient, m ² /s
D	diameter of atomizer orifice, m
D_0	initial droplet diameter, m
DN	compound formed after decomposition of nutritive compound N
d_1	density of product at stage 1, kg/m ³
d_2	density of product at stage 2, kg/m ³
D_1	diameter of droplet at stage 1, m
D_2	diameter of droplet at stage 2, m
E	constant
E	activation energy of reaction, J
E	energy requirement, hp
F	constant
FN	fluid number
g	acceleration due to gravity, m/s ²
G	air mass flow rate, kg/(m ² ·s)
H	volumetric heat transfer coefficient, W/(m ³ °C)
H_w	absolute humidity at wet bulb temperature (saturation temperature), kg/kg
i	enthalpy, J/kg
i_y	enthalpy of water vapor, J/kg
k	drying constant, 1/s
k	reaction rate constant
k_0	drying constant, 1/s
K_m	mass transfer coefficient
K	moisture transfer (diffusion) coefficient
K	thermal conductivity of food material, J/(m h K)
K_f	thermal conductivity of air at droplet and air interface, J/(m h K)
K_g	thermal conductivity of air
K_l	thermal conductivity of liquid portion, J/(m h K)
K_s	thermal conductivity of solid portion, J/(m h K)
K_1	constant

K_2	constant
L	latent heat of vaporization, J/kg
L_x	thickness of solid, m
m	mass, kg
M	moisture content, dry basis, kg/kg
M_r	mass resistance ratio
M_c	moisture content at the center of slab-shaped, cylindrical, or spherical food particles, kg/kg
M_d	dry basis moisture content, kg/kg
MD	mean droplet diameter, μm
M_e	equilibrium moisture content
N	nutritive compound before decomposition
[N]	concentration of nutrient N, kg/m^3
N'	rotation speed of atomizer, rpm
(π)	constant, 3.14159
P	pressure drop over nozzle, Pa
P_a	atmospheric pressure, Pa
P_v	water vapor pressure, Pa
P_s	saturation vapor pressure, Pa
Q	liquid flow rate, gal/min
R	universal gas constant, 1.987
RH	relative humidity, decimal
R_c	constant for cylinder
R_p	constant for infinite plate
R_s	constant for sphere
S_c	constant for cylinder
S_p	constant for infinite plate
S_s	constant for sphere
t	time required to freeze dry the product, min
t_c	drying time during constant rate period, min
t_f	drying time during falling rate period, min
T	temperature, K
T	angle of spray, rad
T_a	air temperature, $^{\circ}\text{C}$
T_g	product temperature, $^{\circ}\text{C}$
T_I	ice temperature, $^{\circ}\text{C}$
T_s	surface temperature, $^{\circ}\text{C}$
T_w	wet bulb temperature, $^{\circ}\text{C}$
T_{inf}	temperature at free stream conditions, $^{\circ}\text{C}$
u	velocity, m/s
V	volume, m^3
V_s	volume of solid particle, m^3
v_f	void fraction
w_d	density of moist product, kg/m^3
X_a	humidity ratio in the air, kg/kg
X_a	mass fraction of ash, decimal
X_c	mass fraction of carbohydrates, decimal
X_f	mass fraction of fat, decimal
X_l	mass fraction of liquids, decimal
X_m	mass fraction of moisture, decimal
X_p	mass fraction of protein, decimal
X_s	mass fraction of solids, decimal
X_w	mass fraction of water, decimal
γ	surface tension, dyn/cm
Gr	Grashof number, $g(T_s - T_{\text{inf}}) L_x^3 / 2$

Le	Lewis number, $H/K_m C_p$
M_c	Mach number, u/a
Nu	Nusselt number, HL_x/K
Pr	Prandtl number, $C_p \mu / K$
Re	Reynolds number, $du L_x / \mu$
Sh	Sherwood number, $K_m L_x / D$
α	volume coefficient of expansion, $1/^{\circ}\text{C}$
μ	dynamic viscosity, $\text{kg/(m}\cdot\text{s)}$
ν	kinematic viscosity, m^2/s
ρ_s	particle (solid) density, kg/m^3
ρ_b	bulk density, kg/m^3
Prime (') means one step further into the dryer.	

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22 Drying of Fish and Seafood

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22.1 INTRODUCTION

Raw foods generally originate from two major sources: plant and animal kingdom. Fish and seafood are the edible flesh for a number of species of animal source. The preservation of foods by drying is the time honored and most common method used by humankind — one of the most important methods for the food-processing industry. The Mesopotamians made salted dried fish as early as 3500 BC [38]. The sundrying of fish and meat was practiced as long ago as 2000 BC and dried vegetables have been sold for about a century and dried soups for much longer [22]. Tannahill [86] noted that dry fish became particularly important when the Roman church banned the eating of meat on Fridays and during Lent.

Drying in earlier times was done in the sun, now many types of sophisticated equipment and methods

are used to dehydrate foods. During the past few decades, considerable efforts have been made to understand some of the chemical and biochemical changes that occur during dehydration and to develop methods for preventing undesirable quality losses [57]. Foods can be divided into three broad groups based on the value added through processing by drying. In the case of cereals, legumes, and root crops, very little value is added per ton processed. More value per unit mass is added to foods such as vegetables, fruits, and fish, and considerably more to high-value crops such as spices, herbs, medicinal plants, nuts, bioactive materials, and enzymes [3].

Fish muscular tissue consists mainly of muscle fibers or cells (86–88%, v/v) and some extracellular space (interstitial space, 9–12% and capillary space, 2–3%). The muscle cells consist of mainly fibrils (working units of cell, 65%), sarcoplasm (transport

and regulatory space filled with liquid and functional units, 20–23%), and finally connective tissue (6%). The muscle cells or fibers, each has a diameter 0.1 to 0.2 mm [89].

22.2 DRYING PRETREATMENT

22.2.1 SALTING OR CURING

22.2.1.1 Salting Preservation

Curing was originally developed to preserve certain foods by the addition of sodium chloride. In food industry, the application of cured is related only to certain meat, fish, and cheese products. Today sodium chloride, sodium and potassium nitrite or nitrate are considered curing salts. Salting is one of the most common pretreatments used for the fish products. Salting converts fresh fish into shelf-stable products by reducing the moisture content, and acting as a preservative. In combination with drying, these processes contribute to the development of characteristic sensory qualities in the products, which influence their utilization as food [29,79]. Although curing was originally a mechanism for preservation by salting, over several millennia additional processes concomitant with curing have evolved, notably fermentation, smoking, drying, and heating. Curing may have different connotations: in meat salt and nitrite or nitrate are always added; in fish salt is always added, but nitrite only rarely; and in cheese, which always contains salt, but infrequently contains nitrate, and the term curing is applied to the production of desirable proteolytic and lipolytic changes. In the past half-century, cured products have been developed that are not stable unless refrigerated. Indeed, most cured meat products must be refrigerated to remain safe and wholesome, and during the past two decades even the packaging of many classes of cured products has become important in extending the period during which the product remains wholesome [81]. Cured meats can be divided broadly into three groups: unheated, mildly heated (pasteurized to center temperature of 65–75°C), and severely heated (shelf stable after heating to 100–120°C) [81].

In addition to the curing salts and related processes mentioned above, additives collectively known as adjuncts are used in many cured meat products. These include ascorbates, phosphates, glucono-D-lactone, and sugars. Adjuncts are used primarily to obtain or maintain desirable changes, the ascorbates in connection with color and the others in connection with pH, texture, and in some cases flavor. Adjuncts may also affect safety. The concentration of each curing agent depends on the nature of

the food products and on the technology used in individual countries [81].

Salting can be done by placing fish in salt solution or by covering with dry salt. During salting, water is removed from the flesh, salt enters the tissues of the fish, and the body juices become a concentrated salt solution. When enough salt enters, it interacts with all proteins causing coagulation. When the tissue cells shrink because of the loss of a large share of the moisture content, the fish flesh loses most of its translucent appearance and does not feel sticky to the touch. At this stage, the salter would say it is struck through [34].

22.2.1.2 Salting Process

Fish are salted over the temperature range of 0–38°C. Higher the temperature, faster the salt infusion, and quicker the process reaches at equilibrium. The osmotic dehydration process (i.e., salting) can be characterized by equilibrium and dynamic periods [64]. In the dynamic period, the mass transfer rates are increased or decreased until equilibrium is reached. Equilibrium is the end point of osmotic process, i.e., the net rate of mass transport is zero. In general, fish absorbs salt faster as the brining temperature increases. It is best to standardize brining at a cool temperature (1.1–1.7°C) to achieve consistent and predictable results and to discourage bacterial growth. Using ice in the brine makeup water is a good way to accomplish this, but caution must be taken to make sure that no ice remains in the finished brine. Brining in a cold room is also a good way to keep brines cool and is advisable for long brining times [30].

In general, salt absorption is affected by brine concentration and temperature, brining time, thickness and geometry of fish, texture and fat content of fish, species, and fish quality [30]. Fish flesh absorbs salt faster from higher salt brine concentration. Brine greater than 15.8% tended to remove moisture from the fish, which can be advantageous in some products. However, strong brines and short times may not allow even distribution of salt into the center of the fish geometry before smoking. Dry salting has the advantage of removing moisture, but has the disadvantage of uneven salt absorption. Dry salting is a technique, which covers fish with a thin layer of salt (0.64–1.27 cm) between layers [30]. *Tilapia* was processed by dry salting (ratio fish/salt, 3:1) varying salting time (0–24 h), air-drying time (6–20 h), and drying temperature (40–60°C). The critical salting times for attaining minimum moisture were 20.5, 12, and 8.5 h, respectively, for products air dried at 40, 50, and 60°C. The hardness, color, and overall acceptability

of salted dried *Tilapia* were found to be dependent on the process variables, salting time, drying time, and temperature [44]. Diffusivity of manganese ion in cured pork was varied from 0.42×10^{-10} to $1.0 \times 10^{-10} \text{ m}^2/\text{s}$ [24]. Salt diffusion in pork meat was found in the range of 3.6×10^{-10} to $1.2 \times 10^{-10} \text{ m}^2/\text{s}$ (temperature -2 to 36°C) and for fat it was $0.07 \times 10^{-10} \text{ m}^2/\text{s}$ [94]. Salt has a profound effect on the ultrastructure and hence moisture binding of fish muscle. It has more effect compared to freezing, drying, or heat treatment [89].

Soft-textured fish tend to absorb salt faster than tough or firm-textured fish. Frozen flesh absorbs negligible salt, thus need thawing. Mishandled fish with gaping (separated flesh fibers) may have decreased brining times. High fat content fish absorb salt slower than low-fat fish. However, they may need less salt to obtain adequate final water phase salt content. Fat content in flesh varies at different locations on the body of the fish. Salmon, for example, tend to have less fat at the tail. Different species of fish have different flesh characteristics and may absorb salt at different rates. Salting times should be specific for each species. Moreover, geometrical shapes of fish having different thicknesses and widths along the length also pose difficulties in controlling the salting process and causes nonuniform salt distribution. Frozen-thawed fish or low-quality fish have flesh characteristics, which may affect (usually increase) the rate of salt absorption. The rate of freezing affects flesh cell structure and therefore the subsequent rate of salt absorption [30]. In some cases, for example, in case of salmon, the fish is soaked overnight in fresh water or for a period of 12–16 h before curing. The water is changed two or three times. About 10 or 12 h of freshening should be sufficient but a more thorough soaking may be required to satisfy some markets.

Salting or solute addition process affects the air-drying process by reducing water diffusion rate [65]. The concentration of salt has also great influence on the rate of surface evaporation [40]. In addition, depending on the salt concentration and relative humidity, the salted fish may reabsorb moisture from the environment during storage [44].

22.2.2 COOKING

Cooking before drying has been recommended for the dehydration of fish. The bacterial load on the final product can thus be much reduced, and cooked fish can be minced and spread evenly on drying trays with much less trouble than raw fish. However, the formation of a superficial pellicle (case-hardening), which may considerably retard drying, is avoided by precooking. It is clear that more severe the initial

conditions of cooking, the more stable is the subsequently dehydrated product [5].

When an animal or plant is killed, its cells become more permeable to moisture as pointed out by Potter [54]. When the tissue is blanched or cooked, the cells may become still more permeable to moisture. Generally, cooked vegetable, meat, or fish dried more easily than their fresh counterparts, provided cooking does not cause excessive shrinkage or toughening [54]. Cooking also results in a decrease in water-holding capacity of meat products [83].

22.3 SMOKING

22.3.1 SMOKING PRESERVATION

Smoking of foods is one of the most ancient, and in some communities one of the most important food-preserving processes. The use of wood smoke to preserve foods is nearly as old as open air-drying. Although not primarily used to reduce the moisture content of food, the heat associated with the generation of smoke also causes an effect of drying. Smoking has been mainly used for meat and fish. The main purposes of smoking are: it imparts desirable flavors and colors to the foods; and some of the compounds formed during smoking have preservative effect (bactericidal and antioxidant) due to presence of a number of compounds [19,57].

Smoking is a slow process and it is not easy to control the process. Smoke contains phenolic compounds, acids, and carbonyls and smoke flavor is primarily due to the volatile phenolic compound [10,20,34]. Wood smoke is extremely complex and more than 400 volatiles have been identified [43]. Guillen and Manzanos [26] identified around 140 compounds in liquid smoke prepared from *Thymus vulgaris* wood. Polycyclic aromatic hydrocarbons are ubiquitous in the environment as pyrolysis products of organic matter. Their concentrations in smoked food can reach levels hazardous for human health, especially when the smoking procedure is carried out under uncontrolled conditions [46]. Wood smoke contains nitrogen oxides, polycyclic aromatic hydrocarbons, phenolic compounds, furans, carbonylic compounds, aliphatic carboxylic acids, tar compounds, carbohydrates, pyrocatechol, pyrogallols, organic acids, bases, and also carcinogenic compounds like 3:4 benzpyrene. Nitrogen oxides are responsible for the characteristic color of smoked food whereas polycyclic aromatic hydrocarbon components and phenolic compounds contribute to its unique taste. All the three are the most controversial chemicals from a health perspective [43].

Commercial processors have therefore adjusted processing conditions to produce the lower salt and moisture products that will sell in today's markets. One result of these changes in processing practices is that processing conditions must be standardized, controlled, monitored, and documented so that the potential for producing toxic, or even lethal, food products is eliminated. This is especially true for seafood products which may contain food-poisoning organisms of marine origin that are more difficult to control than those from land sources. *Clostridium botulinum* type E is the most notorious of these marine organisms and most smoked seafood products are designed to eliminate the potential of toxin production from this bacteria species [30].

Color development in smoked fish is a complex process. Maillard type with glycolic aldehyde and methylglyoxal in the dispense phase of smoke is the dominant role [87]. Hot-smoked fish produced by exposing the fish to sawdust smoke gave good flavor and poor color [1]. Several types of synthetic colors have been used to color kippers in England [75]. Paprika has also been used as a seasoning and to impart color to smoked fish [1]. Abu-Bakar and Abdullah [1] used caramel to improve the acceptable color of hot-smoked Spanish mackerel (*Scomberomorus* spp.), chub mackerel (*Rastrelliger kanagurta*), kurau (*Polynemus* spp.), skinless squid (*Loligo* spp.) mantle. Spanish mackerel, chub mackerel, and squid immersed in brine containing 0.4, 2.0, and 0.6% caramel (w/v), respectively, for 30 min at 25°C produced most acceptable color. Smoked products with golden yellow to light brown were preferred by the panelists.

The use of wood smoke in preventing lipid oxidation in meat and fish products has been investigated [6,39]. Polyphenols derived from the smoke acted as antioxidants. Woolfe [95] found that smoke drying initiated lipid oxidation in herring *Sardinella aurita* as evidenced from peroxide values. The site of initiation was bounded by lipids in contact with the proteins and final moisture content was the predominating factor affecting the rate of oxidation. Sheehan et al. [80] found that level of fats in raw salmon affects texture, oiliness, and color of smoked salmon during storage. Cold smoking, hot smoking, or combination of both did not significantly affect the saltiness, smoke flavor and color, but significant differences were observed on texture and appearance. These were sufficient to give overall acceptance of the product [20].

Fat has an important influence on the nutritional quality of the product, as well as on the eating quality, assessed in terms of texture, flavor, and taste. It was also claimed that a high degree of fat in the connective tissue, between the myomers, can interfere

with the perceived color [18]. In case of smoked salmon, neither the fat content, which varied from 140 to 210 g/kg, nor the estimated fat deposits (7–12%) affected significantly the sensory properties (color, consistency, odor, and taste) of smoked fillets [72]. Cold-smoked fish is lightly preserved fish product, which undergoes a mild salt cure and cold smoking at temperatures below 28°C. It is sold as sliced, vacuum-packed, ready-to-eat product stored at 3–8°C. Freshly packaged cold-smoked product is not sterile and ultimately quality at room temperature storage spoils mainly due to microbiological activity [28].

All smoked fish must be stored chilled or vacuum-packed to prolong shelf life. Hansen and Huss [28] identified the microflora on spoiled, sliced, and vacuum-packed cold-smoked salmon from three different sources. Lactic acid bacteria dominated the microflora, in some cases large number of *Enterobacteriaceae* were also present. The microflora on cold-smoked salmon appeared to be related to the source of contamination, i.e., the raw material and the smokehouse rather than specific for the product.

22.3.2 SMOKING PROCESS

The traditional methods of smoked food preservation typically produced high salt and low moisture content products that are not desirable to most modern consumers. The traditional method of smoking fish uses hot smoke, from a range of woods, passed over the fish to partially dry it and impart the flavor and aroma of the smoke. Disadvantages of this method include a lack of control over the process and the finished product with consequent health concerns if the surface of the fish is not properly dried. Smoking process involves extensive handling of raw and finished products. Fillets are manually turned in the smokehouse to expose cut surfaces and skin for even smoking and drying exposure.

Smoked food is prepared with two basic procedures. One cooks the product (hot smoking) and the other does not (cold smoking). Cold smoking devices have one basic function of applying smoke to the product. Hot smoking devices have the added function of applying heat. Since preservation of fish usually requires moisture removal, systems designed for hot or cold smoking fish have the added function of dehydration. Air movement in a smokehouse is essential to the application of smoke and heat, and removal of water from the product. Traditional smokehouse used natural (gravity) convection to circulate air, whereas modern equipment uses forced (mechanically produced) convection [30]. The hot smoke process for smoking fish differs from the cold smoke process in a fundamental way. The cold smoke process requires

that the fish reaches an internal cooking temperature below 35°C, whereas the hot smoke process cooks the fish to the center at 62.8°C for at least 30 min, and hence there should be at least 3.5% water phase salt in fish for both processes. Between those two extremes are temperatures that can create an environment favorable to growth of food-poisoning bacteria. As an additional safety margin, hot-smoked fish should always be cooled to less than 3.3°C immediately after smoking and held at that temperature until consumed to prevent growth of food-poisoning bacteria. Both hot- and cold-smoked fish are preserved primarily by control of salt and moisture content (water phase salt). Smoke deposition is effective only in controlling surface spoilage [30]. Cold smoking salmon processors are required to maintain temperatures below 32.2°C for a maximum of 20 h.

The conditions in cold smoking do not completely eliminate normal harmless food-spoilage bacteria. Cold-smoked fish is not a fully preserved product and, for the same safety reasons as with hot-smoked fish, must be chilled to 3.3°C and held there until consumed. Yellowfin tuna of 4×4 cm block took 45 min (in 16% brine) to absorb enough salt to reach final water phase salt when the final product reached 60% moisture. For kingfish, the brining time was over 100 min. Large salmon take about the same time and small salmon take much less time, perhaps only 15 to 20 min [30]. Cold smoking at 37.8°C would enable microbiological proliferation if salting is insufficient and smoke deposition and dehydration rate is slow. Cold smoking at low humidity and rapid airflow retarded microflora by showing a slight decrease in surface counts. Hot smoking at 71.1°C, on the other hand, caused a very large reduction in count of vegetative microorganisms [20]. Counts of aerobic microorganism on the surface layer of samples stored at -12.2°C were very low (<10 per gram) and there were no significant increase in count with prolonged storage. Samples stored at 3.3°C showed increasing numbers with prolonged storage with a very steep exponential rise after 40 d of storage. Mold growth was apparent after 45 d at 3.3°C on some samples [20]. Although the heat treatment in 71.1°C process is lesser than what is required to inactivate bacterial spores, the product is of excellent quality and had reasonable storage stability under refrigeration [20].

The product quality of smoked fish depends on how fast it can be dried, cooked, and smoked by deposition of smoke on the product. A smokehouse is simply a drying oven with the ability to apply smoke. The smoke density, surface moisture, air humidity and temperature, and air circulation affect the smoke deposition [30]. The accumulation of surface moisture forms uneven smoke deposition. A relative

humidity of 60% at a temperature of 71.1°C produced maximum smoke deposition in some species [13]. The temperature and humidity need to be controlled at various stages of smoking cycle. While smoke density can be increased by reducing air rejection from the system (closing dampers), the same action raises relative humidity, thus reduced the drying rate. It is useful to be able to generate high smoke density even at high rejection rates. Modern automatic hot plate auger smoke generators are capable of producing large quantities of smoke if properly operated. Species of wood affect smoke deposition and flavor. Most producers have their own preference based on their markets [30].

The hot smoking of fish requires five steps, each with different goals and operating conditions. These steps are surface drying, smoking, drying, heating or cooking, and cooling. Surface drying is the removal of surface moisture leaving a protein coating (pellicle) on each piece of fish so that it accepts an even smoke deposit. Producing a dense smoke atmosphere and conditions where smoke is deposited evenly on the surface of each piece can insure good flavor, color, and surface preservation. Often color does not develop until after the surface of the fish reaches 54.4 to 60°C during the cooking step. Evenly drying the fish to reduce moisture, raise the water phase salt, and establish final texture are critical steps in producing safe products. Heating each piece of fish to at least 62.8°C and holding that temperature for at least 30 min are also critical steps in safe smoked fish. Cooling the fish below cooking temperature (48.9–60°C) in the smokehouse as quickly as possible is needed. Further cooling to less than 3.3°C to reduce growth of food-poisoning bacteria is recommended. A suitable sanitary refrigerated room is usually more practical and cost effective than a refrigerated smokehouse. Cold smoke procedures do not use step 4 of heating or cooking. Usually these five cycles take 8–12 h period. Cycles of 4 h or less are possible with thin and lightly smoked products [30]. The differences in process employed depend primarily upon the type of fish and regional preferences for a particular product. Different schedules for different fish species are specified [20]. Smokehouse is equipped with a smoke generator where smoke is passed over water to remove tar and solid particles. Good Manufacturing Practice (GMP) from U.S. Food and Drug Administration (FDA) sets minimum standards for time and temperature smoking cycles, salt and moisture content, manufacturing, holding and shipping temperatures, process monitoring and record keeping, and packaging.

More modern methods of smoking fish use formulations of liquid smoke to provide flavor and

a range of methods of drying to reduce water activity on the surface. In these methods, the fish are dipped in smoke solutions before drying. Most drying methods use heat to change the relative humidity of the air passing over the fish. This is an inefficient use of energy and in addition the heat drives off many of the aromatic chemicals that go to make up aroma, flavor, and color of the product. This could be overcome by using energy-efficient heat pump drier, where drying is performed in a closed chamber.

Smoke solution are available either as condensed products from the dry distillation of wood or synthetically prepared mixtures of phenols. The use of smoke condensates offers some advantages. They are easily applied and their concentration can be controlled. They can be analyzed, purified if necessary, and the antimicrobial activity can be evaluated. Sunen [84] identified the minimum inhibitory concentration of smoke wood extracts against spoilage and pathogenic microorganisms associated with food. He found that the effectiveness in inhibition varied with the type of commercial liquid smoke. Synthetic smokes are more nearer to actual smoke curing and harmful components can be eliminated from synthetic smokes [11]. The odor, composition of flavor compounds, and antimicrobial activity of the smoke are recognized to be highly dependent on the nature of wood. Some studies have recognized beech and oak woods as those which produce wood smoke with the best sensory properties [25]. Further herbs, spices (bay leaves, black peppers, cloves, coriander seed, and spice), or pinecones may also be added to produce unique aromatic smoke flavors [33,34]. Bacteriocin treatment was found effective inhibiting *Listeria monocytogenes* on salmon packaged under vacuum or modified atmosphere [85].

22.4 DRYING CONDITIONS

22.4.1 PROCESSING

Drying processes can be broadly classified based on the water removing method applied, as (a) thermal drying, (b) osmotic dehydration, and (c) mechanical dewatering. In thermal drying, a gaseous or void medium is used to remove water from the material, thus thermal drying can be divided into three types: (a) air-drying, (b) low air environment drying, and (c) modified atmosphere drying [57]. In osmotic dehydration, solvent or solution is applied to remove water, whereas in mechanical dewatering, physical force is used to remove water. Consideration should be given to many factors before selecting a drying process. These factors are (a) the type of product to be dried, (b) properties of the finished product desired, (c) the

product's susceptibility to heat, (d) pretreatments required, (e) capital and processing cost, and (f) environmental factors. There is no one best technique for all products [15,57].

Drying reduces the water activity, thus preserving foods by avoiding microbial growth and chemical reactions causing deterioration. The heating effects on microorganisms and enzymes activity are also important in the drying of foods. Dehydration preserves fish by destroying enzymes and removing the moisture necessary for bacterial and mold growth. The deterioration or spoilage of fish flesh is particularly due to bacteria. Fatty fish cannot be dehydrated by ordinary dehydration process, and is not possible to store it in the usual way. Fish oils or fats are drying oils, which rapidly absorb oxygen from the air and harden just as paints harden on exposure to air. Fatty fish must be dehydrated quickly in a vacuum, and must be stored in vacuum or in an atmosphere of an inert gas [34].

Earlier only sundrying was used for fish. Whereas climate is not particularly suitable for air-drying or better quality is desired, mechanical air-drying is mainly used. Nowadays, solar and mechanical air-drying is widely used commercially. Fish and seafood can be dried by using convection, vacuum, and freeze-drying methods. Convection air-drying is widely used due to its low cost of equipment and operation compared to vacuum and freeze-drying system. In general, dehydration in vacuum and freeze-drying gave the best results, but this method was considered too expensive.

Factors that affect the rate of drying are temperature, humidity, air velocity and distribution pattern, air exchange, flesh characteristics, and flesh thickness. Removing moisture from fish flesh is a process of surface evaporation and therefore requires heat. In general hotter the air temperature, the faster is the moisture evaporation. Heating the surface too fast can produce a hard crust (mostly dried soluble protein), which retards movement of moisture. This phenomenon (case-hardening) can severely reduce the rate of drying and must be avoided. Dry air picks up moisture from the surface of flesh faster than humid air. The relative humidity (a measure of dryness) is lowered when air temperature is raised. Drier must expel air to get rid of moisture, thereby allowing new, lower humidity air to enter the system. The rate that air is exhausted from a drier affects the entrance of new air and therefore affects the relative humidity and rate of drying. This is the primary way the moisture gets out of the drier after it has evaporated from the fish. The rate of surface evaporation from fish is proportional to the velocity of air passing over it. In general, the higher the velocity, the higher is the rate

of evaporation. Increased air velocity also increases the heating rate of the fish with further increasing evaporation [30]. In case of haddock and herring, the higher the temperature of the air during drying, the more stable is the product [5]. During drying in air at 80 to 90°C, deterioration of the protein or nonfatty part produces substances having antioxygenic properties. Thus, drying at 80 to 90°C gave more stability to the fat in fish.

The drying temperatures as high as 96°C can be used in the initial drying stage without harmful effect. As the product becomes drier, it is necessary to use a lower temperature in order to prevent scorching and in the later stages temperatures above 63°C are inadvisable. Relative humidity between 10 and 40% has no noticeable effect on the quality of the product. Low humidity and high initial drying temperature are helpful in increasing the rate of drying [34]. The results of British Food Investigation Board advised a maximum temperature of 70°C should be used [34].

The fish are usually placed on mesh trays as one layer and hanged from a string for better air circulation over the fish. The air circulation can be horizontal or vertical to the fish layers. Factors such as texture of meat, fat content, and species differences affect migration of moisture from the center to the outside of the piece that is dried, therefore affect the drying rate. In general, firm and high oil content flesh dries slower than soft and low-fat flesh. However, high oil content flesh has less moisture to begin with and may require less drying [30].

The recent applications on energy-efficient heat pump drying, the modified atmosphere drying, could be used for better quality and process efficiency. The use of heat pump dryer offers several advantages over convectional hot air dryers for the drying of food products, including higher energy efficiency, better product quality, the ability to operate independently of outside ambient weather conditions, and zero environmental impact [57]. In addition, the condensate can be recovered and disposed of in an appropriate manner, and there is also the potential to recover valuable volatiles from the condensate [52].

22.4.2 PACKAGING

Dry products are characterized by long shelf life, which is mostly due to the low water activity of the products. Thus, fungal and bacterial growth is seldom a problem under normal storage conditions [53]. The requirement of packaging depends on the types of dried products. The low water content dictates that the products should be kept under dry conditions. A good moisture barrier is the key to successful packaging of dry products. The effects of water activity on

the product stability are reviewed by Rahman and Labuza [67]. Furthermore, increased water activity in the package or ingress of oxygen may accelerate oxidative deterioration. Oxidation in storage may cause serious problem in dried fish products. Light was also found to cause deterioration in the stored product. Usually plastic bags are usually used for dried fish. The best results can be obtained with non-fatty fish packed in hermetically sealed containers and stored in a cool place. In many cases, sacs of desiccant or oxygen absorbers are used inside the bags containing dried fish. Antioxidant treatment packaging materials can also increase the shelf life. Low-grade dried fish also stored in open atmospheric storage conditions. Other changes taking place in dehydrated lean fish include development of a tough texture, darkening of color, and a burnt flavor and odor [34].

22.5 QUALITY CHANGES IN FISH DURING DRYING

Initial freshness plays an important part in determining the stability of dehydrated fish; the fresher the raw material, the more stability is the dehydrated product. The quality characteristics of dried foods can be grouped as microbial, chemical, physical, and nutritional (Table 22.1).

22.5.1 MICROFLORA IN DRIED FISH

Fishes are prone to rapid microbial spoilage, thus adequate care must be taken in drying the fish. Microbial standards are usually based on the total number of indicator organisms or number of pathogens [70]. The microbial load and its changes during drying and storage are important information for establishing a standard that will ensure food safety. Poor processing, handling, and storage practices often result in a limited storage life of the dried salted fish

TABLE 22.1
Quality Characteristics of Dried Foods

Microbial	Chemical	Physical	Nutritional
Pathogens	Browning	Rehydration	Vitamin loss
Spoiling	Oxidation	Solubility	Protein loss
Toxin	Color loss	Texture	Functionality loss
	Aroma development	Aroma loss	Fatty acid loss
		Porosity	
		Shrinkage	
		Pores' characteristics	

[91]. In case of foods to be preserved by drying, it is important to maximize microorganism inactivation for preventing spoilage and enhanced safety. On the other hand, in the case of drying bacterial cultures, minimum inactivation of microorganism is desired. Thus, types of detrimental effects of drying may be desirable or undesirable, depending on the purpose of drying process [57].

Rillo et al. [70] studied the microbiological quality of commercially available dried mackerel in Philippines. Their analysis included total plate count, yeast and mold counts, and tested for pathogens like coliform, *Salmonella*, *Streptococcus*, *Staphylococcus*, *Vibrio*, and *Clostridium*. The microbial load for dried mackerel ranged from 3×10^3 colonies per gram sample to too numerous to count. No evidence of spoilage was detected when the samples having water activity from 0.72 to 0.74. The isolates found were *Alcaligenes*, *Bacillus*, *Leuconostoc*, *Micrococcus*, *Halobacterium*, *Flavobacterium*, *Halococcus*, *Aspergillus*, and *Penicillium*. All the samples were positive for coliform, *Streptococcus*, and *Staphylococcus*. *Vibrio* and *Clostridium* were not detected while *Salmonella* was detected only in some samples. Brining and drying decreased the microbial load but did not eliminate the pathogens. Wheeler et al. [91] studied the common fungi involved in spoiling of dried salted fish in Indonesia. They studied the mycoflora of dried salted fish with emphasis on visible spoiled fish and spoilage fungi. A total of 364 isolates from 74 fish was cultured and identified. Wheeler and Hocking [92] studied the effect of water activity and storage temperature on the growth of fungi associated with dried salted fish. Waliuzzaman et al. [90] studied the microbial growth in trevally (*Caranx georgians*) during heat pump dehumidifier drying at low temperatures. The temperature and relative humidity were varied from 20 to 40°C and 0.20 to 0.60, respectively. It was found that microorganisms did grow during drying of highly perishable products such as fish. Lower temperatures gave lower count regardless of relative humidity of drying. Sulfur-producing organisms were a significant portion of the total flora of fish drying. Rahman et al. [59] studied the microfloral changes in tuna mince during convection air-drying between 40 and 100°C. The drying temperature of 50°C or below showed no lethal effect on the microflora and showed a significant growth. The drying temperature of fish must be at or above 60°C to avoid microbial risk in the product. The actual optimum temperature above 60°C should be determined based on other quality characteristics of the dried fish. Recently the use of heat pump dryer is receiving attention in the food industry due to its several advantages. Potential improvements in the quality of dried products are major advantages

in using heat pump drying. One of the main reasons of quality improvements in heat pump dried products is due to its ability to operate at lower temperatures. Adequate measure should be considered when using heat pump drying below 50°C for highly perishable products such as fish [59].

Reducing the water activity of a product inhibits growth, but does not result in a sterile product. The highest possible drying temperatures should be used to maximize thermal death even though low drying temperatures are best for maintaining organoleptic characteristics [49]. Other alternative is to use high drying temperature initially at high moisture content and then drying at low temperature. It is usual to estimate *D*-value at a specified temperature (isothermal conditions) by maintaining other parameters (such as moisture content) constant. This ideal situation cannot be simulated in the destruction process of microflora during drying. This is due to the change of moisture in the sample during drying process, thus destruction is caused by a combination of temperature and water loss. The microbial deactivation kinetics depends on several factors: variety, water content (i.e., water activity), temperature, composition of the medium (acidity, types of solids, pH, etc.), as well as heating method [35,42,76]. Models to predict the *D*-values were also developed as a function of temperature, pH, and water activity for isothermal conditions [12,23]. These models could not be used in case of drying conditions since the level of water content does not remain same for each temperature studied. Bayrock and Ingledew [8] measured the *D*-values for the changing moisture content (i.e., drying) and for moist conditions (i.e., no change of moisture during heating). They estimated the *D*-values from the slope of log *N* versus time of drying and found that *D*-values for drying condition were much higher than the values from the moist heat. This indicated that heat resistance of microorganism increased significantly during drying compared to the moist heat conditions. During drying of tuna, Rahman et al. [58] found that decimal reduction time (*D*-value) for natural microflora varied from 12.66 to 2.64 h when drying temperature varied from 60 to 100°C, respectively. As expected the values were decreased with the increase of temperature, which indicates that increase in drying temperature increased the lethal effect. The *D*-values at 100°C was much lower than the drying temperature at 90°C or below. This may be due to the high drying rate at 100°C. Bayrock and Ingledew [7] also pointed that higher drying rate at high temperature may be the main cause of microfloral destruction. Rahman et al. [61] investigated the changes of endogenous bacterial counts in minced tuna during dry heating (convection

drying) and moist heating (heating in a closed chamber) as a function of temperature. The *D*-values for total viable counts decreased from 2.52 to 0.26 h for moist heating and 2.57 to 0.34 h for dry heating, respectively, whereas temperature was maintained constant within the range 60–140°C. In both cases, increasing temperature caused significant decrease in *D*-values, whereas the effect of heating methods was not significant. Thus the heat resistance characteristics of microorganisms in fresh tuna mince was not depended on the changing medium moisture content. They also identified types and characteristics of endogenous microbes present in fresh and dried tuna meat. It showed that the predominant microbes in the dried tuna were moderate osmotolerant and heat sensitive.

22.5.2 BROWNING REACTIONS

Browning reactions change color, decrease nutritional value and solubility, create off-flavors, and induce textural changes. Browning reactions can be classified as enzymatic or nonenzymatic with the latter remaining more serious as far as drying process concerned. Two major types of nonenzymatic browning are caramelization and Maillard browning. In addition to the moisture level, temperature, pH, the composition of all parameters affect the rate of nonenzymatic browning. The rate of browning is more rapid in the intermediate moisture range and decreases at very low and very high moistures. Browning tends to occur primarily at the center of drying period. This may be due to migration of soluble constituents toward the center region. Browning is also more severe near the end of drying period when the moisture level of sample is low and less evaporative cooling is taking place, which results in the product temperature rises. Several suggestions are found to reduce browning during drying. In all cases, it was emphasized that product should not experience unnecessary heat when it is in its critical moisture content range [49].

Maillard-type nonenzymatic browning reactions in processed meat products also contribute to their external surface color [78]. Pearson et al. [50,51] demonstrated that the main browning reaction involves the reaction of carbonyl compounds with amino groups, although lesser amounts of carbonyl browning also occur. Muscle usually contains small amounts of carbohydrates in the form of glycogen, reducing sugars, and nucleotides, whereas the amino groups are readily available from the muscle proteins. Browning occurs at temperatures of 80–90°C and increases with time and temperature [16]. A loss of both amino acids and sugars from the tissue occurs as a result of the browning reaction. Lysine, histidine,

threonine, methionine, and cysteine are some of the amino acids that may involved in browning [31].

Potter [54] identified that Maillard browning proceeds most rapidly during drying if moisture content is decreased to a range of 15–20%. As the moisture content drops further, the reaction rate slows so that in products dried below 2% moisture further color change is not perceptible even during subsequent storage. Drying systems or heating schedules are generally designed to dehydrate rapidly through the 15–20% moisture range so as to minimize the time for Maillard browning. In carbohydrate foods, browning can be controlled by removing or avoiding amines and conversely in protein foods by eliminating the reducing sugars [47,48,50,51].

22.5.3 LIPID OXIDATION

Fish oils or fats are more unsaturated than beef or butterfat, and they are usually classified as drying oils because they contain considerable proportions of highly unsaturated acids. The behavior of drying oils toward atmospheric oxygen is well known, and oxidation is a serious problem for commercial drying of fatty fish and seafood. The flesh of some fatty fish, such as herrings, contains a fat prooxidant that is not wholly inactivated by heat [5].

Lipid oxidation is responsible for rancidity, development of off-flavors, and the loss of fat-soluble vitamins and pigments in many foods, especially in dehydrated foods. Factors which affect oxidation rate include moisture content, type of substrate (fatty acid), extent of reaction, oxygen content, temperature, presence of metals, presence of natural antioxidants, enzyme activity, UV light, protein content, free amino acid content, and other chemical reactions. Moisture content plays a big part in the rate of oxidation. At water activities around the monolayer ($a_w \approx 0.3$), resistance to oxidation is greatest.

The elimination of oxygen from foods can reduce oxidation, but the oxygen concentration must be very low to have an effect. The effect of oxygen on lipid oxidation is also closely related to the product porosity. Freeze-dried foods are more susceptible to oxygen because of their high porosity. Air-dried foods tend to have less surface area due to shrinkage, and thus not affected by oxygen [59]. Minimizing oxygen level during processing and storage and addition of antioxidants as well as sequesterants were recommended in literature to prevent lipid oxidation [49].

Antioxidants added to the herrings before drying are ineffective, but the addition to the air during drying of wood smoke, which contains some of the simple antioxygenic phenols, stabilizes the fat of the dehydrated products very considerably [5]. Oxidation

of the fat normally occurs during dehydration. Herings dried at 80 to 90°C compared to lower temperature were found to be more stable during storage [5]. One factor that may be important is the production of browning products, which have antioxidant activity. The effectiveness of nonenzymatic browning products in preventing lipid oxidation was demonstrated and it is one of the mechanisms hypothesized by Karel [36] to prevent lipid oxidation.

The effect of water on the destruction of the protective food structure in some specific dehydrated foods is probably involved in prevention of lipid oxidation in heated meat systems [36]. In systems in which there are both surface lipids and lipids encapsulated within a carbohydrate, polysaccharide, or protein matrix, the surface lipids oxidize readily when exposed to air. The encapsulated lipids, however, do not oxidize until the structure of the encapsulated matrix is modified or destroyed by adsorption of water [82]. Another reason is the increase of oxygen diffusion by increasing molecular mobility above glass-rubber transition [71].

22.5.4 CHANGES IN PROTEINS

The protein matrix in muscle has marked effect upon its functionality and properties [77]. The nonfatty part of fish is very susceptible to changes caused by high temperature of initial cooking, drying, and storage. Every process involved in the conversion of muscle to meat alters the characteristics of the structural elements [83]. Several functional properties may originate from the same internal change of proteins that form the tissue. Denaturation is defined as loss of natural properties such as tertiary or quaternary structure (amino acid sequence, primary structure and peptide strands in a protein, secondary structure). In addition to temperature, ionic environment in the tissue promotes changes in hydrogen bonding and disulfide links [89]. Heating is believed to cause the denaturation of the muscle proteins even below 60°C, but not enough to greatly shear resistance [78]. The decrease in shear observed at 60°C was attributed to collagen shrinkage. Hardening at 70–75°C was believed to be due to increased cross-linking and water loss by the myofibrillar proteins, whereas decreasing shear at higher temperatures may indicate solubilization of collagen [16].

After 1 h at 50°C, the collagen fibrils of the endomysium appear beaded, which is brought about by their close association with the heat-denatured non-collagenous proteins in the extracellular spaces. Heat denaturation of the lipoprotein in plasmalemma results from a temperature of 60°C for 1 h. The breakdown products of the plasmalemma are large granules and are often associated with the basement lamina,

which appears to survive intact even after heating at 100°C for 1 h [73,74].

Heating produces major changes in muscle structure. Voyle [88] reviewed modifications in cooked tissue observable with the scanning electron microscope. Alternation in muscle structure due to heating includes coagulation of the perimysial and endomysial connective tissue, sarcomere shortening, myofibrillar fragmentation, and coagulation of sarcoplasmic proteins [32,88]. Heating and drying intensify the detachment of the myofibrils from the muscle fiber bundles, which is caused mainly by electrical stunning or stimulation and improper conditioning following slaughter [14].

22.5.5 SHRINKAGE AND PORE FORMATION

Rahman [56] provides the present knowledge on the mechanism of pore formation in foods during drying and related processes. The glass transition theory is one of the proposed concepts to explain the process of shrinkage and collapse during drying and other related process. According to this concept, there is negligible collapse (more pores) in material if processed below glass transition and higher the difference between the process temperature and the glass transition temperature, the higher the collapse. The methods of freeze-drying and hot air drying can be compared based on this theory. In freeze-drying, since the temperature of drying is below T'_g (maximally freeze concentrated glass transition temperature), the material is in the glassy state. Hence shrinkage is negligible. As a result the final product is very porous. In hot air drying, on the other hand, since the temperature of drying is above T'_g , the material is in the rubbery state and substantial shrinkage occurs. Hence the food produced from hot air drying is dense and shriveled [2]. The values of T'_g for fish and meat varied from –11 to –15°C [9]. State diagram of tuna meat was developed by measuring freezing curve, glass line, and maximal freeze concentration conditions (X'_w and T'_m) [60]. Rahman [68] provided recent reviews on the development of state diagram. However, the glass transition theory does not hold true for all products. Other concepts such as surface tension, structure, environment pressure, and mechanisms of moisture transport also play important roles in explaining the formation of pores. Rahman [56] hypothesized that as capillary force is the main force responsible for collapse, counterbalancing of this force causes formation of pores and lower shrinkage.

The degree to which a dehydration sample rehydrate is influenced by structural and chemical changes caused by dehydration, processing conditions, sample preparation, and sample composition. Rehydration is maximized when cellular and structural disruption such as shrinkage are minimized [49]. Chang et al. [15]

illustrated the morphological changes that occur in the appearance of the muscle fiber bundles during cooking and drying convectionally in heated rotary dryer. They found that after cooking the fibers are bound together in a compact bundle. The bundle size is gradually reduced by the effects of heating and tumbling during the early stage of predrying in the modified clothes dryer. Apparent bundle size is expanded with the endomysial capillary moisture that is removed during drying.

The apparent shrinkage during processing can be defined as the ratio of the apparent volume at given conditions and initial apparent volume of the materials before processing. The apparent shrinkage coefficient indicates the overall volume shrinkage of a material. Two types of shrinkage usually observed in the case of food materials are isotropic and anisotropic shrinkage. Isotropic shrinkage can be described as the uniform shrinkage in all geometric dimensions of the materials. Anisotropic shrinkage is described as the nonuniform shrinkage in different geometric dimensions. In many cases, it is important to estimate the changes in all characteristics geometric dimensions to characterize a material. In case of muscle, such as fish and seafood, shrinkage in the direction parallel to muscle fibers was significantly different from that perpendicular to the fibers during air-drying [4,66]. This is different from the very isotropic shrinkage of most fruits and vegetables.

The generic shrinkage model was developed by Rahman [69]. Food materials can be considered as multiphase systems (i.e., gas–liquid–solid systems). When the mixing process conserves both mass and volume, then the density of the multiphase system can be written as

$$\frac{1}{\rho_T} = \sum_{i=1}^n \frac{X_i}{(\rho_T)_i} = \xi$$

where $(\rho_T)_i$ and ρ_T are the true densities of component i and composite mixture, respectively, X_i is the mass

fraction of the component i and n is the total number of components present in the mixture. Miles et al. [45] and Choi and Okos [17] proposed the above equation for predicting the density of food materials. However, this equation has limited uses in the cases where there is no air phase present and no interaction between the phases. Rahman [69] has extended the theoretical model, introducing the pore volume and interaction term into the above equation and the equation for apparent density is

$$\frac{1}{\rho_a} = \frac{\xi}{1 - \varepsilon_{ex} - \varepsilon_a}$$

where ε_{ex} and ε_a are the excess volume fraction due to interactions, and void or pore volume fraction or porosity, respectively. The shrinkage can be written as

$$S_a = \frac{V_a}{V_a^0} = \frac{\xi (1 - \varepsilon_{ex} - \varepsilon_a)}{\xi^0 (1 - \varepsilon_{ex}^0 - \varepsilon_a^0)}$$

where V_a is the apparent volume of the material. This model is applied successfully during air-drying of calamari [63]. The formation of pores in foods during drying can be grouped into two generic types: one with an inversion point and another without an inversion point (Figure 22.1 and Figure 22.2). Figure 22.2a shows that during drying initially pores are collapsed and reached at a critical value, and further decrease of moisture causes the formation of pores until completely dried. Opposite condition exists in Figure 22.2. Figure 22.2 shows that pores increased or decreased as a function of moisture content. Pore formation in the case of calamari and squid meat showed type 2A [63,66]. Most of the porosity is predicted from the density data or from empirical correlation of porosity and moisture content. Mainly empirical correlations are used to correlate porosity. Rahman et al. [63] developed the following correlation for open and closed porosity in calamari meat during air-drying up to zero moisture content as

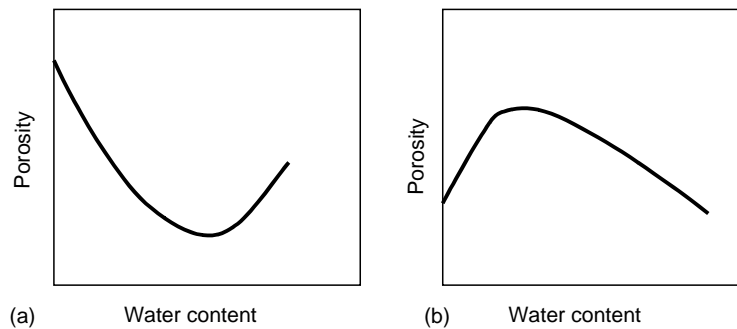


FIGURE 22.1 Change of porosity as a function of water content (with inversion point). (a) Porosity decrease with water content and then increase and (b) porosity increase with water content and then decrease [56].

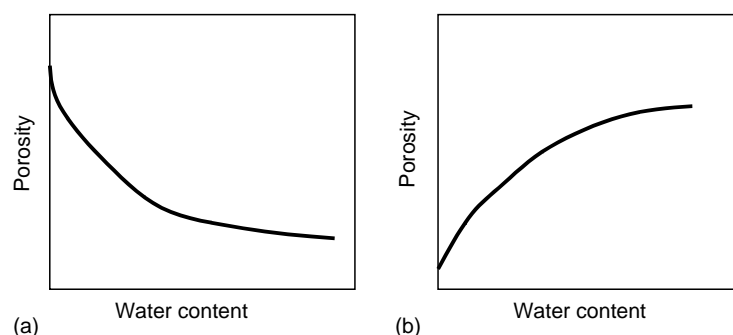


FIGURE 22.2 Change of porosity as a function of water content (no inversion point). (a) Porosity decrease with water content and (b) porosity increase with water content [56].

$$\varepsilon_{op} = 0.079 - 0.164y + 0.099y^2$$

$$\varepsilon_{cp} = 0.068 - 0.216y + 0.138y^2$$

where $y = X_w/X_w^0$. Rahman [69] developed apparent porosity of squid mantle meat during air-drying up to zero moisture content as

$$\varepsilon_a = 0.109 - 0.219y + 0.099y^2$$

Rahman [62] developed a theoretical model (ideal condition) to predict porosity in foods during drying based on conservation of mass and volume principle, and assuming that volume of pores formed is equal to the volume of water removed during drying. As expected the ideal model may not be valid in many practical cases. The ideal model is then extended for on-ideal conditions, when there is shrinkage, collapse, or expansion, by defining a shrinkage expansion coefficient. In addition to porosity, Rahman et al. [59] studied the characteristics of pores in dried tuna produced by air-, vacuum-, and freeze-drying. Pores in different dried tuna samples were characterized by porosimetry as the total intruded volume, total surface area, pore size range, average diameter, and nature of the pore size distribution curves.

22.5.6 REHYDRATION

Rehydration is a process of moistening dry material. Mostly it is done by abundant amount of water. In most cases, dried fish is soaked in water before cooking or consumption, thus rehydration is one of the important quality criteria. In practice, most of the changes during drying are irreversible and rehydration cannot be considered simply as a process reversible to dehydration [41].

In general, water absorption is fast at the beginning and thereafter slows down. A rapid moisture uptake is due to surface and capillary suction.

Rahman and Perera [57] and Lewicki [41] reviewed the factors affecting the rehydration process. These factors are porosity, capillaries and cavities near surface, temperature, trapped air bubbles, amorphous crystalline state, soluble solids, degree of dryness, anions, and pH of soaking water. Porosity, capillaries, and cavities near surface enhance the rehydration process, whereas the presence of trapped air bubbles gives a major obstacle to the invasion of fluid. Until the cavities are filled with air, water penetrates to the material through its solid phase. In general, temperature strongly increases the early stage of water rehydration. There is a resistance of crystalline structures to solvation, which causes development of swelling stresses in the material, whereas amorphous regions hydrate fast. Presence of anions in water affects volume increase during water absorption.

22.5.7 SOLUBILITY

Many factors affect the solubility such as processing conditions, storage conditions, composition, pH, density, and particle size. It was found that increasing product temperatures is accompanied by increasing protein denaturation, which decreases solubility. Thus, more protein is denatured and its solubility gets decreased [49]. Removal of water by evaporation results in the formation of an amorphous state product.

22.5.8 TEXTURE

Factors that affect texture include moisture content, composition, variety or species, pH, product history (maturation or age), and sample dimensions. Texture is also dependent on the method of dehydration and pretreatments. Purslow [55] stated that meat texture is affected by the structure of the solid matrix. He concluded that it is important to have a fundamental understanding of the fracture behavior of meat and how it relates to the structure of the material. Stanley

[83] stated that many researchers now believe that the major structural factors affecting meat texture are associated with connective tissues and myofibrillar proteins. Moreover, two other components muscle membranes and water also deserve consideration not because of their inherent physical properties, but rather as a result of the indirect influence they have on the physical properties. It should be noted that sarcoplasmic proteins may be important for the same reason, although little information on their role is available. He suggested that these structures merit particular attention.

Kuprianoff [37] referred to the possible adverse effects of removing bound water from foods as (i) denaturation of protein by concentration of the solutes, (ii) irreversible structural changes leading to textural modification upon rehydration, and (iii) storage stability problems. Stanley [83] stated that the water-holding capacity of muscle is related to its sorption properties. The bound water in the muscle is primarily a result of its association with the myofibrillar proteins as indicated by Wismer-Pedersen [93]. Protein–water interactions significantly affect the physical properties of meat [27]. Changes in water-holding capacity are closely related to pH and to the nature of muscle proteins.

22.5.9 NUTRITIONAL VALUE

The dehydration of food is one of the most important achievements in the human history making them less dependent upon a daily food supply even under adverse environmental conditions [21]. In general, losses of vitamin B are usually less than 10% in dried foods. Dried foods do not greatly contribute to dietary requirements for thiamin, folic acid, and vitamin B-6. Although vitamin C is largely destroyed by heating–drying, meat *per se* is not a good source [16]. Even though most amino acids are fairly resistant to heating–drying, lysine is quite heat labile and likely to be borderline or low in the diet of humans and especially so in developing countries where high-quality animal proteins are scarce and expensive [21].

22.6 CONCLUSION

In this chapter a brief overview of fish drying is presented. The main focus is given to drying methods, pretreatment, and quality characteristics. In many cases, pretreatment is important to achieve desired level of quality. The microbial, chemical, physical, and nutritional quality characteristics related to fish and seafood are also summarized.

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23 Grain Drying

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23.1 INTRODUCTION

Grain has been an important agricultural commodity and primary food source for centuries. The present distribution of the world's population has made strong demands on grain-handling technology. Irrespective of whether it is international trade or demands within a country, grain needs low moisture levels for safe storage. Drying has always been the most common method of preserving grain. In the days of premechanization of agriculture, enough grain was usually stored by hanging ears of corn in barn lofts and attics to meet the needs of a community. As mechanization of agriculture spreads to meet the needs of a population that was rapidly growing and urbanizing, mechanical methods for drying large quantities of grain were needed. Grain now travels thousands of miles either in large grain-carrying ships

or in different types of carriers on wheels, and must reach its destination in a high-quality state. Proper drying of these huge quantities of grain is a prerequisite to safe storage and delivery.

In 1999, an estimated 884 million metric tons of coarse grains were produced in the world [1]. Assuming harvested grain moisture and storage moisture to be in the range of 20–30% and 10–13%, respectively, 70–197 million tons of water had to be removed from this crop. This water-removal process is very energy consumptive. Therefore, it becomes evident that the efficiency of grain drying, with respect to both energy and time, has important economic consequences for both grain producers and consumers.

Further, there are several advantages for making the grain dryer a standard part of the harvesting system in the Western agricultural sector. First, when a grain dryer is used, extra hours of harvesting

in harvest days each year are possible, potentially reducing the farmer's overall machinery investment. Second, earlier harvesting is possible with a grain dryer, allowing a crop to be harvested nearer to its ideal moisture content for minimizing field loss. This permits farmers to do a better job of weed control through timely chemical application and tillage practices for the following year's crop. Third, weather damage and losses due to wildlife may be reduced by harvesting at the tough or damp stages and then drying the grain. Last, proper drying and aeration of tough or damp grain reduces or eliminates spoilage problems during storage due to hotspots and insect infestation. Not all grain dryers are suitable to a given geographical area and farm. The choice of a system depends upon the annual volume produced, the marketing pattern, the type of farm, and the kind and capacity of existing facilities. This chapter is intended to provide an introduction of the various types of grain dryers presently available on the market so that the reader may understand how a particular dryer is selected for a given farming operation. An attempt has also been made to indicate the importance of solar drying, nonconventional methods of drying, and some aspects of hay drying.

23.2 CROP CONDITIONING

Although the purpose of this chapter is primarily to discuss "heated air" dryers, some mention will be made about the four crop moisture reduction methods, usually called crop conditioning, as they are sometimes used in place of or in combination with heated-air drying.

23.2.1 AERATION

Aeration consists essentially of moving small amounts of unheated air through a pile of grain to equalize the grain temperature and to prevent moisture migration in bins exposed to drastic changes in ambient temperature. It may also be used to cool grain after drying, to keep damp grain cool until it can be dried, to remove storage odors, or to distribute fumigants in the grain mass.

Aeration is usually carried out in a storage bin that is equipped with a fan, duct system, and perforated floor along with exhaust vents to provide escape for moist air. Whether the ventilating air is blown upward or sucked down through the grain is largely a matter of choice. Upward ventilation is more commonly used, although there are advantages and disadvantages to each of these methods. An important advantage of using upward ventilation is that it

allows storage temperatures to be measured easily because the warmest grain is always at the top of the pile. The recommended airflow rate for normal aeration of shelled corn, soybeans, and small grains at 125 Pa (0.5 in. of water) is 5 m³/h per m³ of grain (0.1 cfm/bu) [2]. However, for aerating damp grain at 500–750 Pa (2 to 3 in. of water), flow rates of approximately 50 m³/h per m³ of grain (1 cfm/bu) are needed [3].

It is important to note that the aeration fan should not be run when the relative humidity of the ambient air is too high. For example, during fall and winter the operator should select days when the average relative humidity is less than or equal to 70% [4] and the air temperature is more than −1.1°C (30°F). It should be further noted that bins of 40 m³ (1000 bu) or less generally do not require aeration if the grain put in the bins is dry.

23.2.2 NATURAL-AIR DRYING

Natural-air drying employs a similar setup but higher airflow rates than those used for aeration. Typical rates for a storage depth of 1.2 to 1.8 m (4 to 6 ft) of small grains, peas and beans, and shelled and ear corn are, respectively, 150 to 250 m³/h per m³ of grain (3 to 5 cfm/bu) and 250 to 500 m³/h per m³ of grain (5 to 10 cfm/bu) [2].

23.2.3 IN-STORAGE DRYING WITH SUPPLEMENTAL HEAT

In-storage drying with supplemental heat involves drying of a relatively large batch of grain *in situ* (i.e., in the storage bin). It is carried out in bins of varying capacity up to 100 tons [5]. Ventilation is accomplished by blowing slightly heated air, 4–12°C (7–22°F) above ambient temperature through a duct system or through one centrally placed cylinder, as is the case for batch drying. Drying by this method usually requires continuous operation of the ventilation system for about 1 to 3 weeks.

In-storage drying may also be carried out on a bar floor provided with a powerful and a satisfactory system of floor and lateral ducts. Airflow rates range from 80 to 165 m³/h per ton of grain. The advantages of this method are low cost and simplicity.

23.2.4 MULTISTAGE DRYING

The term multistage drying refers to any process that uses high-temperature drying in combination with aeration or natural-air drying. An outline of two such processes, dryeration and combination drying, follows.

23.2.4.1 Dryeration

Dryeration is the term referring to the two-stage process by which grain is dried in a heated-air dryer to within about 2% of its “dry” moisture content and then moved to and stored in an aerating bin for about 10 h [3]. This allows time for moisture within the kernels to move to the outside for easier removal. Aeration at airflows in the order of 25–50 m³/h per m³ of grain (0.5–1 cfm/bu) is then maintained for about 12 h. The advantages of this system are as follows:

1. The ability to use higher drying temperatures as the grain does not remain in the high-temperature dryer until it is completely dry.
2. Capacity increases of up to 60% of the grain drying system are possible as no cooling time in a high-temperature dryer is required.
3. The last few moisture percentage points, which are especially difficult to remove, are removed in the bin using the heat already contained within the grain, resulting in fuel savings of 20% or more.
4. The grain quality is improved by cooling the grain immediately after it comes out of the dryer.

If the air is blown up through the grain, there is often a considerable amount of condensation on the roof and walls of the bin. The grain must therefore be moved to another bin for storage. The amount of condensation on the roof can be reduced by pulling the air down through the grain or by cooling the grain immediately after it comes out of the dryer.

23.2.4.2 Combination Drying

Combination drying is an extension of the dryeration process and is used primarily for drying grains with very high harvest moisture (>25%) [6]. A high-temperature dryer is used to reduce the grain moisture content to about 19–23%. The grain is then moved to a bin dryer in which drying is completed using natural air or supplemental heat. With this method, the output of the high-temperature dryer is increased to two or three times that obtained when it is used for complete drying. In addition, energy requirements may be reduced by as much as 50%. Airflows for the bin drying portion of the process are between 45 and 90 m³/h per m³ of grain (0.9 and 1.8 cfm/bu).

The selection of dryeration or combination drying will depend on the amount of grain to be dried, its initial moisture content, and the cost of energy and capital investment involved. If small amounts of grain at relatively low moisture contents are to be dried, the

purchase of equipment for combination drying would not be warranted. Combination drying is more suited to high-moisture contents and large volumes of grain. In all cases, for bins 100 m³ or larger, aeration ducts large enough for airflows of at least 36 m³/h per m³ of grain (0.7 cfm/bu) should be provided. Because fully perforated bin floors allow the greatest variety of options, their installation should be seriously considered on all large, new storage bins.

23.3 ARTIFICIAL HEATED-AIR DRYING

An important consideration when dealing with any heated air process is drying temperature. Suggested drying temperatures vary depending on what the grain's use is to be. Table 23.1 lists a few recommendations for natural- and heated-air drying and the maximum drying temperatures to be used on grain for seed, commercial use, and animal feed. Drying time and airflow rate are also important. However, these vary according to the drying temperature and type of dryer used.

The two major types of heated-air grain dryers are bin dryers and portable dryers. Bin dryers are available in batch, recirculating, and continuous categories, whereas portable dryers are available commonly in recirculating and nonrecirculating types.

23.3.1 BIN DRYERS

Bin dryers are manufactured in many sizes and capacities and are used to obtain various drying rates. They are usually operated with lower airflow rates than other types, and hence are generally more energy efficient although slower than most other types of dryers. The general philosophy of bin dryer size selection is to be able to dry as much grain in 24 h as will be harvested in a normal day.

23.3.1.1 Batch Dryers

The least expensive setup for drying is the one using the “batch-in-bin” process. The main components of this system are a bin with a perforated floor, a grain spreader, a fan and heater unit, a sweep auger, and an under-floor unloading auger (Figure 23.1). The heater fan starts when the first load of grain is put in and continues to operate as long as is required to lower the average grain moisture content to the desired level.

Drying rate depends on several variables, such as drying time, grain depth, temperature of the heated air, and airflow rate. Final depth is selected by noting the pressure drop in a manometer. Airflow rates are determined from charts supplied with fan unit. Usually, a rate of 450 m³/h per m³ of grain (9 cfm/bu) is

TABLE 23.1
Recommendations for Drying Grain with Natural Air and Heated Air

	Ear Corn	Shelled Corn	Wheat	Oats	Barley	Sorghum	Soybeans	Rice	Peanuts
<i>Maximum moisture content of crop at harvesting for satisfactory drying:</i>									
With natural air, % ^a	30	25	20	20	20	20	20	25	45–50
With heated air, % ^a	35	35	25	25	25	25	25	25	45–50
Maximum moisture content ^b of crop for safe storage in a tight structure, % ^a	13	13	13 (Seed wheat, 12%)	13 (Seed oats, 12%)	13	12	11	12	13
Maximum relative humidity of air entering crop that will dry crop down to safe storage level when natural air is used for drying, %	60	60	60	60	60	60	65	60	75
Maximum safe temperature of heated air entering crop for drying when crop is to be used for seed, °C	43	43	43	43	41	43	43	43	32
Sold for commercial use, °C ^a	54	54	60	60	41	60	49	43	32
Animal feed, °C ^a	82	82	82	82	82	82			

Source: From Hall, C.W., *Drying and Storage of Agricultural Crops*, AVI Publishing Company, Inc., Westport, Connecticut, 1980, pp 366–367. With permission.

^aMoisture contents on wet basis: (a) higher temperatures than those listed may be used when the corn is dried under carefully controlled conditions so that the maximum temperature of the kernels does not exceed 54°C at any time; (b) if there is any possibility that the crop may be sold, use the lower temperature as listed for commercial use.

^bThe products are to be stored for long periods, the moisture content should be 1–2% lower than shown in this tabulation.

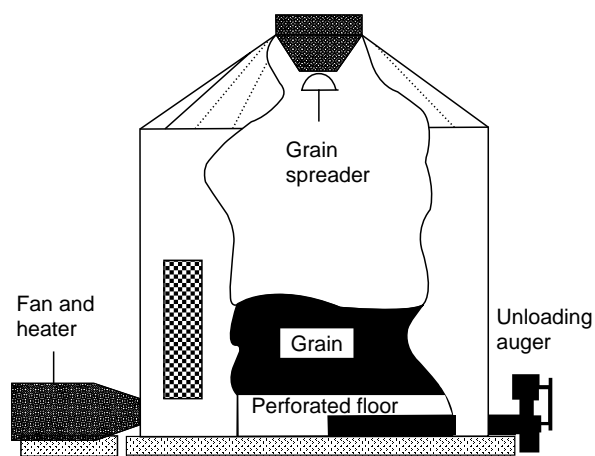


FIGURE 23.1 A typical batch dryer bin.

recommended for efficient drying. For a given grain depth, raising the air temperature speeds up drying but increases the chance of over drying near the floor. Hence, a safe air temperature is chosen for the crop that has to be dried considering its initial moisture content (Table 23.1).

Before storing, newly dried grain must be cooled. This is done by shutting off the heat and using the dryer fan to blow cool air over the grain, or by transferring the warm grain to an aerated storage bin and letting it cool there.

One variation of the batch-in-bin process is to use alternate heating and cooling cycles. This reduces the moisture differential between the drier grain near the perforated floor and the damper grain near the top of the grain column.

Some bin dryers have overhead, perforated, cone-shaped drying floors supported about 1 m below the roof (see Figure 23.2). A heater fan unit is installed

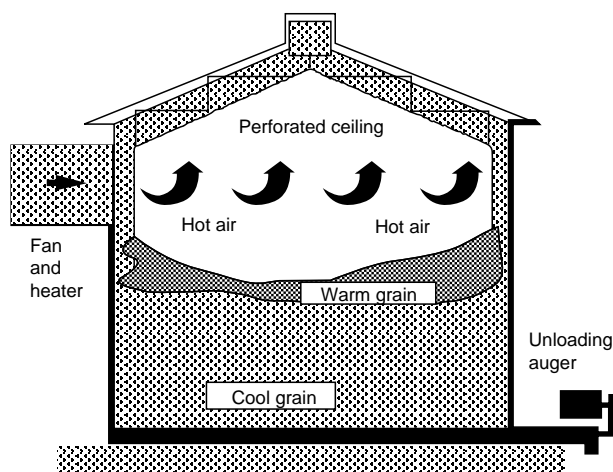


FIGURE 23.2 A bin dryer with an overhead drying floor.

below the perforated floor that blows warm air up through the grain.

When one batch of dry grain is dropped to a perforated floor at the bottom of the bin, where it is cooled by an aeration fan, the next batch is loaded and dried on the dryer floor above. Cool, dry grain is transferred to another storage bin via an auger under the floor. The advantage of this system is that drying can continue as the grain is being cooled and transferred.

Vertical stirring augers may be added to bin dryers, which not only promote more uniform drying, but also permit a higher airflow rate, thus increasing the drying rate for a given crop. Although stirring augers may result in slightly lower fuel efficiencies, the increased drying rate, reduction in over drying at the bottom, and the larger batch size outweigh this disadvantage.

23.3.1.2 Recirculating Dryers

In the recirculating type of dryer, grain is constantly mixed while drying. One example of a recirculating dryer is shown in Figure 23.3. A slanted floor causes the grain to move toward a vertical auger situated in the center of the dryer. The auger picks up the grain and delivers it to the top of the grain bin. The result is a more uniformly dried crop than that obtained using nonrecirculating types.

The dryer shown in Figure 23.4 is used as a recirculating batch or continuous flow dryer. When used as a recirculator, an “under grain” sweep auger moves grain to the center of the perforated bin floor where it is picked up by a vertical auger and delivered to a grain spreader. When the dryer is operated as a continuous flow dryer, the grain traveling up the vertical auger is transferred to an aeration bin via an inclined auger.

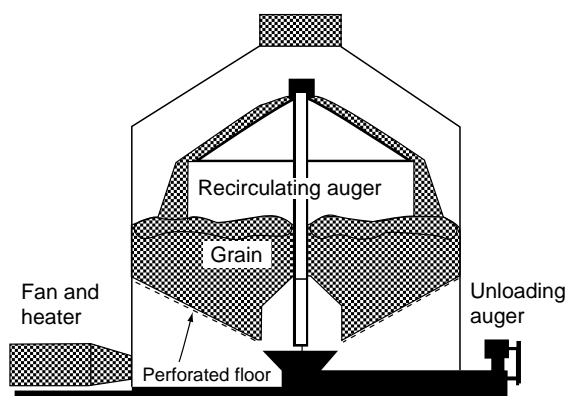


FIGURE 23.3 A recirculating batch bin dryer.

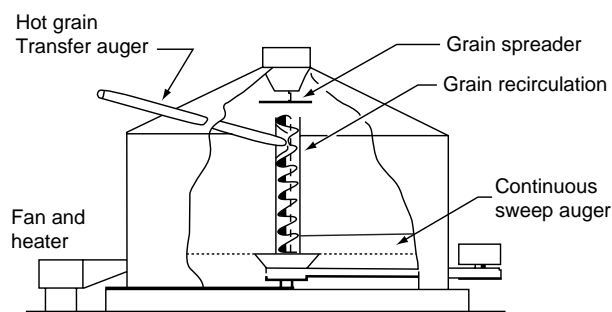


FIGURE 23.4 A recirculating bin dryer.

23.3.1.3 Continuous Flow Dryers

Although there are many types of continuous flow dryers, one of the more common types uses two to four vertical grain columns through which hot air is forced perpendicularly to grain flow (Figure 23.5). The grain is loaded at the top and passed down to both sides of the hot and cold plenums before entering the unloading augers. Grain flow rate is controlled either manually or by a thermostat near the outside of the grain column. As fan capacity is decreased or column width increased, more efficient use of heat results; however, the grain moisture differential between the inside and outside layers increases.

Some continuous flow dryers use three fans and three plenums, each with individual temperature controls. These may be run with two heating sections and one cooling section or else all three with heat, in which case the grain must be cooled in an aerated bin (Figure 23.6a and Figure 23.6b).

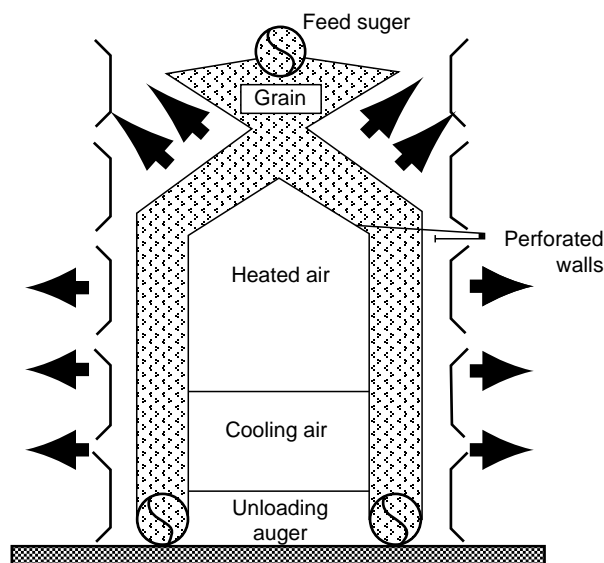


FIGURE 23.5 A typical stationary continuous flow dryer with an air recirculating system.

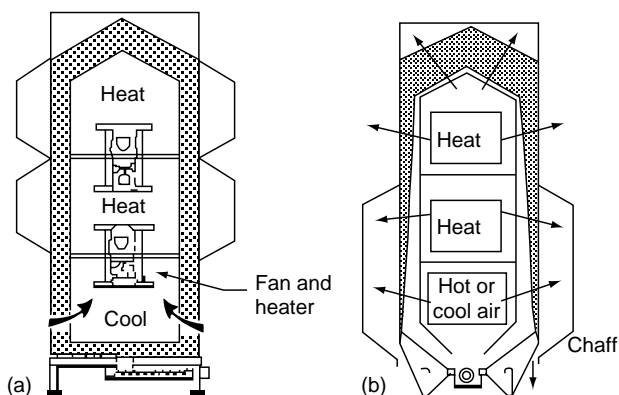


FIGURE 23.6 Heat recovery systems: (a) reverse cooling; (b) one-way airflow.

Farm Fans, Indianapolis, Indiana,* has a series of dryers of this type that they term continuous multi-stage dryers, ranging in capacity from about 5 to 27 tons/h (265 to 1220 bu/h) based on drying and cooling corn from 25% to 15% moisture.

A number of companies recycle drying or cooling air. Two common techniques of accomplishing this are shown in Figure 23.6. Some manufacturers use the system shown in Figure 23.6a. Here, ambient temperature air is drawn through the grain in the cooling section and then passed through the fan heater unit of the midsection. This system results in more energy saving than the system shown in Figure 23.6b due to the fact that air from the first heating section is recycled. Its disadvantage is that chaff and fine material may be drawn into the midsection hot air plenum, necessitating frequent cleaning.

Most continuous flow dryers are of the stationary type, although some of the smaller-size dryers are portable. For example, Gilmore and Tatge Manufacturing Company, Incorporated, Clay Center, Kansas, make a concentric cylinder type portable dryer that handles 7.8 ton/h (350 bu/h) based on moisture removal from 20.5% to 15.5%. Grain column width on many of these dryers is 0.30 m as compared to the 0.45 m found on the GT-Tox-o-Wik recirculating batch dryers. It should be noted that the moisture differential across the grain column is lowered as its width is decreased. A thinner column therefore means that, for a given average moisture content, the inner layer is less overdried. Thus, using a continuous flow dryer might be of some benefit when drying heat-sensitive small grains such as wheat, oats, and barley.

Another type of continuous flow dryer is the parallel flow dryer in which the grain moves in the same

*Mention of proprietary products in this article does not imply any recommendation or endorsement of their particular brands.

direction as the hot airflow. This results in more uniform drying and reduces the danger of heat damage. Furthermore, as no screens are used in parallel flow dryers, small seed crops can be dried without leakage.

Continuous flow dryers are not well suited for the drying of small quantities of different types of grains because startup and emptying of these dryers is inefficient. Accurate moisture control is difficult to achieve until a uniform flow is established. Continuous flow dryers are best in situations in which large quantities of grain must be dried without frequent changes from one type to another.

23.3.2 PORTABLE DRYERS

Portable dryers generally appeal to the farmer who has grain bins scattered in various locations, or who does custom drying off the farm. Portable dryers without a proper grain-handling system may be used to fill an immediate need in an emergency situation; however, they are normally not used where drying is beneficial but not necessary due to the inconvenience of setup and dismantling of the system. The two types of portable batch dryers are nonrecirculating and recirculating.

23.3.2.1 Nonrecirculating Dryers

Most nonrecirculating types of dryers are of the fully enclosed concentric cylinder type. These are loaded at the top and drying is accomplished by blowing hot air radially through a column of grain. Similar to the batch-in-bin drying system, the inside grain layer (the layer near the hot air plenum) becomes overdried as the outside layer remains under-dried. Nevertheless, as the grain is removed from the dryer, the damp and the dry grain are mixed so that a satisfactory product results for further use.

Other types of portable nonrecirculating batch dryers exist, such as wagon or truck-box dryers. These use a heater fan unit similar to that used for bin drying, and which is connected to smaller air ducts suspended at mid-height of the box or located on its floor. If suspended air ducts are used, exhaust ducts on the floor are a necessity.

These types of automatic dryers are equipped with thermostats or timers to control heating and unloading cycles. No manual supervision is required if a completely mechanized grain-handling system is used.

23.3.2.2 Recirculating Dryers

Portable recirculating batch dryers are essentially the same as nonrecirculating models except for a central

auger that picks up grain near the bottom of the column and deposits it at the top (see [Figure 23.7](#)). A complete recirculation of grain occurs roughly every 15 min.

Most common dryers of this type come in sizes ranging from 10 to 18.5 m³ (300 to 525 bu) bin capacity. These dryers are often used by medium-size farms in eastern North America that cannot afford a more expensive continuous flow model. The dryers may be used for virtually any crop, provided that the maximum safe drying temperature is not exceeded. However, their disadvantage is that constant augering can cause damage to certain seeds such as beans, peas, and malting barley, especially when they are nearly dry.

23.4 DRYER SELECTION

The selection of a continuous flow, batch, or batch-in-bin dryer depends largely on the amount of grain to be dried and the facilities already available with a farmer when the dryer is purchased. For example, a farmer who already has a good size storage bin and only a small volume of grain to store would likely use an in-storage dryer rather than purchase a portable dryer and “wet grain” holding bin. This system, however, would not be suitable for farms larger than 160 ha.

The recommendations concerning the type of system to be used can be made based on the annual production of a given farm, as illustrated in [Table 23.2](#). Although the capacity range presented here is for corn, it can also be extended to other grains and cereals. It must be further noted that these recommendations were made for farmers in the area of central United States. For eastern Canada, where temperatures are cooler and humidity is higher, the figures for natural-air drying presented in [Table 23.2](#) are slightly higher. In Quebec, for example, corn at harvest may have 35% moisture content with an average around 30%, whereas in western Canada and the midwestern United States, it is usually harvested at moisture contents in the low 20% range. This factor should therefore be considered when selecting a suitable drying system.

The crops most often dried in Canada and the United States by artificial means are corn (maize) and beans. Wheat, oats, and barley are harvested in the dry season and usually come off the field at a moisture content suitable for safe storage. If need be, the grain may be dried with natural air on sunny, warm days.

Most of the dryers in Canada are found in Quebec and Ontario [7]. Many of these are portable batch types. Larger, continuous flow models may be found at co-ops across the country or on the larger farms in southwestern Ontario, where farmers are growing 320–360 ha of their own crop.

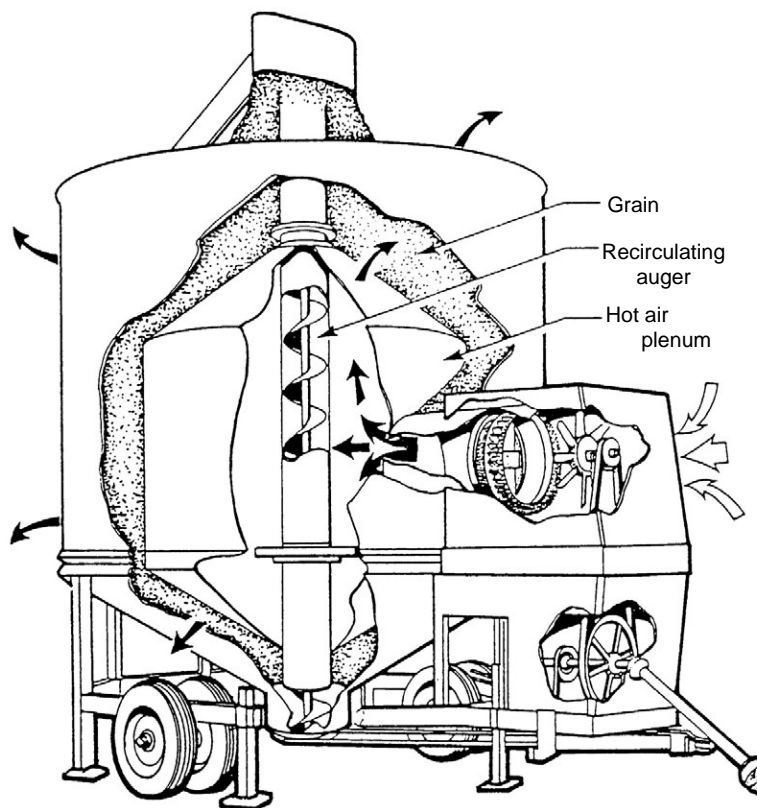


FIGURE 23.7 A typical portable batch dryer.

23.5 SOLAR ENERGY IN DRYING

An alternative drying method encouraged in hot, dry countries of Asia and Africa is solar drying. Solar heat is trapped with a solar collector constructed

TABLE 23.2
Recommended Drying System Based on the Annual Farm Production at Harvest

Annual Production	Type of Drying System ^a
22 to 60 tons (100 to 2700 bu)	Natural-air drying
60 to 445 tons (2700 to 20,000 bu)	Natural-air drying with supplemental heat
445 to 1556 tons (20,000 to 70,000 bu)	Batch-in-bin dryers
Above 1556 tons (>70,000 bu)	Portable and continuous flow dryers

^aBased on D.I. Chang, D.S. Chung and T.O. Hodges, Grain Dryer Selection Model, Am. Soc. Agr. Eng. Paper No. 79-3519, St. Joseph, Michigan, 1979.

from an aluminum sheet painted black. The collector may be fixed to the drying bin in such a way that an airspace exists between it and the bin wall. Energy absorbed by the collector heats the ventilating air by a few degrees as it is forced through the airspace. In North America, these types of dryers have been known to operate satisfactorily with grain moisture contents up to 25%, even on cloudy days [5]. The reason for this is that solar energy comprises of about half visible light and half infrared rays, with the latter having the ability to penetrate the clouds. On rainy days and nights supplemental heat may be supplied electrically.

In countries where harvesting time occurs at the beginning of the dry season, the most popular method of drying is exposure to the sun. Crops are often left to dry in the field before harvesting. In some countries, various crops are dried on scaffolds or inverted latticework cones. Another method is to lay paddy, maize, cobs, and other crops on heaps of stubble and then to cover them with stubble. At the village level, probably the most common practice is to spread the harvested threshed or shelled crop on the ground or on a specially prepared area (e.g., matting,

sacking, mud and cow dung mixture, or concrete) exposed to the sun.

In humid countries, initial crop drying may take place as outlined above; however, further drying is accomplished by placing the crop in a ventilating storage area. A more effective type of drying than sun drying is shallow-layer drying. This form of drying may be achieved by spreading the produce in a layer on the ground or on wire bottom trays that are supported above the ground. Cribs may also be constructed for drying maize on the cob or unthreshed legumes and cereals. These are usually oriented so that the long axis is facing the prevailing wind. They often have roofs or wide overhangs to protect the drying crop from rain.

In most warm, developing countries, a commercial dryer is too expensive and not essential enough for a single farmer to consider its purchase. China, India, and countries on the continent of Africa are examples of places where solar drying by direct exposure or by a cheaply constructed collector is employed.

23.6 ARTIFICIAL DRYING IN DEVELOPING COUNTRIES

Where humidity is too high to allow grain to be adequately dried by natural means, it is necessary to supply heat to the drying crop. The most popular forms of artificial drying may be categorized according to the depth of grain that is dried. These are: (a) deep-layer drying; (b) in-sack drying; and (c) shallow-layer drying.

Deep-layer dryers consist of silo bins (rectangular warehouses) fitted with ducting or false floors through which air is forced. Depths of up to 3.5 m of grain may be dried at onetime [8].

An in-sack dryer is made of a platform that contains holes just large enough to hold jute sacks full of grain. Heated air is blown up through the holes (and grain) via a heater or fan unit. The platform may be constructed from locally available material. A typical oil-fired unit that deals with two to five tons of grain is equipped with a fan that delivers 9700 m³/h of air heated to 14°C above ambient temperature and consumes about 4.5 l of oil per hour [8]. For two-ton loading, the moisture removal rate is about 1% per hour.

Shallow-layer dryers are those consisting of trays, cascades, or columns in which a thin layer of grain is exposed to hot air. In these dryers, the hot air stream is at the highest safe temperature and the amount of drying is determined by the length of time the grain is allowed to remain in the dryer, either as a stationary batch or as a slow-moving stream. Due to the fact that the layer of grain that is dried is thin (less than

about 0.20 m), no significant moisture gradient develops through the grain [8]. This means that the drying temperature is limited only by the possibility of heat damage to the grain.

Another simple but effective type of artificial dryer utilizes a locally built platform dryer in which the products of combustion of local fuel are not allowed to pass through the grain. The heated air passes through the produce by means of natural-air movement or convection currents. One such dryer built at Mokwa, Nigeria, uses a pit (which became the hot air plenum) covered by the drying floor, the firebox being located outside the plenum chamber.

Yet another type of dryer is the horizontal dryer, which contains a number of chambers, each being divided by horizontal, equidistant, screen-bottomed trays placed on horizontal pivots. Damp grain is placed on the top tray in a layer 0.16–0.18 m deep and is tipped to the next set of trays after an initial drying period. Because this type of dryer is normally operated as a batch dryer, it is an advantage to have two cooling chambers per unit so that one batch may be loaded into the dryer as the other is removed from the machine. A typical setup of this type would include a double drying chamber, a cleaning unit, and augers or elevating units for filling the dryer and elevating the grain to storage.

23.7 NONCONVENTIONAL METHODS

Recent increases in the cost of fossil fuels have prompted researchers to investigate and develop more energy efficient dryers [9–12]. One attempt at reducing fuel cost was to pass unheated air through large beds of absorbent material such as silica gel before passing it through the grain. The problem with this was that the gel itself had to be dried at high temperatures, making the operation expensive.

Heat pump dehumidifiers in drying equipment have been shown to offer many benefits. The removal of moisture from, and the subsequent transfer of the latent heat to the drying air enable drying at lower temperatures, lower cost and operation even under humid ambient conditions. Electrical heat pumps cause minimum environmental pollution.

Other methods related to enhancement of heat transfer techniques are also studied. Particle–particle heat transfer is one such technique that has led to the design of many experimental dryers. Richard and Raghavan have dealt with this topic extensively [13]. They discuss the theoretical aspects, experimental data, and demonstrate the potential of this method. The main advantage of this type of dryer is its rapidity.

Following this concept, a continuous flow conduction grain processor or dryer was developed in the late 1980s at McGill University, Quebec, Canada; it is shown in Figure 23.8 and fully described in a paper and two patents by Pannu and Raghavan [14–16]. It is based on particle–particle heat transfer and was designed to control mixing and heating time and provide ease of separation of the grain and the particulates.

The dryer consists of three concentric conical drums rotating about a common axis. The inner cone is fitted with a propane burner and buckets to carry the heat transfer medium toward the hottest part of the flame. The particulate medium then flows into the second drum, where it is mixed with moist grain and is carried in the opposite direction by the helical walls of the second drum. As the mixture proceeds toward the opposite end, it is separated by screen mesh, the grain carried on to the second cone outlet as the particulate medium drops into the outer cone and is recirculated to the heating chamber. The outer cone is insulated with R-10 glass wool to reduce heat losses. Grain residence time, and therefore the heating and moisture removal rates, is controlled by adjusting the rotating speed of the unit.

The dryer performance can be adjusted by varying different parameters such as the heating medium, medium and grain mass ratio, grain moisture, medium temperature, and angular velocity. The possibility of improving drying performance by using zeolites (molecular sieves) rather than sand was investigated [17]. It was found that the difference in moisture removed between the molecular sieves and sand was a function of residence time. When using sand, the

relative humidity in the dryer reached saturation within 2 min ($>90\%$), whereas it dropped to a steady value of 10–20% when zeolites were used. Thus the differences in moisture removal increased with time.

The high heat transfer efficiency makes the unit, using sand as a particulate, suitable for heat treatment applications such as pasteurization, precooking, insect eradication, and other applications in which moisture removal is not a priority. With zeolites as the transfer medium, moisture removal rates double, thus bringing drying efficiency up to the standards.

The enhanced drying performance using molecular sieves is encouraging. However, questions can be raised as to possible hazards associated with synthetic zeolite residues and as to nutritional quality of corn dried at higher temperatures.

The moisture removal increase was 50–130% higher for zeolites than sand, depending on the operating conditions. Upon nutritional analysis, no differences were found in digestibility or acceptability, as indicated by daily feed intake [18]. Unavailable proteins tended to increase with medium temperature; however, the increase was not significant.

23.8 HAY DRYING

Hay is also an important crop, like grain and cereals. In Canada and the United States, it is estimated that in any given year, 30% of hay is lost during harvesting and storage. Proper drying and handling techniques might reduce these losses. The hay is usually field dried to approximately 40% moisture and then dried

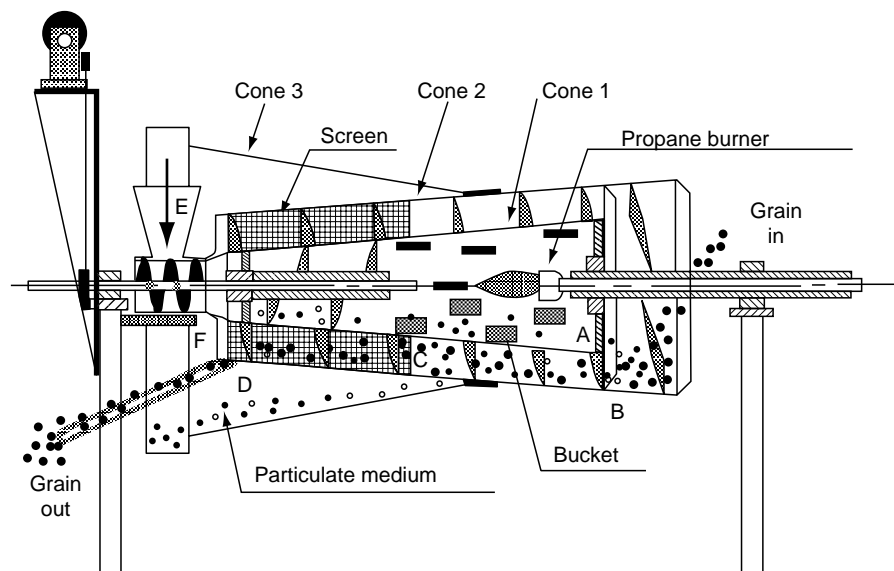


FIGURE 23.8 Schematic diagram of conduction grain processor.

to 20% with barn hay dryers. By employing suitable management techniques for barn drying systems, harvest losses can be reduced, produce can be harvested at its optimum stage of growth, and storage losses can be minimized. Although these advantages are acceptable, the number of barn drying systems has not increased in recent years because of the difficulty of providing a handling system compatible with the harvesting method. Forced-air drying and heated-air drying systems are generally used for hay drying. Further information on hay drying is given in Ref. [19].

ACKNOWLEDGMENTS

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24 Grain Property Values and Their Measurement

Digvir S. Jayas and Stefan Cenkowski

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24.1 INTRODUCTION

The world produces annually about 2 billion tonnes (Gt) of grains and oilseeds [1] that are handled and stored on- and off-farm for periods of up to 3 y. Often the storage period may be longer than 3 y, for example, when the grain is stored for potential famine relief. To design handling, inspection, and storage systems for

grains and oilseeds, data on many properties of individual seeds and seeds in bulk are needed. The properties of interest are: bulk and particle densities, porosity, roundness, sphericity, friction coefficients of grains against commonly used bin wall materials, emptying and filling angles of repose, equilibrium moisture content (EMC), specific heat, thermal conductivity, dielectric constant, electrical conductivity, reflectance, terminal

velocity, and drag coefficient. There are many methods of measuring these properties. In this chapter, only the methods that are currently in use or widely accepted methods are described. Representative property values for common grains and oilseeds are summarized from the published literature.

The properties of grains and oilseeds are measured on representative samples and are affected by many factors such as moisture content, growing location, amount and type of foreign material in the sample, and conditions of the surroundings. Many properties also depend on other properties. For example, thermal conductivity of bulk grain depends on its bulk density. Care must be taken in obtaining a representative sample because the great care taken in measuring a property will only give a good property value for the sample used. Ideally, representative samples from many different growing locations, growing years, and cultivars should be used to arrive at a grain property value that can be used in engineering design. Unfortunately, such studies become cost prohibitive and coordinated efforts among world scientists are needed to improve on the database for grain properties. At times, results of a well planned and executed project on measurement of grain properties become less useful when incomplete information about the sample is given in the published literature. Therefore, when reporting, a complete description of the sample should be given. For example, usefulness of data on bulk density of wheat without knowing at least its moisture content and class of wheat may be diminished considerably for an engineering design.

24.2 PHYSICAL PROPERTIES

24.2.1 BULK DENSITY (TEST WEIGHT)

Bulk density is defined as the ratio of the mass of the sample to the volume occupied by the bulk sample and is expressed in the units of kg/m^3 . The bulk volume includes the volume of intergranular air and grain. In the grain trade, the term test weight is used, which is defined as the mass of a measured volume of grain expressed in kg/hL (lb/bu). Standard methods for determining test weights are used by regulatory agencies around the world (e.g., Canadian Grain Commission [2]). The bulk density values reported in the literature are usually determined for clean grain at a specified moisture content by using the equipment and procedure for determination of test weight. The values thus determined are lower than the values that are expected in storage. The bulk density of grain in storage can be affected by the method used for filling the structure and by the amount and type of foreign material in the grain. For example, a bin filled using a spreader gives a

higher bulk density than a bin filled using a spout, or increasing the drop height during spout filling increases the bulk density. Also, the presence of foreign materials that are finer and heavier than grain kernels increases storage bulk density.

Bulk density of a clean grain sample (for corn, the sample is not cleaned) is determined by filling a 500-mL metallic container (90 mm diameter and 78 mm high) from a funnel with a 38.1 mm opening and having a flat slide gate. The opening of the funnel is maintained at 44.1 mm above the top of the container. Grain required to fill the container plus a small additional amount are loaded into the funnel whereas its gate is closed. When the gate is opened, the sample flows freely from the funnel into the center of the container and fills the container to overflowing. The grain in the container is leveled by striking off the excess grain with a round rod (19-mm diameter hardwood) using three equal zigzag movements at an angle to the direction of movement of approximately 45° . The mass of the grain in the container is measured and the bulk density (kg/m^3) or the test weight (kg/hL) is calculated (Canadian Grain Commission [2]). The bulk density for common grains and oilseeds are summarized in Table 24.1 (ASAE [3]).

TABLE 24.1
Approximate Bulk Density of Grains and Other Seeds

Seed	Bulk Density (kg/m^3)
Alfalfa	772
Barley	618
Beans Lima dry	721
Buckwheat	618
Canola (rapeseed)	669
Corn shelled	721
Lentils	772
Oats	412
Peanuts, unshelled	
Virginia type	219
Spanish	322
Rice, rough	579
Rye	721
Sorghum grain	721
Sunflower seed (nonoil)	309
Sunflower seed (oil)	412
Soybeans	772
Timothy seed	579
Wheat	772

Source: From ASAE, D241.4 Feb. 93, Density, specific gravity, and mass-moisture relationships of grain for storage, 40th ed., Standards, Engineering Practices, and Data (*Am. Soc. Ag. Eng.*), St. Joseph, MI, 1993, pp. 408–410. With permission.

TABLE 24.2
Porosity and Particle Density of Selected Seeds

Seed	Cultivar	Moisture Content (% wet basis)	Porosity (%)	Particle Density (kg/m ³)
Barley	Coast (6 rows)	10.3	57.6	1130
Barley	Hannchen	9.7	44.5	1260
Barley	Synasota	9.8	45.4	1210
Barley	Trebi (6 rows)	10.7	47.9	1240
Barley	White hullness	10.4	39.5	1330
Buckwheat	Japanese	10.1	41.0	1100
Canola	Tobin	6.5	38.4	1150
Canola	Westar	6.7	38.9	1100
Corn, mixed	Yellow and white	9.0	40.0	1190
Corn, shelled	Yellow, dent	25.0	44.0	1270
Corn, shelled	Yellow, dent	15.0	40.0	1300
Flaxseed		5.8	34.6	1100
Grain sorghum	Blackhull kafir	9.9	36.8	1260
Grain sorghum	Yellow milo	9.5	37.0	1220
Millet	Siberian	9.4	36.8	1110
Oats	Iowar	9.7	51.4	950
Oats	Kanota	9.4	50.9	1060
Oats	Red Texas	10.3	55.5	0990
Oats	Victory	9.8	47.6	1050
Rice	Honduras	11.9	50.4	1110
Rice	Wataribune	12.4	46.5	1120
Rye	Common	9.7	41.2	1230
Soybeans	Wilson	7.0	33.8	1130
Wheat, hard	Turkey, winter	9.8	42.6	1300
Wheat, hard	Turkey, winter (yellow)	9.8	40.1	1290
Wheat, soft	Harvest Queen	9.8	39.6	1320

Source: From ASAE, D241, 4 Feb. 93, Density, specific gravity, and mass–moisture relationships of grain for storage, 40th ed., Standards, Engineering Practices, and Data (*Am. Soc. Ag. Eng.*), St. Joseph, MI, 1993, pp. 408–410. With permission.

24.2.2 PARTICLE DENSITY

Two other terms true density and kernel density are used as synonyms to the particle density. Particle density is defined as the ratio of the mass of a sample to the volume occupied by the kernels (excluding the intergranular air) of the sample. The kernel volume can be determined using either of these two methods: liquid displacement method (LDM) and air comparison pycnometer (ACP) (e.g., Model 930, Beckman Instruments Inc., Fullerton, CA). In the LDM, a known mass of grain is poured into a graduated cylinder filled with a liquid to a known level. The change in the volume of the liquid is determined by subtracting the initial liquid volume reading from the final liquid volume reading. The liquid should not be sorbed (adsorbed or absorbed) by the solid particles; therefore, toluene is commonly used for grains and oilseeds. The main problem with the LDM is that tiny air bubbles on the surface of the solid particles may be present and included in the volume of the particles.

Also, liquid may not displace all the intergranular air. The ACP measures the true volume of the solid particles. The particle densities of some grains are given in Table 24.2 (ASAE [3]).

24.2.3 POROSITY

The terms percent voids, percent airspace, percent pore volume, and porosity are used interchangeably in the literature. Porosity is defined as the ratio of the volume occupied by the intergranular air to the total volume of the bulk sample and is expressed in percent. Usually, the porosity is calculated from the bulk and particle densities of a sample and thus can be considered as a derived quantity*:

$$\epsilon = \left[1 - \frac{\rho_b}{\rho_t} \right] \times 100. \quad (24.1)$$

*Symbols are defined in the List of Symbols.

Porosity can be measured directly using a method described by Day [4]. The porosities for common seeds are given in Table 24.2 (ASAE [3]).

$$R = \frac{A_p}{A_c} \quad (24.2)$$

24.2.4 PROJECTED AREA

Kernels of all grains and oilseeds, when dropped on a flat horizontal surface, rest in their most stable position. The area covered (viewed from the direction perpendicular to the surface) by a kernel in its most stable position is defined as the projected area (mm²) and can be determined by tracing the kernel on a graph paper and by estimating the number of squares in the enclosed tracing. A planimeter can also be used for quantification of the area. The length and width of the minimum rectangle that encloses the projected area are defined as principal major and minor axes or length and width of grain kernels, respectively. The length obtained by tracing along the circumference of the projected area is defined as the perimeter of a kernel. The projected area, length, width, and perimeter of kernels can be measured easily using a digital image processing system [5] and the values for many seed types are given in Table 24.3.

24.2.5 ROUNDNESS

Roundness of a grain kernel is defined as the ratio of the kernel's projected area in its most stable position to the area of the smallest circumscribing circle:

24.2.6 EQUIVALENT VOLUME (EQUIVALENT DIAMETER)

A sphere whose volume is equal to the volume of a grain kernel is defined as the equivalent sphere of the kernel. The diameter of the sphere of equivalent volume is defined as the equivalent diameter of the kernel. The average equivalent volume of kernels can be determined by using a particle density method (see Section 24.2.2) and counting the kernels. The equivalent diameter is calculated from the measured volume as

$$d_e = \left(\frac{6V_e}{\pi} \right)^{1/3} \quad (24.3)$$

Commonly, the volume occupied by 1000 kernels (randomly selected) is measured and the average equivalent volume and the diameter are calculated and reported (Table 24.4).

The geometric mean diameter, D_g , is another term that is used to describe the shape of kernels. It is calculated as

$$D_g = (a \times b \times c)^{1/3} \quad (24.4)$$

Physically, the terms a , b , and c are the length, width, and height, respectively of the smallest parallelepiped that will fully enclose the kernel.

TABLE 24.3
Physical Dimensions of Various Seeds

Seed	Projected Area (mm ²)		Perimeter (mm) ^a		Length (mm)		Width (mm)	
	Mean	SD ^b	Mean	SD ^b	Mean	SD ^b	Mean	SD ^b
Wheat (HRS) ^c	15.0	2.2	14.6	1.1	5.3	0.4	3.2	0.4
Barley	23.0	2.9	20.5	1.7	8.3	0.8	3.4	0.3
Canola	2.4	0.5	5.1	0.6	1.6	0.2	1.5	0.2
Brown mustard	1.9	0.3	4.5	0.5	1.5	0.2	1.3	0.2
Yellow mustard	3.9	0.6	6.8	0.6	2.2	0.2	1.9	0.2
Oriental mustard	2.4	0.3	5.1	0.4	1.7	0.2	1.5	0.2
Laird lentils	36.0	3.6	21.8	1.2	6.7	0.4	6.4	0.4
Eston lentils	16.0	2.0	14.5	1.0	4.5	0.3	4.3	0.3
Pea beans	40.0	4.6	23.4	1.4	7.8	0.5	6.1	0.5
Green peas	35.0	3.3	21.6	1.0	6.6	0.3	6.4	0.4
Black beans	45.0	5.7	25.3	1.7	8.4	0.6	6.6	0.6
Buckwheat	19.0	2.9	16.5	1.5	5.6	0.7	4.2	0.4
Flaxseed	6.4	0.5	10.1	0.4	3.9	0.2	1.8	0.1

^aRefer to Section 24.4.4.

^bSD = Standard deviation based on $n = 1000$.

^cHRS, Hard red spring.

Source: From Shatadal, P., Jayas, D.S., Hehn, J.L., and Bulley, N.R., *Can. Agric. Eng.*, 37(3), 163, 1995. With permission.

TABLE 24.4
Equivalent^a Diameter (d_e), Surface Area (A), and Volume (V_e) of Kernels of Selected Seeds

Seed	d_e (mm)	A (mm ²)	V_e (mm ³)	Ref.
Beans ^b	6.64	138	153	103
Corn	7.37	170	209	104
Corn	7.88	195	256	105
Corn, Inra 258	7.28	166	199	106
Corn, Velox	7.14	160	190	106
Corn, Dekalb XL72A ^c	7.61	182	231	107
Corn, Pioneer 3388 ^c	7.16	161	192	107
Corn, N7A X N28 ^c	8.18	210	286	107
Flax	1.90	11.34	3.59	104
Fababeans ^b	8.63	234	336	103
Lupin ^b	5.73	103	98	103
Peas ^b	6.28	124	130	103
Poppy seed	0.99	3.11	0.52	104
Wheat	3.48	38.04	22.07	104

^aEquivalent diameter was calculated from measured V_e and A was calculated from d_e .

^bMoisture content = 0% wb.

^cMoisture content = 10.5% wb.

24.2.7 SPHERICITY

Sphericity of a kernel is defined as: the ratio of the volume of a kernel to the volume of the smallest circumscribing sphere; or the ratio of the equivalent diameter of the kernel to the diameter of the smallest circumscribing sphere; or the ratio of the geometric mean diameter of a kernel to the diameter of the smallest circumscribing sphere.

24.2.8 SURFACE AREA

Surface area is the area of the outer surface of a kernel. An approximate surface area can be determined by assuming grain kernels are ellipsoids with major or minor axes a and b , and by calculating the surface area of the ellipsoid using known mathematical relationships [6]. An approximate surface area can also be estimated by assuming the kernel as a sphere of equivalent diameter. A method of measuring surface area of grains is by coating grains with a single layer of metal particles [7–9]. A known mass of kernels is dipped in varnish. The excess varnish is removed by rolling kernels on paper towel and by air drying. The kernels are swirled with nickel particles and the mass of coated kernels is determined. For converting the change in mass of the kernels to the surface area, particles of known geometry (surface area) having specific gravity similar to the specific gravity of grain kernels are coated in a similar manner and change in their mass is determined. By applying the proportionality rule the

surface area of particles is calculated. The surface area of some grains is given in Table 24.4.

24.2.9 EMPTYING AND FILLING ANGLES OF REPOSE

Emptying angles of repose of samples are measured by emptying grain from a box (e.g., a wooden box 430 mm long, 200 mm wide, and 430 mm high). The dimensions of the box are arbitrary except that the accuracy of measurement improves with increased length of slope. The box is filled with samples to a depth of 350 mm. Samples are allowed to flow out through a 50-mm high and 200-mm wide rectangular opening provided along the width of the box at the bottom of one end wall. Emptying angles are calculated from measurements of horizontal and vertical scale readings.

Filling angles of repose can be measured using a box with one side made of Plexiglass (e.g., a wooden box 1200 mm long, 100 mm wide, and 760 mm high). Samples are allowed to flow freely through a 50-mm square opening in a wooden hopper, whose center is maintained 1000 mm above the bottom of the receiving box. Filling angles are calculated from measurements of sample profile depth at two horizontally spaced points 300 mm apart. The first point is chosen approximately 100 mm away from the impact flattened apex of the cone. Filling angles are measured on both sides of the apex and averaged for individual replicates. Typical values of emptying and filling angles of repose of grains and oilseeds are given in Table 24.5.

TABLE 24.5
Emptying and Filling Angles of Repose of Selected Seeds

Seed	Moisture Content (% wb)	Emptying Angle (°)	Filling Angle (°)	Ref.
Barley (cv. 'Bedford')	12.7	26	24	108
Durum wheat (cv. 'Wakoma')	12.7	24	24	108
Oats (cv. 'Harmon')	12.7	27	28	108
Rye (cv. 'Gazelle')	12.7	21	25	108
Rapeseed (cv. 'Candle')	8.1	26	24	108
Sunflower (cv. 'Sundak')	8.1	22	21	108
Soybean (cv. 'McCall')	8.1	29	—	108
Triticale (cv. 'Carman')	12.7	21	23	108
Wheat (cv. 'Neepawa')	12.7	27	26	108
Fababeans (cv. 'Ackerperle')	12.6	28	29	10
Flaxseed ('McGregor')	7.0	30	26	10
Lentils (cv. 'Laird')	13.8	24	24	10
50%-Hulless barley (cv. 'Condor')	14.5	24	23	109
95%-Hulless barley (cv. 'Condor')	14.0	24	23	109

24.2.10 FRICTION COEFFICIENTS AGAINST STRUCTURAL MATERIALS

Coefficients of sliding friction against various structural surfaces are determined by using a tilting table. The surface of interest is attached to the tilting table. A wooden frame (305 mm long and 255 mm wide), made of 18-mm square wood, is placed lengthwise on the surface to prevent kernels from rolling down the surface. It is filled with the sample and leveled. The frame is lifted slowly to an approximate height of 2–3 mm, so that the frame does not rest on the surface. Using a manually driven screw, the table is tilted slowly until the sample starts to slide. The angle of the tilting table is measured using a protractor and a plumb bob. The coefficient of friction is calculated as the tangent of the angle measured [10]. The friction coefficient against vertical surfaces is needed in bin design. To measure the friction coefficients against vertical walls, a system described by Irvine et al. [11] can be used. The values of sliding friction against four structural materials for common grains and oilseeds are given in [Table 24.6](#).

24.3 HYGROSCOPIC PROPERTIES

24.3.1 MOISTURE MEASUREMENT

A common method for determining moisture content of grains and oilseeds is to dry samples in triplicate in a convection air oven at a specified temperature for a specified duration ([Table 24.7](#); ASAE [12]). About 10–15 g samples are weighed in covered aluminum dishes. The dishes are uncovered and placed with

their covers in the oven at the set temperature. At the end of the drying period, dishes are covered and placed in a desiccator for cooling to the room temperature. The dishes with dried sample are weighed again. Moisture content of samples is calculated as the ratio of the mass loss divided by the mass of the original sample and is expressed in percentage on a wet mass basis (wb). In theories of grain drying, the moisture contents of samples are usually expressed on a dry mass basis (db), which is calculated as the ratio of the mass of water divided by the mass of dry matter and is expressed as a percentage or decimal fraction. To avoid confusion as to which basis the moisture content is reported in a particular chapter, it is suggested that the scientific community adopt a convention to report moisture contents on wet basis in percentage and moisture contents on dry basis in decimal fraction.

24.3.2 MEASUREMENT OF EQUILIBRIUM MOISTURE CONTENT

There are two common methods (static and dynamic) for measuring the EMC–equilibrium relative humidity (ERH) relationships of grains and oilseeds. The static method is also known as the EMC method and the dynamic method is also known as the ERH method.

In the static method, a sample of known mass is allowed to reach equilibrium with air maintained at a constant relative humidity and temperature. The moisture content of the sample at equilibrium is measured and is defined as the EMC. The constant relative humidity environments are usually created using saturated salt solutions in containers [13].

TABLE 24.6
Coefficients of Sliding Friction against Four Different Structural Materials for Selected Seeds

Seed	Moisture Content (% wb)	Galvanized Steel	Steel Troweled Concrete	Wood Floated Concrete	Plywood	Ref.
Barley (cv. 'Bedford')	12.7	0.29	0.38	0.45	—	108
Durum wheat (cv. 'Wakoma')	12.7	0.29	0.39	0.45	—	108
Oats (cv. 'Harmon')	12.7	0.27	0.40	0.44	—	108
Rye (cv. 'Gazelle')	12.7	0.30	0.38	0.41	—	108
Rapeseed (cv. 'Candle')	8.1	0.24	0.30	0.39	—	108
Sunflower (cv. 'Sundak')	8.1	0.35	0.40	0.40	—	108
Soybean (cv. 'McCall')	8.1	0.27	0.33	0.34	—	108
Triticale (cv. 'Carman')	12.7	0.39	0.38	0.39	—	108
Wheat (cv. 'Neepawa')	12.7	0.32	0.42	0.47	—	108
Fababeans (cv. 'Ackerperle')	12.6	0.29	0.31	0.29	0.28	10
Flaxseed ('McGregor')	7.0	0.27	0.42	0.44	0.33	10
Lentils (cv. 'Laird')	13.8	0.25	0.34	0.31	0.24	10
50%-Hulless barley (cv. 'Condor')	14.5	0.29	0.39	0.40	0.31	109
95%-Hulless barley (cv. 'Condor')	14.0	0.32	0.40	0.41	0.32	109

A container with a 10- to 15-g sample suspended in the environment above the saturated salt solution is kept at a constant temperature. The experiment must be repeated at several temperatures and relative

humidities. The sample is weighed at a regular interval of 3–12 h until the change in sample mass between two successive readings is less than 0.01 g (at this stage it is assumed that the sample has reached

TABLE 24.7
Oven Temperature and Heating Period for Moisture Content Determinations

Seed	Oven Temperature $\pm 1^{\circ}\text{C}$	Heating Temperature		Sample Size (g)
		h	min	
Alfalfa	130	2	30	10
Barley	130	20	0	10
Beans, edible	103	72	0	15
Bluestem, yellow	100	1	0	1
Corn	103	72	0	15 or 100 ^a
Fescue	130	3	0	5
Flax	103	4	0	5–7
Mustard	130	4	0	10
Oats	130	22	0	10
Orchard grass	130	1	0	5
Parsnip	100	1	0	10
Rape (Canola)	130	4	0	10
Rye	130	16	0	10
Ryegrass	130	3	0	5
Safflower	130	1	0	10
Sorghum	130	18	0	10
Soybeans	103	72	0	15
Sunflower	130	3	0	10
Timothy	130	1	40	10
Wheat	130	19	0	10

^aUse 100 g if moisture exceeds 25%.

Source: From ASAE, S352, 2 Dec. 92, Moisture measurement—unground grain and seeds, 40th ed., Standards, Engineering Practices, and Data (*Am. Soc. Ag. Eng.*), St. Joseph, MI, 1993, p. 449. With permission.

equilibrium). Depending on the vapor pressure of the moisture in the seeds and the vapor pressure of the air above the saturated salt solution the sample may reach equilibrium by picking up or giving off moisture, thus resulting in either a sorption or desorption EMC value. For grains and oilseeds, the sorption EMC is lower than the desorption EMC at the same relative humidity because of the hysteresis phenomenon that is exhibited by biological materials. The moisture uptake can be by chemisorption, adsorption, absorption, or a combination as moisture content increases. The term sorption includes all mechanisms of moisture uptake. The time for the samples to reach equilibrium may vary from 1 to 5 weeks depending on the relative humidity and temperature. Therefore, mold usually develops on samples in high humidity environments and treatment of the sample with a mold inhibitor such as propionic acid is required.

To reduce the time to reach equilibrium and to reduce the problem with mold development, another version of the static method is to force conditioned air of known temperature and relative humidity over the sample until the change in mass of the sample is small (<0.01 g). A variation to this method can be used when determining thin-layer drying or wetting characteristics of a sample. The variation is that it is not necessary to bring the sample to equilibrium, and the EMC is determined as M_e by nonlinear regression of the equation:

$$\frac{M - M_e}{M_i - M_e} = \exp(-Kt^N) \quad (24.5)$$

In the dynamic method, a small amount of air is brought into equilibrium with a 0.5–1.0 kg sample of known moisture content by recirculating the air in a sealed unit [14] that is housed in a room at a constant temperature within $\pm 0.1^\circ\text{C}$. The relative humidity of the recirculating air is monitored until it becomes constant at which stage it is assumed that equilibrium has been attained and the measured relative humidity is the ERH. The moisture content of the sample is measured again. The average of the initial and final moisture contents is taken as the EMC. Because the amount of recirculating air is small, the change in moisture content of the sample usually is within the error limits of the method of moisture measurement and some researchers take the initial moisture content of the sample as the EMC. The time to reach equilibrium is reduced to 6–12 h depending on the conditions of the sample. When determining a desorption isotherm by the dynamic method, the initial relative humidity of the air must be well below the expected ERH so the grain loses moisture to the air; and the reverse must be guaranteed when determining the sorption isotherm.

24.3.3 ANALYSIS OF EMC–ERH DATA

The EMC–ERH data of grains and oilseeds are analyzed by fitting various equations to the data using nonlinear regression. The commonly used equations are: the modified Henderson, Chung–Pfof, Halsey, Oswin, and Guggenheim–Anderson–de Boer (G.A.B) (Table 24.8). The modified Henderson [15,16] and modified Chung–Pfof [17,18] equations have been adopted as standard equations by the American Society of Agricultural Engineers for describing EMC–ERH data for cereals and oilseeds. The modified Halsey [19,20] and modified Oswin [21,22] equations have been shown to describe the EMC–ERH data of many seeds satisfactorily [22,23]. The G.A.B. equation has recently been recognized as the most satisfactory theoretical isotherm equation, but it does not

TABLE 24.8
Equilibrium Moisture Content–Equilibrium Relative Humidity Relationships Used to Analyze Sorption and Desorption Isotherms of Grains and Oilseeds

Modified Henderson Equation

$$RH = 1 - \exp(-A(T + C)M^B)$$

Modified Chung–Pfof equation

$$RH = \exp\left(-\frac{A}{T + C} \exp\left(\frac{-BM}{100}\right)\right)$$

Modified Halsey equation

$$RH = \exp\left(-\frac{\exp(A + BT)}{M^C}\right)$$

Modified Oswin equation

$$RH = \frac{1}{\left(\frac{A + BT}{M}\right)^C + 1}$$

Guggenheim–Anderson–de Boer (G.A.B.) equation

$$M = \frac{ACB \times RH}{(1 - B \times RH)(1 - B \times RH + CB \times RH)}$$

Modified Guggenheim–Anderson–de Boer (G.A.B.) equation

$$M = \frac{A(C/T)B \times RH}{(1 - B \times RH)(1 - B \times RH + (C/T)B \times RH)}$$

A , B , C are constants, M is percent water content dry basis, RH is equilibrium relative humidity, decimal, and T is temperature, $^\circ\text{C}$.

Source: From Henderson, S.M., *Agric. Eng.*, 33(1), 29, 1952; Thompson, T.L., Peart, R.M., and Foster, G.H., *Trans. ASAE*, 11(4), 582, 1968; Chung, D.S. and Pfof, H.B., *Trans. ASAE*, 10(4), 552, 1967; Pfof, H.B., Maurer, S.G., Chung, D.S., and Milliken, G.A., Summarizing and reporting equilibrium moisture data for grains, Paper No. 76–3520, ASAE, St. Joseph, MI, 1976; Halsey, G., *J. Chem. Phys.*, 16, 931, 1948; Iglesias, H.A. and Chirife, J., *J. Food Technol.*, 11, 109, 1976; Oswin, C.R., *J. Soc. Chem. Ind. London*, 65, 419, 1946; Chen, C. and Morey, R.V., *Trans. ASAE*, 32(3), 983, 1989; Jayas, D.S. and Mazza, G., *Trans. ASAE*, 34(5), 2099, 1991; Jayas, D.S. and Mazza, G., *Trans. ASAE*, 36(1), 119, 1993.

TABLE 24.9
Constants of Selected Equations^a for the Isotherm of Various Seeds

Seed	Equation ^b	Isotherm Equation Constants		
		A	B	C
Barley	PF	475.12	0.14843	71.996
Corn (shelled corn)	HE	6.6612E-05	1.9677	42.143
Oats (cv. 'Dumont')	PF	433.157	21.581	41.439
Rough rice, long grain (Australia)	HE	4.1276E-05	2.1191	49.828
Medium grain (California)	HE	3.5502E-05	2.31	27.396
Short grain (Japan)	HE	4.8524E-05	2.0794	45.646
Wheat durum ('Wakooma')	OS	13.101	-0.052626	2.9987
Wheat hard red ('Waldron')	OS	15.868	-0.10378	3.0842
Wheat hard red ('Napayo')	OS	14.736	-0.05459	3.3357
Rapeseed ('Candle')	HL	3.0026	-0.0048967	1.7607
Canola ('Tobin')	HL	3.489	-0.010553	1.86
Flaxseed ('Linnot')	HE	0.000176	1.9054	56.228
Peanut kernel	HL	3.9916	-0.017856	2.2375
Safflower seed	HE	0.000203	1.8883	57.4013
Sunflower seed	HE	0.00031	1.7459	66.603

^aEquations are given in Table 24.8.

^bHE, modified Henderson; PF, modified Chung-Pfost; HL, modified Halsey; OS, modified Oswin.

Source: From ASAE, D245.5. Moisture relationships of plant-based agricultural products, 43rd ed., Standards, Engineering Practices, and Data (*Am. Soc. Ag. Eng.*), St. Joseph, MI, 1996. With permission.

incorporate the effect of temperature on the EMC–ERH relationship. Jayas and Mazza [24] modified the G.A.B. equation by dividing the constant C by temperature to give a three-parameter equation. The constants of the most appropriate equations for common seeds are given in Table 24.9 (ASAE [25]).

24.4. THERMAL PROPERTIES

24.4.1 SPECIFIC HEAT

Specific heat, c , is the amount of heat in kilojoules required to change the temperature of 1 kg of material by 1°:

$$c = \frac{1}{m} \frac{dq}{d\theta} \quad (24.6)$$

Specific heat of a moist agricultural product can be related to its dry, c_{db} , or wet, c_{wb} , mass. From the relationships between dry and wet masses it follows that

$$c_{db} = c_{wb}(1 + M) \quad (24.7)$$

It has been empirically shown that the specific heat of moist agricultural products can be presented as a sum of the specific heat of its dry mass, c_d , and the specific heat of water held in the product, c_w [26–29]. The

following relationship holds for 1 kg of dry mass of the product:

$$c_{db} = c_d + c_w M \quad (24.8)$$

Comprehensive reviews of the methods of measurement of thermal properties of grains have been published [28,30,31]. Precision and accuracy of measurement are important factors to consider when a method is considered for possible use. However, the variation in composition, size, and shape of agricultural products precludes the need for accuracies greater than ± 2 –5%.

The common procedures for measurement of the specific heat of grains at constant pressure are ice calorimetry [32], mixture methods [33], indirect methods, where the specific heat is calculated from other thermal properties such as thermal conductivity and diffusivity [34–37], method of differential scanning calorimetry (DSC) [38], guarded plate method, and the adiabatic method [28]. Only the most common method—the method of mixtures and the most modern method that utilizes sophisticated instrumentation—the DSC method, are discussed in this section.

The method of mixtures [33,39] consists of adding a known mass of material at one elevated temperature to a known mass of water at another temperature, and measuring the equilibrium temperature. The test

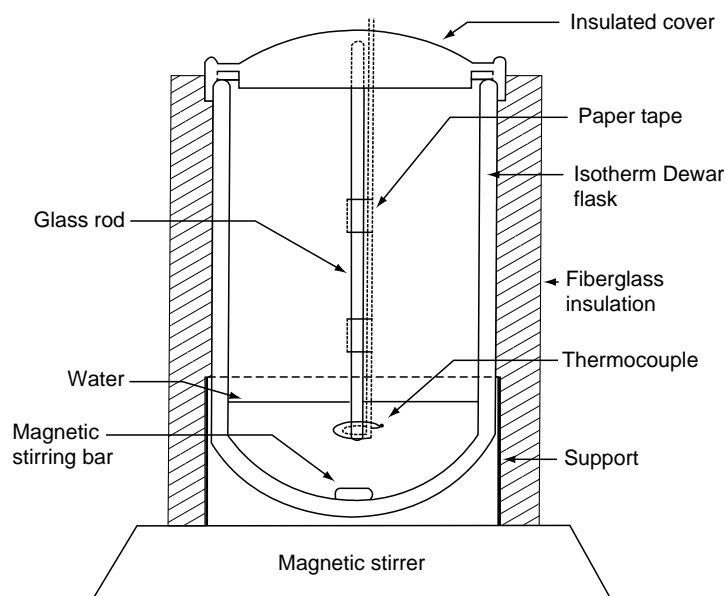


FIGURE 24.1 Schematic diagram of calorimeter. (From Viranichai, S., Effect of moisture content and temperature on specific heat of wheat. M.Sc. dissertation (unpublished), Department of Agricultural Engineering, University of Manitoba, MB, 1971. With permission.)

apparatus usually consists of an isothermal Dewar flask with a capacity of 1000 mL from which the metal jacket is removed to allow the use of a magnetic stirrer (Figure 24.1). The flask is insulated with approximately 4 cm of fiberglass insulation to minimize heat exchange with the surroundings. An insulating cover is divided into two parts to facilitate measurement of the calorimeter water temperature and to reduce heat losses while transferring a grain sample into the calorimeter. A glass rod, 13 cm long, is attached to the smaller part of the cover and protrudes into the calorimeter flask. A copper–constantan thermocouple (36-gauge) is attached to the end of the rod to measure the temperature of the calorimeter water. A magnetic stirrer is used to maintain a constant temperature throughout the flask. Calibration of the calorimeter can be accomplished using granular aluminum of known specific heat. The reliability of the equipment can be checked with granular copper as a reference sample. The typical testing procedure is as follows [33]: about a 20-g wheat sample (± 0.0001 g) at various initial temperatures (-36 to 21°C) is placed into the calorimeter flask filled with 200 ± 0.1 g of distilled water at room temperature, and the temperature of the water is recorded until thermal equilibrium of the mixture is established. A typical temperature–time curve of the water in the calorimeter is shown in Figure 24.2. The temperature of the mixture (T_m) is determined by extrapolating the straight-line portion of the temperature–time

curve back to the transient time, which is the time at which the sample is dropped into the calorimeter water. By equating the changes in the thermal energy contents of the water and the sample, the following equation can be written:

$$cm_s(T_m - \theta_s) = (m_w + E)c_w(T_c - T_m) \quad (24.9)$$

The water equivalent value, E , is obtained by rearranging Equation 24.9 and conducting experiments with a material such as water of known specific heat. The error associated with specific heat values obtained with the above-described apparatus is within 4.0% [40].

When using the method of mixtures water can be replaced with toluene. Toluene has the added advantages of a lower specific gravity (0.86) and specific heat ($0.39 \text{ kJ}/(\text{kg K})$) than water; thus enabling seeds to sink more readily than in water, and resulting in a substantially greater temperature rise than is obtained with water. With water, condensation on the calorimeter's surface usually occurs, causing a loss of measurement precision. Therefore, a calorimeter of low mass can be used [41]. The apparatus consists of two closely fitting, thin-walled aluminum tubes of about 2.5 cm diameter placed in a calorimeter (Figure 24.3). The inner tube is used to house chilled toluene and a sample. The outer tube is mounted by its top edge to a sealing disk, which is attached to the cap of a screw-top aluminum can. Air inside the can is dehydrated by desiccant. The can is packed in melting ice, inside a

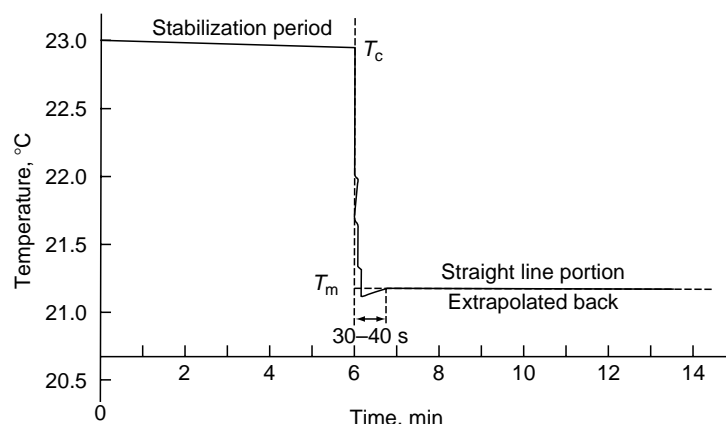


FIGURE 24.2 Temperature-time characteristic from calorimetric test. (From Viranichai, S., Effect of moisture content and temperature on specific heat of wheat. M.Sc. dissertation (unpublished), Department of Agricultural Engineering, University of Manitoba, MB, 1971. With permission.)

vacuum flask, to provide a constant and uniform surface temperature around the calorimeter. Temperature recording accuracy should be 0.01°C.

When low moisture content grain is used in the method of mixtures, absorption of water by the grain generates heat of hydration. This heat results in increased temperature rises in the water, which can lead to erroneous results. Therefore, to minimize the error, it may be necessary to incorporate a correction factor in the results [27].

The method of DSC is based on measuring the very small thermal effects produced during thermal processes and is recommended as well suited for determining the effect of temperature on specific heat [38,42]. In the DSC method, any gain or loss of thermal energy is recorded as the equipment warms the

test material at a controlled rate of temperature rise over a selected temperature interval. The disadvantages of this method are that it requires a small sample size (5–15 mg), which makes it difficult to obtain a homogeneous, representative sample, and it is a comparative device that must be calibrated. The measurement of specific heat with DSC is based on the assumption that the temperature is uniform in the sample and the sample pan during a test. However, due to the low thermal diffusivity of biological materials, thermal lag within a sample may introduce error in the measured specific heat.

The measured specific heats of dry mass of some agricultural products are given in Table 24.10. Because of differences in the chemical composition between crops and among cultivars of the same

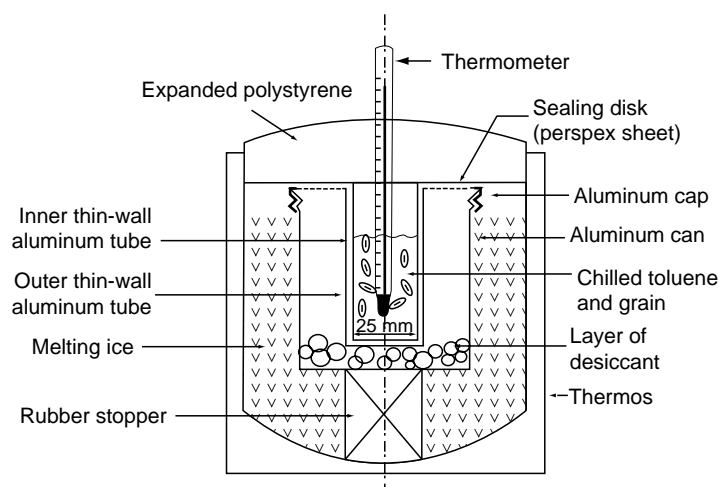


FIGURE 24.3 The calorimeter assembly. Modified from Sharp, R.B. and Nash, J.E., *J. Agric. Eng. Res.*, 10, 355, 1965. With permission.)

TABLE 24.10
Specific Heat of the Dry Mass of Agricultural Seeds

Seed	Specific Heat c_d , J/(kg K)	Ref.
Alfalfa	1172	110
Barley	1245	111
Beans	1293	111
Corn	1534	44
Corn	2035	105
Oats	1277	46
Oats	1282	111
Oats	993–1278	112
Rapeseed	1553–1569	111
Rice (rough)	1109	46
Rice	1637	47
Rice (white)	1197	46
Rye	1272	111
Sorghum	1397	27
Wheat	1097	33
Wheat	1185–1260	28
Wheat	1287–1299	112
Wheat	1276	111
Wheat	1454	44

crop, empirical equations have been developed. Usually, these equations describe the relationship between the specific heat and moisture contents of various crops [27,28,33,34,43–47] (Table 24.11):

$$c = a + bM + c_1\theta \quad (24.10)$$

Much of the published data on specific heat and other thermal properties of grains are of limited value because not enough supporting data are included, such as a detailed product description, and the estimated error in measurement. The description of grain should include the cultivar, the size of the individual kernels, the maturity, and the pretreatment. Details of an experiment should include the sample size, the surface conditions of the kernel, the porosity, the temperature, the relative humidity, the pressure, and the sampling procedure. The equipment description should provide sufficient detail so that one can duplicate the experiment.

24.4.2 THERMAL CONDUCTIVITY AND DIFFUSIVITY

Thermal conductivity, k , is described by the formula:

$$q = -kA \frac{d\theta}{dx} \quad (24.11)$$

For solid bodies the heat flux, q , is directly proportional to the temperature gradient, $\nabla\theta$, and the thermal conductivity, k , of the body. The minus sign refers to the direction of the flux, which is opposite to the

TABLE 24.11
Formulas for Specific Heat Determination of Selected Agricultural Products

Seed	Equation ^{a,b} and its Working Range ^c	Ref.
Alfalfa seed	$c = 1172 + 33M'$, $0 < M' < 28\%$	110
Chick-peas	$c = -4.19 \times 10^{-3} + 1.19 \times 10^1\theta + 2.15 \times 10^{-2}\theta^2 - 3.73 \times 10^4M - 1.65 \times 10^3M^2 + 1.38 \times 10^2M\theta$ $292 \leq T \leq 308 \text{ K}$, $0.12 \leq M \leq 0.32$	39
Corn (shelled)	$c = 1470 + 36M'$, $1 \leq M' \leq 30\%$	44
Lentils	$c = 1370 + 27M'$, $0 \leq M' \leq 60\%$ $c = [0.577 + 0.0071\theta + (0.0622 - 0.0914M) \times 10^2M]1000$ $10 < T < 80^\circ\text{C}$, $0.02 < M < 0.35$	45 113
Oats	$c = 1277 + 32M'$, $10 \leq M' \leq 17\%$ $c = 992 + 50M'$, $12 \leq M' \leq 18\%$	46
Rapeseed	$c = 1356 + 32M'$, at 19.4°C $c = 1288 + 28.4M'$, at 1.7°C $1 \leq M' \leq 20\%$ $c = 1328 + 28.0M'$, at -4.4°C	34
Rice (rough)	$c = 1109 + 45M'$, $10 < M' < 17\%$	46
Rice (white)	$c = 1197 + 38M'$, $10 < M' < 17\%$	
Sorghum	$c = 1397 + 32M'$, $0 < M' < 30\%$	27
Soybeans	$c = 1637 + 19M'$, $0 \leq M' \leq 24\%$	47
Wheat	$c = 1260 + 36M'$, $5 < M' < 35\%$ $c = 1098 + 40 \times 10^2M$, $0 \leq M < 0.25$ $c = 1184 + 30M'$, $0 \leq M' \leq 13.6\%$ $c = 1452 + 30M'$, $1 \leq M' \leq 32\%$	28 33 43 44

^a c , specific heat (J/(kg K)).

^b θ Measured as temperature of air in equilibrium with seeds.

^c M , moisture content (kg H₂O/kg db) and M' , moisture content % wb.

direction of the temperature gradient. Therefore, if the temperature gradient (for a certain time period) is constant, the thermal conductivity defines the amount of thermal energy that is transmitted within a unit time and through a unit cross-sectional area. The area is perpendicular to the direction of flow.

Factors influencing the choice of method for the determination of the thermal conductivity have been discussed in the literature [28,31,48]. Basically, methods of measurement of the thermal conductivity are divided into three categories: (1) steady-state techniques, (2) quasistatic techniques, and (3) transient-state techniques [28,31].

The advantages of the steady-state techniques are simplicity of the mathematical equations and high control of experimental variables. The disadvantages are: long equilibration time associated with possible moisture migration, restriction to the simple geometry of the sample (sphere, cylinder, or slab), and the presence of convection in granular materials. Thermal

conductivities of grains have been measured by the guarded hot-plate method [30] and the concentric spheres method [49].

The Fitch method [50] and its various modifications [31] are the most common quasistatic techniques used to measure the thermal conductivity. The main advantage of this method is that the test is simple and can be carried out in 10 min. For absolute measurements, however, the accuracy is rather low. Figure 24.4 shows a modified Fitch apparatus [51]. The sliced sample is placed between two copper plates. One plate acts as a heat source and the other plate as a heat sink. The thermal conductivity is calculated by Equation 24.12, which is the solution of the governing differential equation for the temperature field within the sample [51].

$$\ln \left(\frac{\theta_0 - T_\infty}{\theta(t) - T_\infty} \right) = \frac{Akt}{Lm_{cp}c_{cp}} \quad (24.12)$$

The plot of the temperature ratio $(\theta_0 - T_\infty)/(\theta(t) - T_\infty)$ versus time on a semilog paper is a straight line. The

thermal conductivity is calculated from the slope $(Ak/Lm_{cp}c_{cp})$ of the temperature history.

The essential difference between a steady state and a transient state is that the temperature at a particular location changes with time under transient conditions. A line heat source probe has been recommended by many researchers [28,29,52,53]. The method is simple, fast, and requires a relatively small sample. A schematic representation of the thermal conductivity probe, the direct current (dc) supply, and the temperature measuring system is shown in Figure 24.5 [54]. The probe is inserted into a sample of a uniform temperature and is heated at a constant rate. The temperature adjacent to the line heat source is recorded. Various modifications of the line heat source probe can be found in the literature. The probe attached to a 20-cm diameter aluminum cylinder as a sample holder is one of them (Figure 24.6) [35]. Other modifications are related to placement of thermocouples directly on the heating element [55,56] or at a fixed distance from it [43].

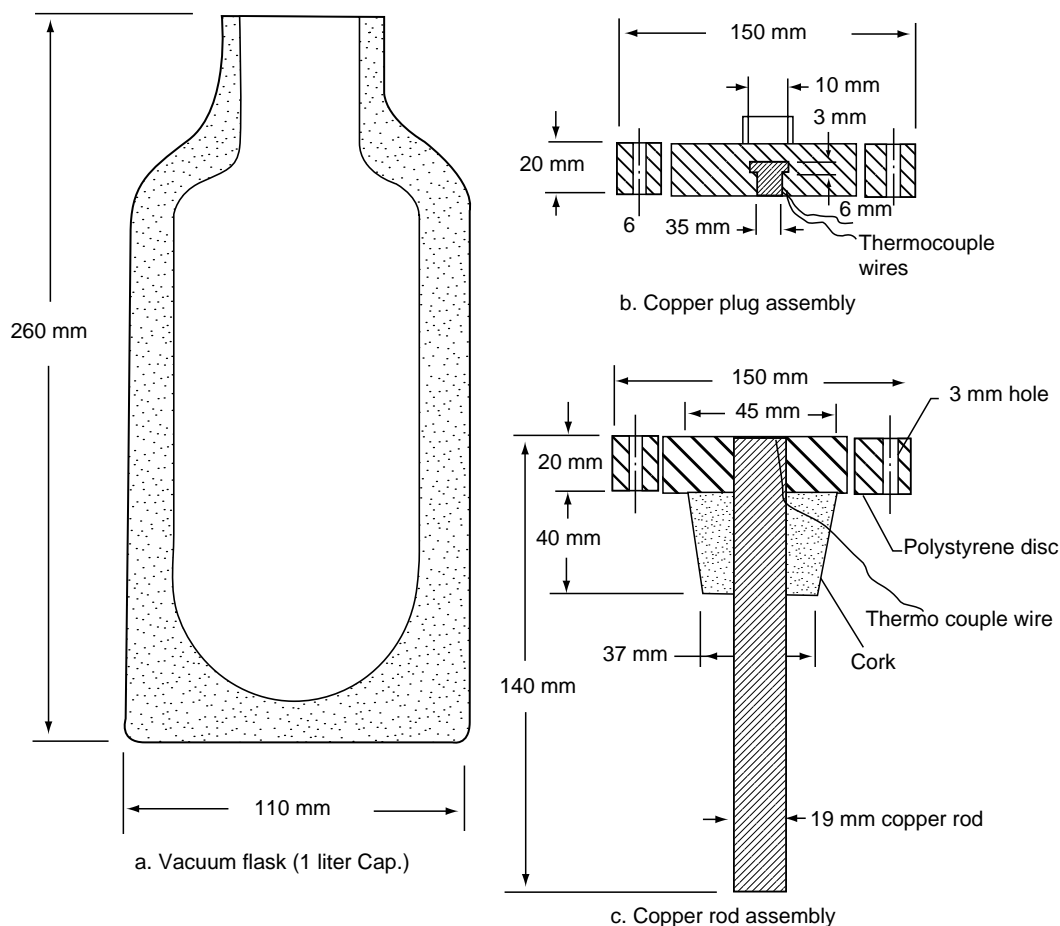


FIGURE 24.4 Modified Fitch apparatus. (From Zuritz, C.A., Sastry, S.K., McCoy, S.C., Murakami, E.G., and Blaisdell, J.L., *Trans. ASAE*, 32, 711, 1989. With permission.)

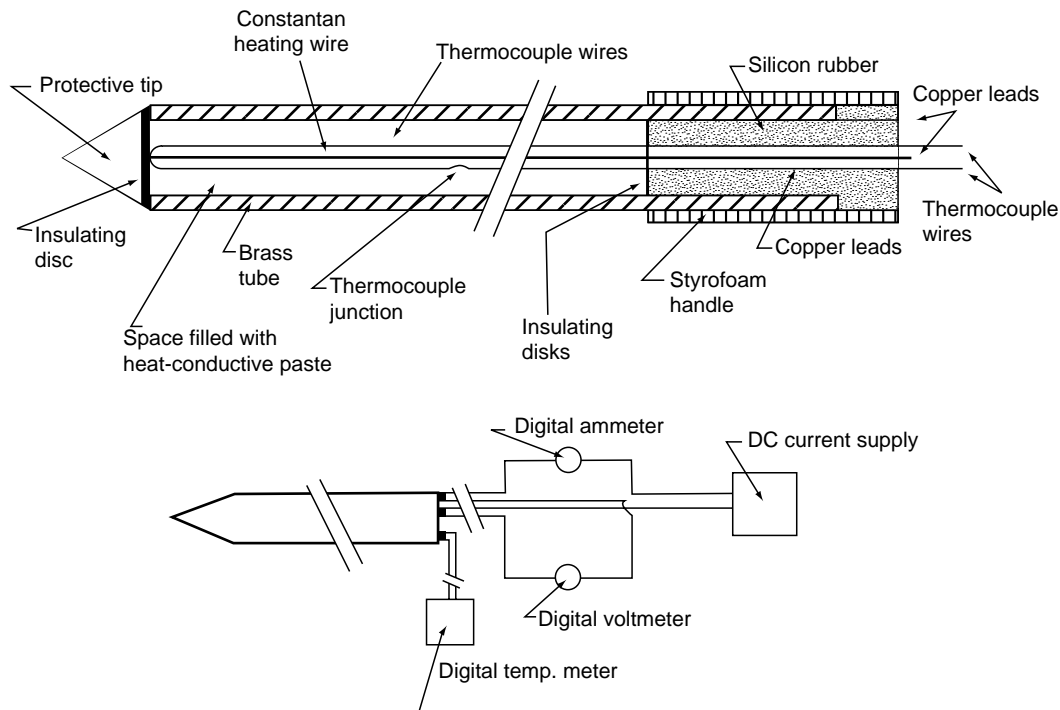


FIGURE 24.5 Thermal conductivity device. (From Papadakis, G., Giaglaras, P., and Kyritsis, S., *J. Agric. Eng. Res.*, 45, 281, 1990. With permission.)

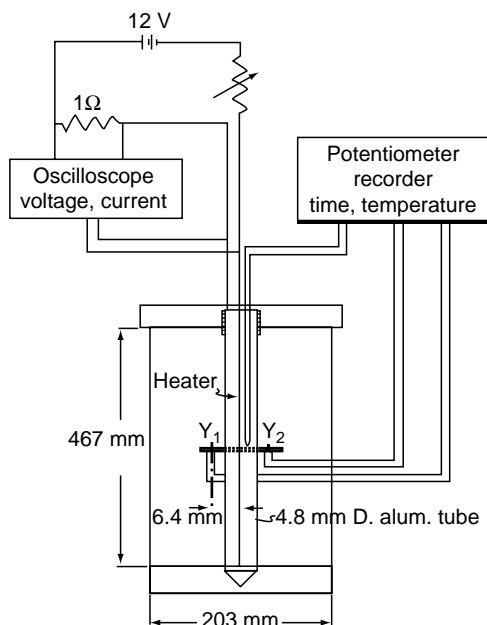


FIGURE 24.6 Schematic of thermal property apparatus. (From Suter, D.A., Agrawal, K.K., and Clary, B.L., *Trans. ASAE*, 18, 370, 1975. With permission.)

The determination of the thermal conductivity of grain is based on the comparison of the temperature history data obtained by using the line heat source

probe with the approximate analytical and numerical methods [35,54]. The analytical method has the advantage of being quick in calculating thermal conductivity. This method, however, requires a perfect line source and a small diameter tube holding the line heat source. In reality, this requirement is difficult to meet. Therefore, a time-correction procedure has been introduced [52,54,56]. Another objection to the analytical method is that it cannot easily be used to calculate the temperature distribution in the heated grain and to compare it with the measured one. Such a comparison can be easily accomplished by a numerical method, where the estimated accuracy for thermal conductivity is determined and the thermal conductivity of the device is taken into account [54].

The analytical method for determination of the thermal conductivity is presented below. The heat flow from the line heat source (a heating wire) of infinite length and infinitely small diameter imbedded in an infinite homogeneous medium can be expressed by the Fourier equation:

$$\frac{\partial \theta}{\partial t} = \alpha \left(\frac{\partial^2 \theta}{\partial r^2} + \frac{1}{r} \frac{\partial \theta}{\partial r} \right) \quad (24.13)$$

Equation 24.13 is solved for a particular case of the temperature at the line heat source for the heat input

of q_1 . The temperature rise $\Delta\theta$ in the time interval between t_1 and t_2 is determined as [52,57]

$$\Delta\theta = \frac{q_1}{4\pi k} \ln\left(\frac{t_2}{t_1}\right) \quad (24.14)$$

To accommodate the time-correction factor, t_0 , Equation 24.14 is modified to the following form [35,52,56]:

$$\Delta\theta = \frac{q_1}{4\pi k} \ln\left(\frac{t_2 - t_0}{t_1 - t_0}\right) \quad (24.15)$$

The time-correction factor, t_0 , takes into account various effects (i.e., contact resistances, positions of thermocouples, specific heat of the probe and sample) and has to be determined from experimental data. Thus, $dt/d\theta$ is plotted in relation to time. By differentiation Equation 24.14 gives

$$\frac{dt}{d\theta} = \frac{4\pi k}{q_1} t_c \quad (24.16)$$

Equation 24.16 can be represented as a plot of $dt/d\theta$ in relation to corrected time t_c . Thus, the expression $4\pi k/q_1$ represents a slope and at $dt/d\theta = 0$, $t_c = t_0$ which is the time-correction factor utilized in Equation 24.15.

The thermal conductivity of solid engineering materials varies with chemical composition, physical structure, state of the substance, temperature, and moisture content. Because grains are stored, ventilated, and dried in bulk, the bulk density of such products also influences their thermal conductivity. At constant moisture content the thermal conductivity can be expressed as a linear function of the bulk density, ρ_b :

$$k = a + b\rho_b \quad (24.17)$$

Coefficients a and b for wheat, corn, and grain sorghum at various moisture contents and at 22°C have been given by Chang [58]. Also, the thermal conductivity of many grains can be expressed as a linear function of moisture content:

$$k = k_d + a_1M + b_1\theta \quad (24.18)$$

Table 24.12 gives common relationships for determination of the thermal conductivity of selected major seed types. To improve the fit of data to mathematical expressions, polynomial equations are also used or temperature is introduced into the expression [39,56].

Unsteady-state or transient heat conduction commonly occurs during heating or cooling of grains. It involves the accumulation or depletion of heat within

a body, which results in temperature changes in the kernel with time. The rate at which heat is diffused out of or into a kernel or layer of kernels is dependent on the thermal diffusivity coefficient, α , of the grain:

$$\alpha = \frac{k}{c\rho_t} \quad (24.19)$$

The recommended method [29] for the determination of the thermal diffusivity of individual kernels is to calculate it from experimentally measured values of the thermal conductivity of kernel material, specific heat, and kernel (particle) density—the so-called indirect method. The method may lead to approximate results with a relative error, which is difficult to estimate in respect to a true (real) value, which can only be determined by direct measurements. The results of thermal properties for wheat and corn [44] and for single soybeans [59] confirm the above.

The thermal diffusivity can also be measured directly by employing transient heat conduction. The basic differential equation (Fourier heat conduction equation) governing heat conduction in isotropic bodies is used in this method. A rectangular copper box filled with grain is placed in an ice bath (0°C), and the temperature at its center is recorded [44]. The solution of the Fourier equation for the temperature at the center of a slab is used:

$$\frac{\theta_c}{\theta_0} = \frac{4}{\pi} \left(\exp(-\pi^2 z) - \frac{1}{3} \exp(-9\pi^2 z) + \frac{1}{5} \exp(-25\pi^2 z) - \dots \right) \quad (24.20)$$

and

$$z = \frac{\alpha t}{x^2}, \text{ dimensionless} \quad (24.21)$$

For selected values of θ_c/θ_0 obtained from the experiments, the values of z can be obtained, and with the measured values of time, t , and x the diffusivity is calculated from Equation 24.21. The values of the thermal diffusivities can also be determined for seeds by solving the Fourier equation for either an infinite cylinder [60], or a sphere [59]. **Table 24.13** gives values of thermal diffusivities for selected seed types obtained with the direct or indirect method.

24.4.3 CONVECTIVE HEAT TRANSFER COEFFICIENT

The main mechanism of air-to-particle heat transfer during heating, cooling, and drying processes of grains is forced convection. The forced convection heat transfer coefficient (surface conductance), h , is determined from the Nusselt number, Nu :

TABLE 24.12
Equations for Calculating Thermal Conductivity, k (W/(m K)), for Grains and Oilseeds

Seed	Equations ^{a,b} (W/(m K))	Range ^c	Ref.
Barley	$k = 0.173 + 7.51 \times 10^{-4}T + 1.51 \times 10^{-3}M'$	$9 \leq M' \leq 23\%$ $28 \leq T \leq 29^\circ\text{C}$	56
Beans	$k = 0.0671 + 32.84M$	$0.115 \leq M \leq 0.414$	51
Corn	$k = 0.1409 + 0.0011M'$	$0.9 \leq M' \leq 30.2\%$	44
Corn	$k = 0.1326 + 0.1547M - 0.1454M^2$	$0 \leq M \leq 0.6$	106
Chick-peas	$k = -5.07 \times 10^{-1} + 2.55 \times 10^{-3}\theta$ $-2.13 \times 10^{-6}\theta^2 + 4.24 \times 10^{-1}M$ $-6.56 \times 10^{-2}M^2 + 6.48 \times 10^{-4}M\theta$	$283 \leq T \leq 312\text{ K}$ $0.115 \leq M \leq 0.272$	39
Lentils (cv. Laird)	$k = 0.193 + 1.0 \times 10^{-3}T + 1.51 \times 10^{-3}M'$	$9 \leq M' \leq 23\%$ $-28 \leq T \leq 29^\circ\text{C}$	56
Oats	$k = 0.0988 + 0.307M$	$0 \leq M \leq 0.19$	114
Peas	$k = 0.168 + 8.4 \times 10^{-4}T + 3.05 \times 10^{-3}M'$	$9 \leq M' \leq 23\%$ $-28 \leq T \leq 29^\circ\text{C}$	56
Rapeseed	$k = 0.1600 + 0.043M$	$0 \leq M \leq 0.30$	115
Rice, rough	$k = 0.0865 + 0.0013M'$	$9.9 \leq M' \leq 19.3\%$	116
Rice bran	$k = -0.0943 + 3.87 \times 10^{-3}M'$ $+ 6.19 \times 10^{-4}T + 3.14 \times 10^{-4}\rho_b$	$7 \leq M' \leq 15\%$ $410 \leq \rho_b \leq 490\text{ kg/m}^3$ $42 \leq T \leq 68^\circ\text{C}$	117
Sorghum	$k = 0.0976 + 0.0015M'$	$0 \leq M' \leq 25\%$	27
Wheat (HRS)	$k = 0.1398 + 0.0014M'$, for $T = 20^\circ\text{C}$ $k = 0.1440 + 0.0009M'$, for $T = 5^\circ\text{C}$ $k = 0.1365 + 0.0014M'$, for $T = 1^\circ\text{C}$ $k = 0.1327 + 0.0015M'$, for $T = -6^\circ\text{C}$ $k = 0.1407 + 0.0009M'$, for $T = -17^\circ\text{C}$ $k = 0.1436 + 0.0009M'$, for $T = -27^\circ\text{C}$	$4.4 \leq M' \leq 22.5\%$	118
Wheat	$k = 0.1170 + 0.0011M'$	$0.7 \leq M' \leq 20.3\%$	44

^a k , Thermal conductivity W/(m K).

^b θ , Measured as temperature of air in equilibrium with seeds.

^c M , Moisture content kg H₂O/kg db and M' , moisture content % wb.

TABLE 24.13
Thermal Diffusivities of Selected Seeds

Seed	Moisture Content % wb	Thermal Diffusivity (α , m ² /s)	Method	Ref.
Chick-peas	12.0	11.6×10^{-8}	Indirect	39
Corn, yellow dent	9.8	9.4×10^{-8}	Direct	44
	20.1	8.6×10^{-8}		
Rapeseed	10.5	9.2×10^{-8}	Indirect	34
Rice, bran	7.0	9.7×10^{-8}	Indirect	117
Rice, rough	12.0	16.4×10^{-8}	Indirect	116
Soybeans	11.2	11.7×10^{-8}	Direct	59
Wheat, soft	10.3	8.3×10^{-8}	Direct	44
	20.3	8.1×10^{-8}		
Wheat	9.2	11.4×10^{-8}	Direct	43
Wheat	10.0	8.3×10^{-8}	Direct	119

$$Nu = \frac{hd_e}{k_a} \quad (24.22)$$

The Nu number is determined from empirical correlations between the Reynolds number, Re , and the Prandtl number, Pr :

$$Nu = CRe^m Pr^n \quad (24.23)$$

The Reynolds and the Prandtl numbers are expressed as

$$Re = \frac{vd_e}{\nu} \quad \text{and} \quad Pr = \frac{\nu}{\alpha} \quad (24.24)$$

For the temperature range used in drying grains, the Prandtl number is assumed to be constant and its value is usually incorporated into a constant, C , which simplifies Equation 24.23 to one variable—the air velocity (dependent on the Re number).

One of the first experiments to determine heat transfer coefficients commonly cited in agricultural engineering publications is that of Löf and Havley [61] who investigated heat transfer coefficients from air to a 0.9-m deep bed of granitic gravel ranging in size from 4.8 to 38.1. The temperature of the hot air passing through the gravel was monitored at selected locations through the bed. The experiments were conducted for air velocities from 0.08 to 0.44 m/s and the entrance air temperatures were maintained over the range 38 to 121°C.

To determine the mean volumetric heat transfer coefficient, h_{cv} , in a deep bed of gravel it was assumed that: (i) any particle is at a uniform temperature at any given time, (ii) resistance to heat transfer by conduction in the fluid or solid is negligible, and (iii) the rate of heat transfer is described by Newton's equation [61]:

$$h_{cv} = A \left(\frac{m_a}{d_e} \right)^{0.7} \quad (24.25)$$

The constant A depends on the nature of the material, the porosity, and the average temperature of the deep bed.

Using the same theoretical approach, the heat transfer coefficient in a 240-mm deep bed of bone-dry barley malt was determined [62]. The experiments were performed for airflow rates in the range 0.35 to 0.65 kg/(m² s) and inlet temperatures from 50 to 70°C. The results were expressed as

$$h_{cv} = 49,320 m_a^{0.6906} \quad (24.26)$$

The heat transfer coefficient can also be determined based on a heat and mass balance for a thin layer of

grain in which no moisture transfer takes place [63]. Thus, the following differential equation is used:

$$\frac{d\theta}{dt} = \frac{h_{cv}}{c_g \rho_b} (T - \theta) \quad (24.27)$$

with the solution

$$h_{cv} = \frac{c_g \rho_b}{t} \ln \left(\frac{T - \theta}{T - \theta_0} \right) \quad (24.28)$$

The slope of the logarithm of the temperature ratio versus time is used to evaluate h_{cv} for specific values of t and θ determined experimentally. For barley dried in a thin layer in the airflow range 0.0056–0.023 kg/s (assuming no shreokage) the following relationship is valid [63]:

$$h_{cv} = 21.0 m^{0.6} \quad (24.29)$$

The results from Equation 24.25, Equation 24.26, or Equation 24.28 can be used for the calculation of Nusselt numbers and presented in the form of Equation 24.23.

24.4.4 LATENT HEAT OF VAPORIZATION

To overcome the attractive forces between the adsorbed water molecules and the internal surfaces of grain kernels extra energy is needed in addition to the heat required to change the water from liquid to vapor. As the moisture content is lowered, there is an increase in the amount of energy required to evaporate the water molecules in seeds. The change of state from liquid to vapor at constant temperature and pressure is expressed by Clapeyron's equation, which after rearranging and integrating has the following form:

$$\ln P_v = \frac{h_{fg}^*}{h_{fg}} \ln P_{vs} + C \quad (24.30)$$

The vapor pressure, P_v , can be calculated as

$$P_v = RHP_{vs} \quad (24.31)$$

where RH is the relative humidity in decimal form.

The values of the saturation vapor pressure, P_{vs} , at different temperatures can be obtained from the steam tables [64]. The values of the RH in Equation 24.31 can be replaced by the ERH obtained from the EMC versus ERH relationship. The values of P_v and P_{vs} at the same EMCs can be plotted on a log–log scale. The slope of the resulting straight line gives the ratio h_{fg}^*/h_{fg} .

The following formula has been established to determine the dependence of the latent heat of vaporization of water in grain as a function of the moisture content, M [65]:

$$h_{fg}^* = h_{fg}[1 + a \exp(bM)] \quad (24.32)$$

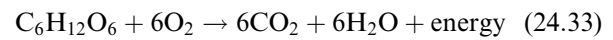
The coefficients a and b for selected seeds have been determined and are given in Table 24.14 [66]. The variation in the latent heats of barley and three cultivars of wheat are shown in Figure 24.7. The symbols represent the values of h_{fg}^*/h_{fg} calculated from the EMC data [25] for selected moisture contents. The curves represent the best-fit line based on Equation 24.32. Also, predicted results based on the coefficients given for wheat by Gallaher [65] are incorporated in Figure 24.7.

The differences in h_{fg}^* for wheat can be attributed to a difference in the chemical composition between wheat types, which affect the EMC characteristics on which the calculation of h_{fg}^* is based.

24.4.5 HEAT OF RESPIRATION

Under storage conditions grains generate heat as a result of respiration and mold activity. From the engineering point of view, determination of the amount of heat generated allows for the proper design of grain aeration systems to maintain crop quality. In extreme cases rain, snow, humid air from a roof leak or ventilation opening cause wet pockets in a grain bin. Heat generated within the pocket of wet grain provides a favorable environment for growth of microorganisms and excessive heat generation.

No theory adequately explains the heat of respiration in stored grain, although the respiration equation has often been used to estimate heat production [67,68]:



For each gram of dry matter ($C_6H_{12}O_6$), which is oxidized, 15.7 kJ of heat is produced. Equation 24.33 provides a simple way of calculating the heat production from the CO_2 production. According to Equation 24.33 the respiratory quotient (ratio of oxygen consumption to carbon dioxide production) is equal to 1.0. However, measured quotients frequently deviate from 1.0 [69].

Heat production under the adiabatic conditions similar to those in naturally occurring pockets of wet grain is measured in a specially designed calorimeter [70] (Figure 24.8). A grain sample is split and placed in two identical 1-L Dewar flasks, which are housed in an insulated chamber. One of the flasks, from which grain is sampled regularly for determination of the moisture content and microfloral infection, is used as a reference flask. To avoid disturbing the test flask during grain sampling, the temperatures inside and outside the test flask are continuously monitored by six thermocouples, three inside and three outside. When grain in the test flask starts to heat, due to metabolic activity, the inside temperature rises, causing a difference between the inside and outside temperatures. If the difference exceeds the preset value of 0.5–0.7°C, the electric heater is turned on to heat the air in the chamber, and thus equalizing the outside and the inside temperatures. The temperature gradient across the flask wall is minimized

TABLE 24.14
Derived Coefficients a and b in Equation 24.32 for Some Major Crops

Seed	Coefficients		Moisture Content Range (db)
	a	b	
Barley	1.0	−19.9	$0.09 \leq M \leq 0.22$
Beans	0.5	−16.0	$0.09 \leq M \leq 0.28$
Corn	2.1	−17.0	$0.10 \leq M \leq 0.24$
Peanut kernel	1.5	−29.2	$0.06 \leq M \leq 0.13$
Rice	3.2	−21.7	$0.10 \leq M \leq 0.14$
Sorghum	1.2	−19.6	$0.10 \leq M \leq 0.24$
Soybeans	0.4	−13.9	$0.06 \leq M \leq 0.26$
Wheat, durum	0.8	−18.1	$0.10 \leq M \leq 0.26$
Wheat, hard	1.7	−17.6	$0.10 \leq M \leq 0.26$
Wheat, soft	3.9	−23.6	$0.10 \leq M \leq 0.20$
Wheat (Gallaher [65])	23.0	−40.0	$0.10 \leq M \leq 0.18$

Source: From Cenkowski, S., Jayas, D.S., and Hao, D., *Can. Agric. Eng.*, 34, 281, 1992.

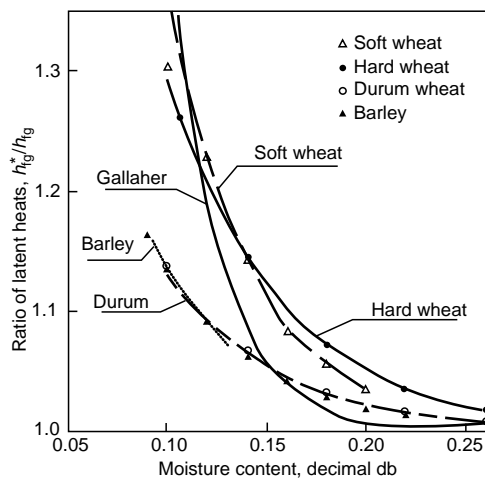


FIGURE 24.7 Effect of moisture content on the latent heat of vaporization of selected grains compared to Gallaher's model. (From Gallaher, G.L., *Agric. Eng.*, 32, 34, 1951; Cenkowski, S., Jayas, D.S., and Hao, D., *Can. Agric. Eng.*, 34, 281, 1992. With permission.)

throughout the course of heating; consequently, no heat is transferred from or to the grain in the flask. Under the adiabatic condition, the metabolic heat produced inside the flask by the seed and microflora is measured directly.

During the initial 11 d, the heat production rate in wheat, calculated from measured CO_2 production, followed the directly measured rate of heat production (Figure 24.9). On the average, the directly measured rates were 27 and 14% higher than those calculated from CO_2 production for 23.0 and 27.2% moisture contents, respectively.

The respiratory quotient is not constant during adiabatic heating (Figure 24.10). For wheat at 27.2% moisture, it increased from 1 to 4.2 in 4 d, and then decreased gradually. The peak respiratory quotient was lower for 23.0% moisture content wheat. At both moisture contents, the respiratory quotient stayed at about 0.8 after the temperature of the grain reached 52°C.

24.5 ELECTRICAL PROPERTIES

24.5.1 DIELECTRIC PROPERTIES

A biological material (a dielectric) can be exposed to electric or magnetic fields in frequencies from direct current (0 Hz) to x-rays ($\approx 10^{18}$ Hz). From direct current to ac frequencies up to about 8 Hz, lumped circuits (composed of serial and parallel resistances and capacitance elements) are used to measure

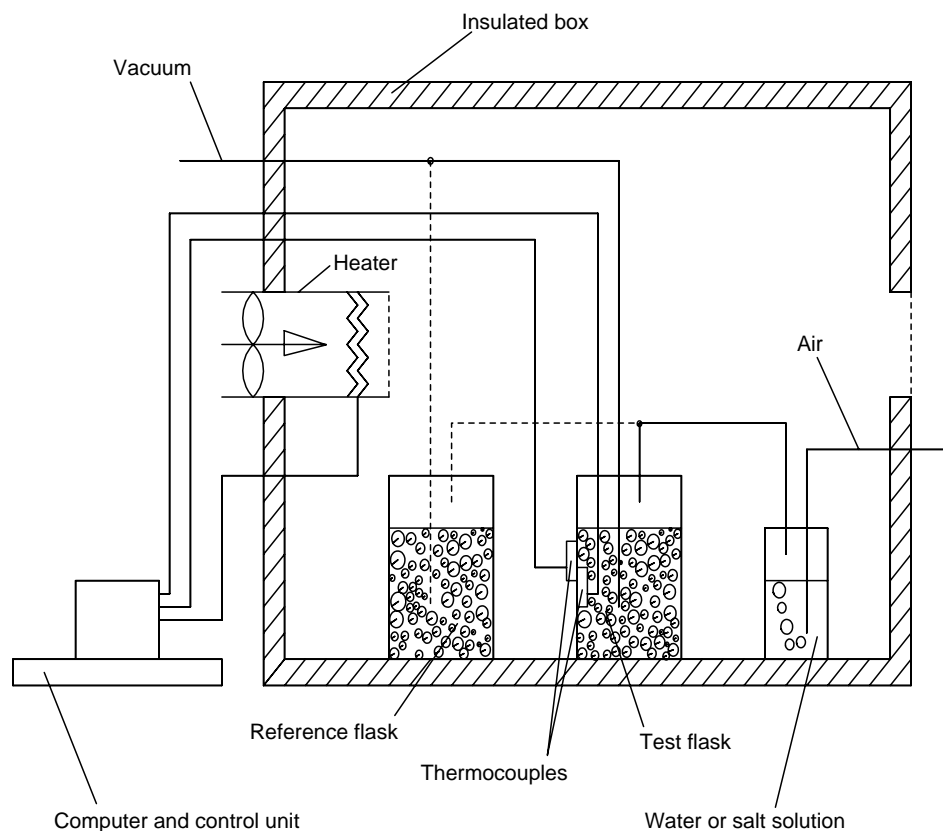


FIGURE 24.8 Computer controlled calorimeter. (From Zhang, Q., Muir, W.E., Sinha, R.N., and Cenkowski, S., *Can. Agric. Eng.*, 34, 233, 1992. With permission.)

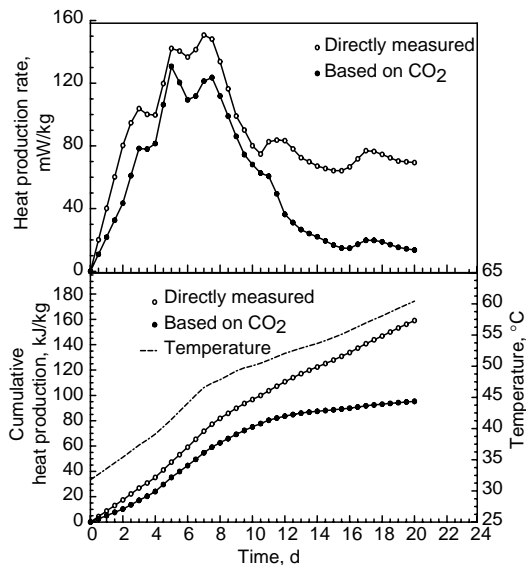


FIGURE 24.9 Means of cumulative heat production, heat production rate, and temperature of wheat at 27.2% wet basis moisture content. (From Zhang, Q., Muir, W.E., Sinha, R.N., and Cenkowski, S., *Can. Agric. Eng.*, 34, 233, 1992. With permission.)

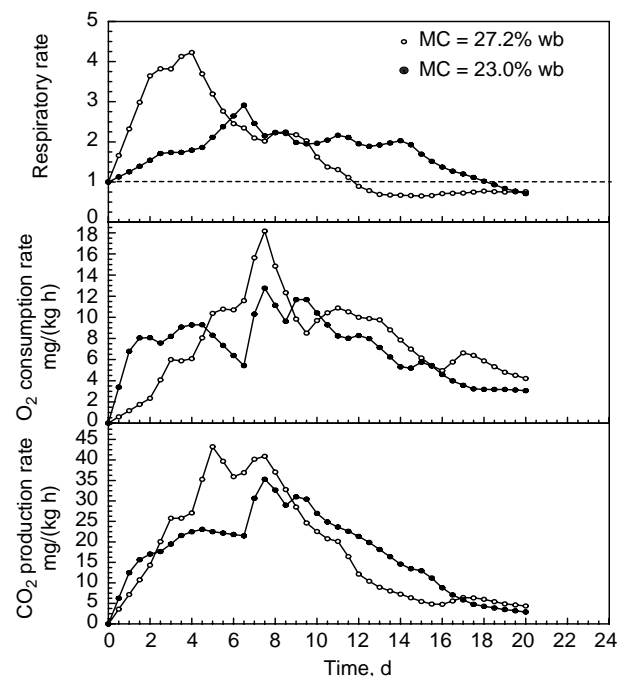


FIGURE 24.10 Comparison of CO_2 production and O_2 consumption between wheat of different moisture content. (From Zhang, Q., Muir, W.E., Sinha, R.N., and Cenkowski, S., *Can. Agric. Eng.*, 34, 233, 1992. With permission.)

dielectric properties (Figure 24.11). The properties of the circuit element may be measured in the range up to 10^7 Hz by bridge arrangements and at the higher frequencies (between 10^4 and 10^8 Hz) in resonant circuits [71].

For wavelengths between meters and millimeters, the dimensions of the dielectric (crop) become comparable to those of the waves, and the physical distinction between coil and condenser begins to disappear. The tested material is inserted into a waveguide and its characteristics can be determined by standing wave patterns. The wavelength range is known as the interference optics. As the wavelength shrinks, the dielectric properties of a material can be determined from infrared to ultraviolet by reflection and transmission measurements (range of geometrical optics).

Biological materials have the ability to store and dissipate electrical energy from an applied electromagnetic field. The properties resulting from electrical charging and loss currents generally related to the material's electrical capacitance and resistance are the fundamental dielectric properties. These properties have been of interest because of their influence on energy absorption in dielectric heating. The dielectric properties that are of interest include the dielectric constant, ϵ' , the dielectric loss factor, ϵ'' , which are respectively the real and imaginary parts of the complex relative permittivity, $\epsilon^* = \epsilon' - j\epsilon''$, the loss tangent or dissipation factor, $\tan(\delta) = \epsilon''/\epsilon'$, and the ac conductivity, $\sigma = \omega\epsilon_0 \epsilon''$ [72].

The dielectric constant may be measured by a Q -Meter within the frequency range in which it operates and with a suitable sample holder [73]. The Q -Meter is one of the common instruments in radio frequency measurement for determining the inductance, the capacitance, the Q of electronic components, and the resonant frequencies of circuits. The Q -Meter includes a variable frequency oscillator with the output coupled to a series resonant circuit (Figure 24.12) [73].

Electrical properties of grain have been utilized for quick moisture tests based on the measurement of resistance, capacitance, or electrical conductivity. Many studies have been devoted to the development of density-independent functions of the dielectric properties that would permit on-line measurement of moisture content [74–76]. Also, measurements of electrical properties of grain and seed have been employed for purposes other than determining moisture content. For example, viable seeds of corn were sorted from dead seeds by measuring the current conducted by individual soaked kernels between electrodes connected to a 6-V dc source [77]. Another application of electrical properties is electrostatic separation where the ability of a seed to hold a surface charge is determined mainly by its conductivity.

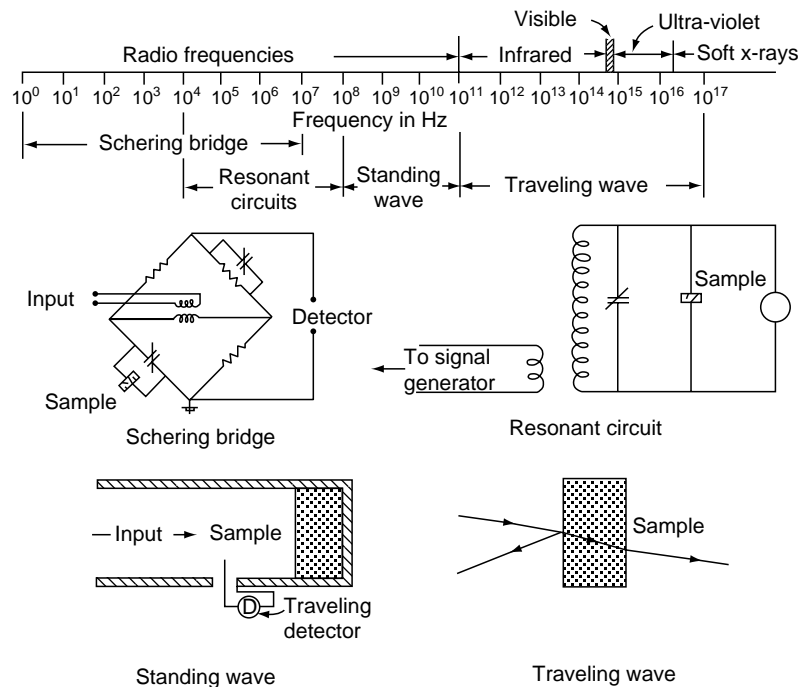


FIGURE 24.11 Frequency ranges and measuring techniques. (From Hippel, A.R., *Dielectrics and Waves*, John Wiley & Sons, New York, 1954. With permission.)

24.5.2 CAPACITANCE- AND RESISTANCE-BASED MEASUREMENTS

For sensing moisture content, a parallel-plate capacitor may be used [78–80]. Capacitance measurements are taken at frequencies of 1.0–4.5 MHz with a Hewlett-Packard 4192 LF Impedance Analyzer equipped with the 16096A test fixture and a specially constructed electrode assembly. Variations in kernel shape and thickness cause capacitance variations. For these reasons, the thickness, mass, and projected area of individual kernels must also be measured [80].

Using a combination of a dc resistance measuring meter (highly sensitive ohmmeter with a specially designed electrode) and a capacitance-type meter, a measure of moisture distribution in corn kernels can be obtained. This information has been used to detect heat damage in artificially dried corn [81,82].

24.5.3 MEASUREMENTS BASED ON ELECTRICAL CONDUCTIVITY

A moisture meter measuring the dc conductance in a circuit was developed from the Wheatstone bridge [83]. Because the electrical properties of whole grain are preferentially affected by the moisture content of the surface layers, the kernel is ground before the test. In a test cell, the meal is compressed against

two coplanar electrodes to a standardized pressure of 6.9 MPa. The electric current penetrates the sample only to a depth of the order of the electrode separation, so that the quantity of meal in the cell is unimportant above a certain minimum.

The rate of water penetration in sorghum kernels during tempering has been studied using a Tag-Heppenstall (C.T. Taniabue, Brooklyn, NY) electrical conductance moisture meter [84]. The tester consists of a pair of corrugated rolls. When grain passes between the rolls, it is slightly crushed and thus acts as a path for the electric current. From the current, I , that has passed through the grain, the resistance, R , is determined by Ohm's law ($I = E/R$)

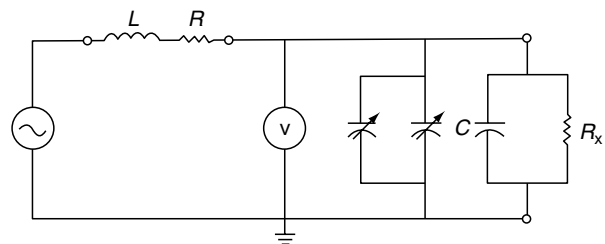


FIGURE 24.12 Simplified diagram for Q-Meter series resonant measuring circuit with sample holder connected. (From Nelson, S.O., *Trans. ASAE*, 22, 950, 1979. With permission.)

where E is the voltage used. The resistance is inversely proportional to the moisture content. By measuring the decrease in the apparent moisture in the outer layers of grain at regular time intervals, the rate of the water penetration into the kernel is determined.

A conductance-type meter similar to the Tag-Heppenstall device has been designed to test single corn kernels [85]. The rolls are about one third the size of those of the Tag-Heppenstall; i.e., approximately 50 mm in diameter and 30 mm long, and the idler roll is attached by a spring that gives the rolls freedom to adjust to different size kernels. The rolls are connected to a logarithmic amplifier that converts the natural logarithmic response into a more useful line response. The schematic of the instrument is shown in Figure 24.13. The 100 V passing through the rolls is reduced by a series of resistors so that the signal reaching the recorder is 100 mV, and when the rolls are shorted, the peak ammeter response is 100 units. Tests showed that the response reached 100 units for kernels with about 25% moisture.

The problems associated with blending grain shipments that vary widely in moisture content are well recognized. They include hazards of mold infection and increased breakage susceptibility. A simple and rapid method for detecting blends of corn varying widely in moisture content has been reported in the literature [86]. The detection is based on dual determination of moisture, i.e., by an oven drying method and electrical conductance meter (Tag-Heppenstall, model 8004, type 14, Weston Electrical Instruments Corp. Newark, New Jersey) or by two electrical methods (conductance—Tag-Heppenstall method versus capacitance—Motomco moisture meter model 914, Motomco Inc., Patterson, New Jersey). In both

methods, the detection of blends of corn is based on the difference in moisture content resulting from the two methods.

24.6 OPTICAL PROPERTIES

Information on optical properties can be used for automatic sorting, quantifying foreign materials [87], and for objective classification of grain class and variety [88]. The use of light transmittance or reflectance offers a means for determining internal quality of grains, moisture content [89], color, appearance [90], the extent of smut spores on wheat [91]. Infrared and far-red light has been used to change lettuce-seed germination characteristics [92].

Energy can be transmitted from a body by electromagnetic waves. These vary from very short to very long waves over a wide spectrum from cosmic to long radio and electric waves. The spectrum of visible light is usually divided into eight intervals, corresponding to the following characteristic colors and wavelength bands: violet (380–450 nm), light blue (450–480 nm), blue (480–510 nm), green (510–550 nm), yellow-green (550–575 nm), yellow (575–585 nm), orange (585–620 nm), and red (620–760 nm). The infrared spectrum is broader and falls between 0.7 and 100 μm . When electromagnetic waves strike a body, they may be absorbed, transmitted, or reflected. Absorption, reflection, and transmission may vary with wavelength [93] and with the type of grain. Only 4% of the incident radiation from a beam of light striking a typical grain kernel is reflected [94]. The remaining radiation is transmitted into the object. Of this portion, part is absorbed by the object, part is reflected back to the

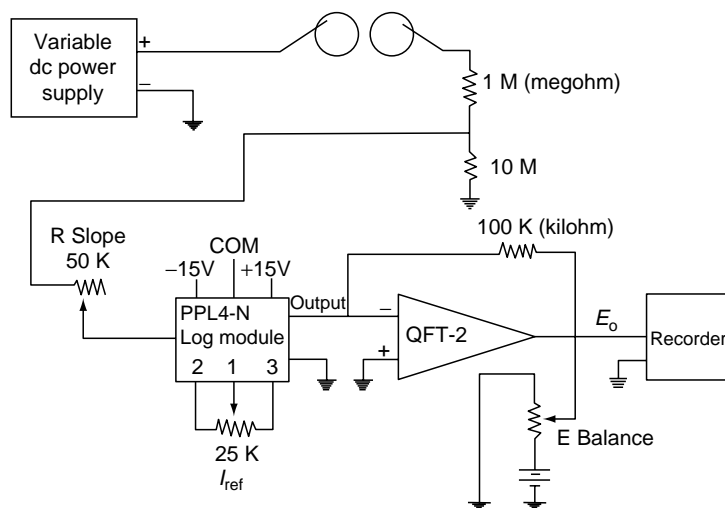


FIGURE 24.13 Electrical schematic of the instrument for determining moisture content of single kernels of corn. The PPL4-N log module and the QFT-2 are manufactured by Teledyne Philbrick. (From Watson, C.A., Greenway, W.T., Davis, G., and McGinty, R.J., *Cereal Chem.*, 56: 137–140, 1979. With permission.)

surface (body reflection), and part is transmitted through the object. The absorbed portion is transformed to other forms of energy. For infrared radiation (0.7–100 μm) the energy is transformed into heat. The amount of transmitted rays absorbed by the material depends on the wavelength and the physical characteristics of the material. Energy passing through a bulk sample of grain is a combination of transmitted and reflected light. In the near-infrared part of the spectrum (700–1400 nm), a kernel of wheat transmits about 2% and reflects 50% of the radiation on the kernel [91]. A layer of steel balls of a size similar to the kernels of wheat and of the same layer thickness reflect a similar amount of energy as kernels of wheat but the amount of transmitted light is 30 times less. This indicates that some energy is transmitted through individual grain kernels.

Absorption spectra are often characterized by the transmittance, τ , at a given wavelength:

$$\tau = \frac{I}{I_0} \quad (24.34)$$

$$\alpha = \log\left(\frac{I_0}{I}\right) = -\log \tau \quad (24.35)$$

The absorbance is related to the thickness, L , of the sample and the concentration, C , of absorbing molecules by the Beer–Lambert law [95]:

$$\alpha = \varepsilon CL \quad (24.36)$$

According to the law of conservation of energy, the radiation incident on an object, I_0 , must be equal to the sum of reflected, I_R , absorbed, I_α , and transmitted, I_τ , radiation:

$$I_0 = I_R + I_\alpha + I_\tau \quad (24.37)$$

Specular reflection occurs when the angle of reflection equals the angle of incidence, whereas diffuse reflection is spread out evenly over a full hemisphere. The main difficulty in measuring the spectral transmittance properties of a kernel is the problem of collecting sufficient energy from the transmitted signal. To collect the maximum amount of light under such conditions the phototube should surround the entire sample. The nearest approach to this condition can be attained by enclosing the sample in a light-integrating sphere with a phototube viewing a small port in the sphere to measure the brightness of the sphere surface. Nearly, all transmitted light can be

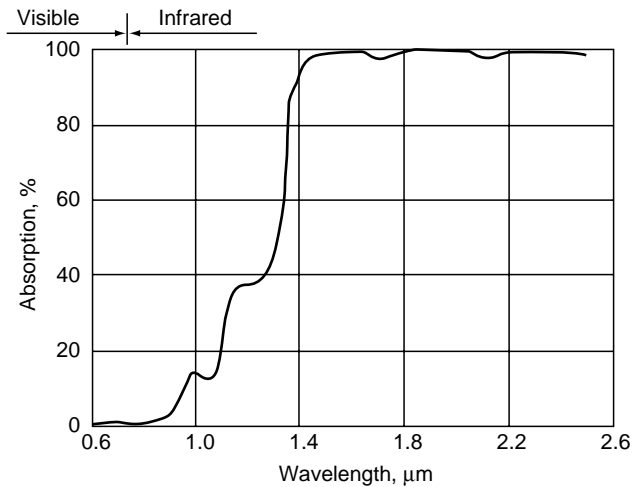


FIGURE 24.14 Absorption of infrared radiation by a 3 mm layer of water. (From Shuman, A.C. and Staley, C.H., *Food Technol.*, 4, 481, 1950. With permission.)

collected by this arrangement regardless of where it emerges from the grain [96].

Knowledge of the spectral reflectance of grain is essential to design an infrared grain dryer. The choice of the source of infrared radiation and the thickness of the grain sample both require this knowledge. Figure 24.14 shows the variation in the absorption of infrared radiation for a 3 mm depth of water. Wavelengths longer than 1400 nm are completely absorbed. The interaction between the absorption characteristics of a material and the intensity of the light source can be illustrated in the following example. Consider the drying of a material, which has absorption characteristics as described by the curve in Figure 24.14. It is necessary to select a light source. The relative intensity of the light should be high enough under the highest absorption level (above 1400 nm) for the material. From the three available sources of infrared radiation (Figure 24.15), the desired radiation source characteristic is the one with a filament temperature of 2200 K. In this case 65% of the total energy of the lamp (this is measured by the area under the relative intensity curve and above 1400 nm wavelength) will be completely absorbed when passing through the sample. The light sources with filament temperatures of 3500 or 1100 K are not desirable. In the first case (3500 K) the maximum light intensity falls in the range where absorption of the material is small. In the second case (1100 K) the higher intensity characteristic of the light source is beyond the range of the absorption characteristics of the dried material.

Spectrophotometers are instruments that measure absorption or reflectance, and are not, as the name implies, restricted to the visible light range.

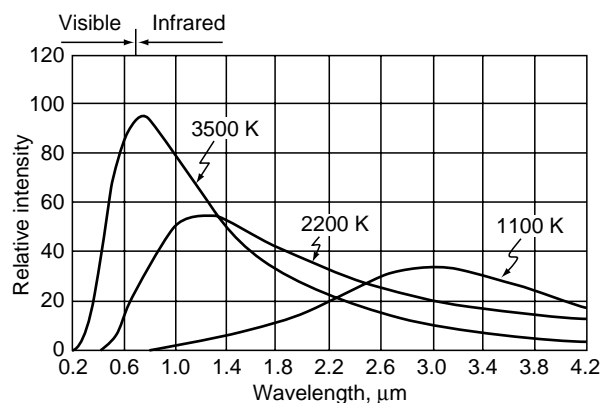


FIGURE 24.15 Relative intensities of various sources of infrared radiation. (From Shuman, A.C. and Staley, C.H., *Food Technol.*, 4, 481, 1950. With permission.)

They include a light source, wavelength selector, and detector. There are many commercial spectrophotometers that are suitable for experiments in the visible and nonvisible ranges. A greater flexibility can be obtained by assembling an instrument from modular components available from such sources as Oriol, Bomem, PTR, Optics, and Spex.

24.7 AERODYNAMIC PROPERTIES

In the development of equipment for harvesting, handling, and cleaning operations, a knowledge of the aerodynamic properties of various grains is necessary. The aerodynamic drag coefficient and the terminal velocity are the two most important aerodynamic properties.

The geometry of a solid body held in a free stream is the major factor determining the drag force, F_D , exerted on a body. The drag force can be calculated from

$$F_D = C_D \frac{\rho_t v^2}{2} A_p \quad (24.38)$$

Values for the drag coefficient have been determined for bodies of regular shape such as spheres, cylinders, and flat disks [97,98]. The drag coefficients are usually shown on a log-log plot, as a function of the Reynolds number.

Most seed kernels are irregular in shape. Their drag coefficients depend not only on the shape but also on the orientation of the kernels in the airstream. Thus, an equivalent diameter is used in the determination of the Reynolds number. Drag coefficients for various crops are given in Table 24.15.

A free-falling body ceases to accelerate after some time and the body attains a constant terminal velocity. The resisting force becomes equal to the mass of the body. The terminal velocity can be calculated theoretically for a spherical body by replacing the drag force, F_D , in Equation 24.38 by its mass. For irregularly

TABLE 24.15
Terminal Velocity of Seed Grains

Seed	Terminal Velocities m/s	Drag Coefficient	Reynolds Number	Ref.
Alfalfa	5.5	0.50	601	120
Barley	7.3–9.0	—	—	101
Barley	7.0	0.50	2280	120
Corn	9.7–11.3	—	—	101
Corn	10.6	0.56	5770	120
Corn	11.4	0.56–0.70	5700	121
Flax	4.7	0.52	836	120
Oats	7.0–8.3	—	—	101
Oats	5.9–6.3	0.47–0.51	1900–2480	120
Oats	6.6	0.47–0.51	2000	121
Soybeans	14.5	0.45	6300	121
Soybeans	11.2–11.9	—	—	101
Soybeans	13.5	0.45	6280	120
Wheat	7.7–9.6	—	—	101
Wheat	9.0	0.50	2720	120

shaped kernels, the terminal velocity is determined experimentally. There are two types of experiments: (i) free-fall tube and (ii) vertical wind tunnel.

In the free-fall tube, a kernel is held by partial vacuum by a vacuum drop head (Figure 24.16); the kernel free-falls after breaking the vacuum [99]. The passage of the kernel through a light beam produces a voltage drop from the photocell which starts a counter. The kernel's impact on the receiver diaphragm, at the bottom of the tube, causes the second voltage drop from the microphone stopping the counter. The height of fall is controlled by changing the length of the drop tube. Since the terminal velocity of seeds ranges from approximately 5 to 15 m/s, the tube always must be of a sufficient length. For example, testing soybeans requires a tube at least 15 m long. Other disadvantages are that: (i) the seeds have a tendency to fall off the center line and (ii) the movement of the free-falling seeds is difficult to observe.

The vertical wind tunnel has approximately a 1.8-m working section with a taper of 3° on each side (Figure 24.17). This divergence causes a reduction of the air velocity by about 25% per 1 m in length and is essential for particle flotation and stability. The taper enables a particle to float to a height corresponding to its terminal velocity. The floating height can be observed and measured. This technique has been used to measure the terminal velocity of three cultivars of wheat kernels [100], corn, soybeans, wheat, oats, and barley [101], and milkweed pods [102]. The advantage of the vertical wind tunnel over the free-falling tube is that the first one does not require as much headspace. The vertical wind tunnel

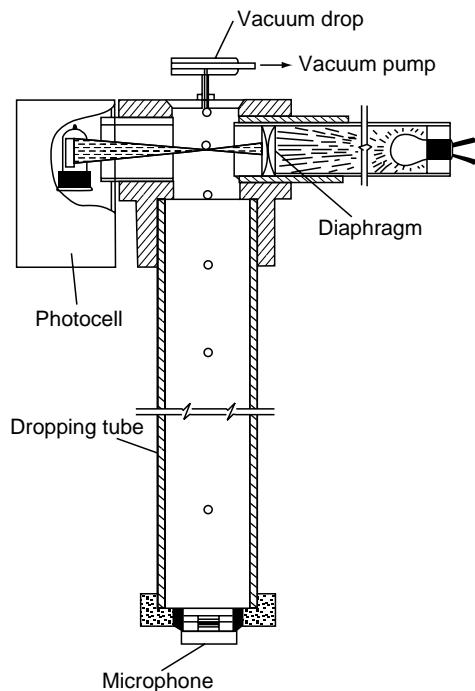


FIGURE 24.16 Free-fall drop equipment. (From Keck, H. and Goss, J.R., *Trans. ASAE*, 8:553–554, 557, 1965. With permission.)

is of a compact structure and spinning of seeds can be quite clearly observed and monitored, however, the slightest nonuniformity in the air velocity causes undesired rotation and side movements of the seeds.

After determining the terminal velocity, the drag coefficient can be calculated using Equation 24.38. The terminal velocity of selected seed grains is summarized in Table 24.15. The indicated range reflects the influence of the geometric characteristics of different varieties and the different-sized kernels within the same variety.

LIST OF SYMBOLS

- A heat transfer area, m^2
- A constant
- A_c area of the smallest circumscribing circle, m^2
- A_p the projected area of a body normal to the flow or surface, m^2
- a the longest dimension of a seed, mm
- a constant
- a_1 coefficient which depends on a material
- B constant
- b constant
- b the longest dimension of a seed normal to a , mm
- b_1 constant
- C constant
- C concentration of absorbing molecules, mole/L

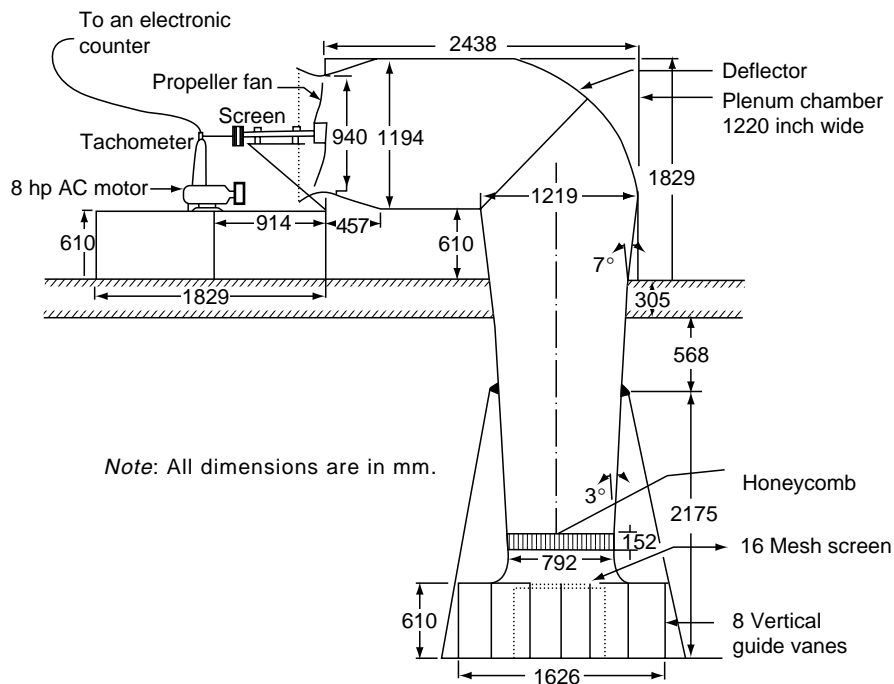


FIGURE 24.17 Diagram of a vertical wind tunnel. (From Bilanski, W.K. and Lal, R., *Trans. ASAE*, 8, 411, 416, 1965. With permission.)

C_D the drag coefficient (dimensionless)
 c the longest dimension of a seed normal to a and b , mm
 c specific heat of grain, J/(kg K)
 c_1 constant
 c_{cp} specific heat of a copper plug, J/(kg K)
 c_d specific heat of dry matter of seeds, J/(kg K)
 c_w mean specific heat of water, J/(kg K)
 D_g geometric mean diameter, m
 d_e equivalent diameter, m
 E mass of water having an equivalent thermal capacity to that of the calorimeter flask, stirrer, thermocouple, g
 F_D the total drag force, N
 h heat transfer coefficient (surface conductance), W/(m² K)
 h_{cv} volumetric heat transfer coefficient, kJ/(m³s K)
 h_{fg}^* latent heat of vaporization of water in seeds, kJ/kg H₂O
 h_{fg} latent heat of vaporization of free water, kJ/kg H₂O
 I intensity of light transmitted through the sample, decimal
 I_o intensity of light incident on the sample, decimal
 j imaginary number
 K drying constant, 1/s
 k thermal conductivity, W/(m K)
 k_a thermal conductivity of the air, W/(m K)
 k_d thermal conductivity of a layer of dry matter of seeds, W/(m K)
 L sample thickness, m
 M moisture content db, kg H₂O/kg dry matter
 M' moisture content, % wb
 M_e equilibrium moisture content, kg H₂O/kg db
 M_i initial moisture content, kg H₂O/kg db
 m mass of the material, kg
 \dot{m} airflow rate, kg/s.
 \dot{m}_a airflow rate, kg/(m² s)
 m_{cp} mass of a copper plug, kg
 m_s mass of the sample, kg
 m_w mass of water, kg
 N drying coefficient
 n number of particles
 Nu Nusselt number, dimensionless
 Pr Prandtl number, dimensionless
 P_v actual pressure of water vapor, Pa
 P_{vs} saturation pressure of water vapor, Pa
 q heat flow, or heat flux, W or W/m²
 Q amount of heat, kJ
 q_l heat input per unit length of line heater, W/m
 R roundness (decimal)
 Re Reynolds number, dimensionless
 RH relative humidity, decimal
 r distance of the grain layer from the zero axis which represents the heat source, m
 T drying air temperature, °C

T temperature of a copper plug (Equation 24.12), °C
 T_c initial temperature of water at transient time, °C
 T_m temperature of the mixture extrapolated at transient time, °C
 T_∞ temperature of a copper rod (Equation 24.12), °C
 t time, s
 t_c corrected time, s
 V_e equivalent volume, m³
 V_n net volume of particles, m³
 v drying air velocity, m/s
 v free stream velocity, m/s
 x distance from the face of a slab, m
 z dimensionless number

GREEK SYMBOLS

α absorbance, decimal
 α thermal diffusivity of seeds, m²/s
 ∇ gradient
 δ loss angle of a dielectric, deg
 ε molar absorptivity, L/(mole cm)
 ε_0 permittivity of free space, $\varepsilon_0 = 8.854 \times 10^{-12}$ farad/m.
 ε porosity, decimal
 θ temperature of a grain layer, °C
 θ grain temperature, °C
 θ_0 initial grain temperature, °C
 θ_0 initial temperature of both sample and copper plug, (Equation 24.12) °C
 θ_0 initial temperature of a slab, °C
 θ_c temperature at center of a slab, °C
 θ_s initial temperature of the sample, °C
 ν kinematic viscosity of drying air, m²/s
 ρ density of fluid (air), kg/m³
 ρ_b bulk density, kg/m³
 ρ_t particle density, kg/m³
 σ ac conductivity, 1/(ohm m) or mhos/m
 τ transmittance, decimal
 ω angular frequency of the applied electric field ($\omega = 2\pi f$ where f is the frequency of the applied voltage), rad/s.

SUBSCRIPTS

db dry basis
 R reflected
 wb wet basis
 α absorbed
 τ transmitted

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25 Drying of Fruits and Vegetables

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25.1 INTRODUCTION

From the point of view of consumption, fruits are plant products with aromatic flavor that are naturally sweet or normally sweetened before usage [1]. Apart from providing flavor and variety to human diet, they serve as important and indispensable sources of vitamins and minerals although they are not good or economic sources of protein, fat, and energy. The same is true in the case of vegetables, which also play an important role in human nutrition in supplying certain constituents in which other food materials are deficient and in adding flavor, color, and variety to the diet [2].

After moisture, carbohydrates form the next most abundant nutrient constituent in fruits and vegetables, and are present as low-molecular-weight sugars or high-molecular-weight polymers like starch and so on. The celluloses, hemicelluloses, pectic substances, and lignin characteristic of plant products together form dietary fiber, the value of which in human diet is increasingly realized in recent years, especially for the affluent society of the Western countries. Virtually all human's dietary vitamin C, an important constituent of human diet, is obtained from fruits and vegetables, some of which are rich in provitamin A (β -carotene) (e.g., mango, carrot, etc.). They are important suppliers of calcium, phosphorus, and iron.

Fruits and vegetables have gained commercial importance and their growth on a commercial scale has become an important sector of the agricultural industry. Recent developments in agricultural technology have substantially increased the world production of fruits and vegetables. Consequently a larger proportion of several important commodities is handled, transported, and marketed all over the world than before with concomitant losses calling for suitable postharvest techniques for storage and processing to ensure improved shelf life. Production and consumption of processed fruits and vegetables are also increasing.

25.2 POSTHARVEST TECHNOLOGY OF FRUITS AND VEGETABLES

25.2.1 WORLD PRODUCTION

The present world production of fruit (excluding melons) according to Food and Agricultural Organization (FAO) was about 444.65 million metric tons (mt) in 1999 [3]. China with a production of 59.5 mt (13.4%) is a leading producer of fruits in the world. India, with 38.56 mt (8.7%) occupies second position, followed by Brazil (8.45%), United States (6.4%), and Italy (4.3%).

World production of vegetables (including melons) is about 628.75 mt. The major vegetable producing countries were China, India, United States, Turkey, Italy, Japan, and Spain. China was the largest producer accounting for about 250.0 mt (39.8%) whereas India was the second contributing about 59.4 mt (9.45%).

25.2.2 LOSSES

Most fruits and vegetables contain more than 80% water and are therefore highly perishable. Water loss and decay account for most of their losses, which are estimated to be more than 30–40% in the developing countries in the tropics and subtropics [1] due to inadequate handling, transportation, and storage facilities. Apart from physical and economic losses, serious losses do occur in the availability of essential nutrients, notably vitamins and minerals.

The need to reduce postharvest losses of perishable horticultural commodities is of paramount importance for developing countries to increase their availability, especially in the present context when the constraints on food production (land, water, and energy) are continually increasing. It is being increasingly realized that the production of more and better food alone is not enough and should go hand in hand with suitable postharvest conservation techniques to minimize losses, thereby increasing supplies and availability of nutrients besides giving the economic incentive to produce more [1].

25.2.3 ROLE OF PRESERVATION

One of the prime goals of food processing or preservation is to convert perishable foods such as fruits and vegetables into stabilized products that can be stored for extended periods of time to reduce their postharvest losses. Processing extends the availability of seasonal commodities, retaining their nutritive and esthetic values, and adds variety to the otherwise monotonous diet. It adds convenience to the products. In particular it has expanded the markets of fruit and vegetable products and ready-to-serve convenience foods all over the world, the per capita consumption of which has rapidly increased during the past two to three decades.

Several process technologies have been employed on an industrial scale to preserve fruits and vegetables; the major ones are canning, freezing, and dehydration. Among these, dehydration is especially suited for developing countries with poorly established low-temperature and thermal processing facilities. It offers a highly effective and practical means of preservation to reduce postharvest losses and offset the shortages in supply.

25.2.4 PRESERVATION BY DRYING

The technique of dehydration is probably the oldest method of food preservation practiced by humankind. The removal of moisture prevents the growth and reproduction of microorganisms causing decay and minimizes many of the moisture-mediated deteriorative reactions. It brings about substantial reduction in weight and volume, minimizing packing, storage, and transportation costs and enables storability of the product under ambient temperatures. These features are especially important for developing countries and in military feeding and space food formulations.

A sharp rise in energy costs has promoted a dramatic upsurge in interest in drying worldwide over the last decade. Advances in techniques and development of novel drying methods have been made available for a wide range of dehydrated products, especially instantly reconstitutable ingredients, from fruits and vegetables with properties that could not have been foreseen some years ago. The growth of fast foods has fueled the need for such ingredients. Due to changing lifestyles, especially in the developed world, there is now a great demand for a wide variety of dried products with emphasis on high quality and freshness besides convenience.

This chapter is intended to provide a comprehensive account of the various drying techniques and appliances developed and applied over the years specifically for the dehydration of fruits, vegetables, and their products. Theoretical and practical aspects of drying as applied to foodstuffs in general have been covered by Sokhansanj and Jayas in the earlier edition of the *Handbook of Industrial Drying* [4]. Therefore, discussion will be restricted to fruit and vegetable drying besides quality changes during drying and storage as specifically applied to these commodities.

25.3 PRETREATMENTS FOR DRYING

Fruits and vegetables are subjected to certain pretreatments with a view to improve drying characteristics and minimize adverse changes during drying and subsequent storage of the products. These include alkaline dips for fruits and sulfiting and blanching for fruits and vegetables [5].

25.3.1 ALKALINE DIP

The alkaline dip involves immersion of the product in an alkaline solution before drying and is used primarily for fruits that are dried whole, especially prunes and grapes. A sodium carbonate or lye solution (0.5% or less) is usually used at a temperature ranging from

93.3 to 100°C [1]. It facilitates drying by forming fine cracks in the skin. Oleate esters constitute the active ingredients of commercial dip solutions used for grapes. They accelerate moisture loss by causing the wax platelets on the grape skin to dissociate, thus facilitating water diffusion.

25.3.2 SULFITING

Sulfur dioxide treatments are widely used in fruit and vegetable drying as sulfur dioxide is by far the most effective additive to avoid nonenzymatic browning [NEB] [6]. It also inhibits various enzyme-catalyzed reactions, notably enzymic browning, and acts as an antioxidant in preventing loss of ascorbic acid and protecting lipids, essential oils, and carotenoids against oxidative deterioration during processing and storage. It also helps in inhibition and control of microorganisms, especially microbial fermentation of sugars in fruits such as sun-dried apricots as encountered during prolonged drying. It has the advantage of allowing higher temperatures, hence shorter drying times, to be used. It is intended to maintain color, prevent spoilage, and preserve certain nutritive attributes until marketed.

Fruits for dehydration are often treated with gaseous SO₂ from burning sulfur as used in the manufacture of dried apricots, peaches, bananas, raisins, and sultanas. Alternatively, apple slices are generally dipped in solutions of the additive (prepared by dissolving sodium bisulfite or SO₂ in water) and may receive an extra treatment with gaseous SO₂ during drying.

Treatment of vegetables with SO₂ gas is impractical. Sulfite solutions are preferred as the most practical method of controlling absorption. As vegetables are blanched before drying, generally the additive is incorporated at the blanching stage either in the blanch liquor if the vegetable is to be dipped or as a spray in the case of steam blanching.

Sufficient SO₂ must be absorbed by the prepared material to allow for losses that occur during drying and subsequent storage. The various methods of application of SO₂ result in varying levels of uptake, which is a function of SO₂ concentration, length of treatment, and time allowed for draining, size and geometry of the food, and the pH of the blanch liquor or spray. Drying times in excess of 12 h for fruits and vegetables and of several days as in sun drying of fruits necessitate use of large amounts of SO₂. It has been shown that only 35–45% of the additive initially incorporated is measurable after drying. The subsequent loss of SO₂ from dried products occurring during storage determines the practical shelf life with respect to spoilage through NEB.

Table 25.1 and Table 25.2 show suggested levels for SO₂ in vegetables and fruits, respectively, after the completion of drying [6].

25.3.3 BLANCHING

Blanching consists of a partial cooking, usually in steam or hot water, before dehydration. It is intended to denature enzymes responsible for bringing about undesirable reactions that adversely affect product quality such as enzymic browning and oxidation during processing and storage. The effectiveness of the treatment is judged by the degree of enzyme inactivation. Thus, activity of polyphenoloxidase is followed in fruits, that of catalase in cabbage and of peroxidase in other vegetables. The other beneficial effects produced by blanching include [5] reduced drying time, removal of intercellular air from the tissues, softening of texture, and retention of carotene and ascorbic acid during storage. Commercially both continuous- and batch-type blanchers are employed, involving 2- to 10-min exposure to live steam. Series blanching in hot water is also used, in which the solids content of the water is maintained at an equilibrium level to minimize leaching losses.

In addition to water and steam blanching, use of microwave energy was demonstrated to be a convenient and effective method of blanching [7] and superior in retention of ascorbic acid. The texture of rehydrated, microwave-blanching freeze-dried spinach was firm, chewy, and highly acceptable.

Low-temperature long-time (LTLT) blanching (65–70°C for 15–20 min) was found to improve the

TABLE 25.1
Suggested Sulfur Dioxide Levels in Dried Vegetables

Vegetable	SO ₂ (ppm)
Beans	500
Cabbages	1500–2500
Carrots	500–1000
Celery	500–1000
Peas	300–500
Potato granules	250
Potato slices	200–500
Sweet potatoes (diced)	200–500
Beets	Not necessary
Corn	2000
Peppers ^a	1000–2500
Horseradish	Destroys flavor

^a0.2% antioxidant BHA gives better color retention.

Source: From Dunbar, J., *Food Tech. New Zealand*, 21(2), 11, 1986. With permission.

TABLE 25.2
Suggested Sulfur Dioxide Levels in Dried Fruits

Fruit	SO ₂ (ppm)
Apples	1000–2000
Apricots	2000–4000
Peaches	2000–4000
Pears	1000–2000
Raisins	1000–1500

Source: From Dunbar, J., *Food Tech. New Zealand*, 21(2), 11, 1986. With permission.

quality (texture) of dried carrot (together with calcium treatment) [8] and dried sweet potato [9] as compared to high-temperature short-time (HTST) blanching (95–100°C for 3 min). Because at this temperature pectin methyl esterase was active to deesterify and increase the free carboxyl group of pectin, which could then form salt bridges with divalent cations to produce a firmer textured product [8].

The prevalence of water blanchers in the industry necessitates the comparison of different types of blanching for their energy utilization. On the basis of a theoretical requirement of 134 kg of steam per 10³ kg of raw vegetables, energy efficiency of a steam blancher was estimated at 5%, a hydrostatic steam blancher at 27%, an IQB unit at 85%, and a water blancher at 60% [10].

25.4 DRYING TECHNIQUES AND EQUIPMENT

25.4.1 DEHYDRATION

Dehydration involves the application of heat to vaporize moisture and some means of removing water vapor after its separation from the fruit and vegetable tissue. Hence it is a combined and simultaneous heat and mass transfer operation for which energy must be supplied.

Several types of dryers and drying methods, each better suited for a particular situation, are commercially used to remove moisture from a wide variety of fruits and vegetables [11]. Whereas sun drying of fruit crops is still practiced for certain fruits such as prunes, grapes, and dates, atmospheric dehydration processes are used for apples, prunes, and several vegetables. Continuous processes, such as tunnel, belt-trough, and fluidized bed (FB), are mainly used for vegetables. Spray drying is suitable for fruit juice concentrates and vacuum dehydration processes are useful for low-moisture, high-sugar fruits.

Factors on which the selection of a particular dryer or drying method depends include form of raw material and its properties, desired physical form and characteristics of the product, necessary operating conditions, and operation costs.

Three basic types of drying processes may be recognized as applied to fruits and vegetables: sun drying and solar drying; atmospheric drying including batch (kiln, tower, and cabinet dryers) and continuous (tunnels, belt, belt-trough, fluidized bed, explosion puff, foam mat, spray, drum, and microwave heated) processes; and subatmospheric dehydration (vacuum shelf belt/drum and freeze dryers). Recently the scope has been expanded to include use of low-temperature and low-energy processes like osmotic dehydration.

In the following sections only a few types of dryers and drying techniques of importance to fruit and vegetable drying are briefly discussed. Detailed information on their design, operation, and economics may be obtained from references quoted in the relevant sections.

25.4.2 SOLAR DRYING

One of the oldest uses of solar energy since the dawn of civilization has been the drying and preservation of agricultural surpluses. It was also the cheapest means of preservation by which water activity was brought to a low level so that spoilage would not take place. It has been used mainly for drying of fruits such as grapes, prunes, dates, and figs.

There is no accurate estimate of the vast amount of material dried using this traditional technique. Since the method was simple and originated and utilized in most of the developing countries, its acceptance created no problem. But there were many technical problems associated with the traditional way of drying in the direct sun. These problems include rain and cloudiness; contamination from dust and by insects, birds, and animals; lack of control over drying conditions; and possibility of chemical, enzymic, and microbiological spoilage due to long drying times. The recent increase in the cost of fossil fuels associated with depletion of the reserve and scarcity has led to renewed interest in solar drying.

Bolin and Salunkhe [12] have exhaustively reviewed the drying methods using solar energy alone and with an auxiliary energy source, besides discussing the quality (nutrient) retention and economic aspects. They suggested that to produce high-quality products with economic feasibility, the drying should be fast. Drying time can be shortened by two main procedures: by raising the product temperature to that moisture can be readily vaporized, whereas at the same time the humid air is constantly removed, and by treating the

product to be dried so that moisture barriers such as dense hydrophobic skin layers or long water migration paths will be minimized. Developments in solar drying of fruits and vegetables up to 1990 have been reviewed by Jayaraman and Das Gupta [13].

To design a solar dryer for drying fruits and vegetables, two important stages are to be considered: to heat the air by the radiant energy from sun and to bring this heated air in contact with the material inside a chamber to evaporate moisture.

Solar dryers are generally classified [14] according to their heating modes or the manner in which the heat derived from solar radiation is utilized. These classes include sun or natural dryers, direct solar dryers, indirect solar dryers, hybrid systems, and mixed systems.

25.4.2.1 Sun or Natural Dryers

Solar or natural dryers make use of the action of solar radiation, ambient air temperature, and relative humidity and wind speed to achieve the drying process.

25.4.2.2 Solar Dryers—Direct

In direct solar dryers the material to be dried is placed in an enclosure with a transparent cover or side panels. Heat is generated by absorption of solar radiation on the product itself as well as on the internal surfaces of the drying chamber. This heat evaporates the moisture from the drying product. In addition it serves to heat and expand the air, causing the removal of the moisture by the circulation of air.

25.4.2.3 Solar Dryers—Indirect

In indirect solar dryers, solar radiation is not directly incident on the material to be dried. Air is heated in a solar collector and then ducted to the drying chamber to dehydrate the product. Generally flat-plate solar collectors are used for heating the air for low and moderate temperature use. Efficiency of these collectors depends on the design and operating conditions. The main factors that affect collector efficiency are heater configuration, airflow rate, spectral properties of the absorber, air barriers, heat transfer coefficient between absorber and air, insulation, and insolation. By optimizing these factors, a high efficiency can be obtained. More sophisticated designs of flat-plate collectors are now available. Imre [15] described such collectors and their efficiency.

25.4.2.4 Hybrid Systems

Hybrid systems are dryers in which another form of energy, such as fuel or electricity, is used to supplement solar energy for heating and ventilation.

25.4.2.5 Mixed Systems

Mixed systems include dryers in which both direct and indirect models of heating have been utilized (Figure 25.1). Several experimental methods were evaluated for the solar dehydration of fruits (apricots): (a) wooden trays; (b) solar troughs of various materials designed to reflect radiant energy onto drying trays; (c) natural convection, solar-heated cabinet dryers with slanted plate heat collectors; (d) dryers incorporating inflated polyethylene (PE) tubes as solar collectors; and (e) PE semicylinders either incorporating a fan blower to be used in inflated hemispheres or incorporating a similar dome used as a solar collector, the air from which is blown over fruit in a cabinet dryer [16]. Method (d) was found to be cheap, 38% faster than sun drying, and could be used as a supplementary heat source for conventional dehydrators.

Solar drying incorporating a desiccant bed for heat storage has been used to dry fruits and vegetables [17]. Hot air up to 27°C above ambient was obtained in a single glass-covered collector with an airflow of about 140 kg/h and raised to 52°C for airflow of 25 kg/h. In the absorbent circuit, which used a double glass-covered collector, temperature

differences were 10% higher. Other forms of heat storage involving use of natural materials such as water, pebbles or rocks, and the like, and salt solutions or absorbents have also been used.

Design and construction of a dryer was described [18] to utilize solar energy in the two-step osmovac dehydration of papaya consisting of a 56-by-25-by-25-cm plexiglass (3.8-cm thick) and a portable condenser vacuum unit (Figure 25.2). Solar osmotic drying had higher drying rates and sucrose uptake than in the nonsolar runs. Similarly, drying rates from solar vacuum drying were about twice those of nonsolar vacuum drying.

Solar drying of a number of vegetables using a solar cabinet dryer fitted with three flat-plate collectors was described [19]. It was concluded that use of three flat-plate collectors instead of one improved the performance of solar cabinet dryer by increasing cabinet temperature and air circulation as compared to drying using single flat-plate collector.

Since solar drying of fruits and vegetables is usually long because of large amount of water to be removed, Grabowski and Mujumdar [20] examined the possibility of coupling osmotic drying with solar drying for more effective drying. They have also

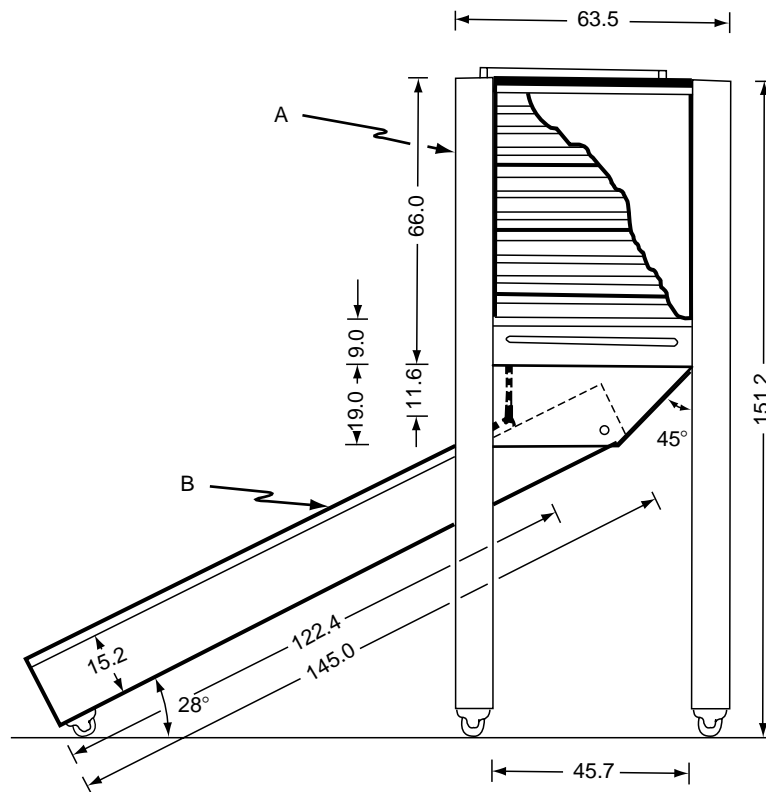


FIGURE 25.1 Dimensions of a combined-mode solar layer (A, dryer; B, solar collector). (From Bolin, H.R. and Salunkhe, D.K., *Crit. Rev. Food Sci. Nutri.*, 16, 327, 1982. With permission.)

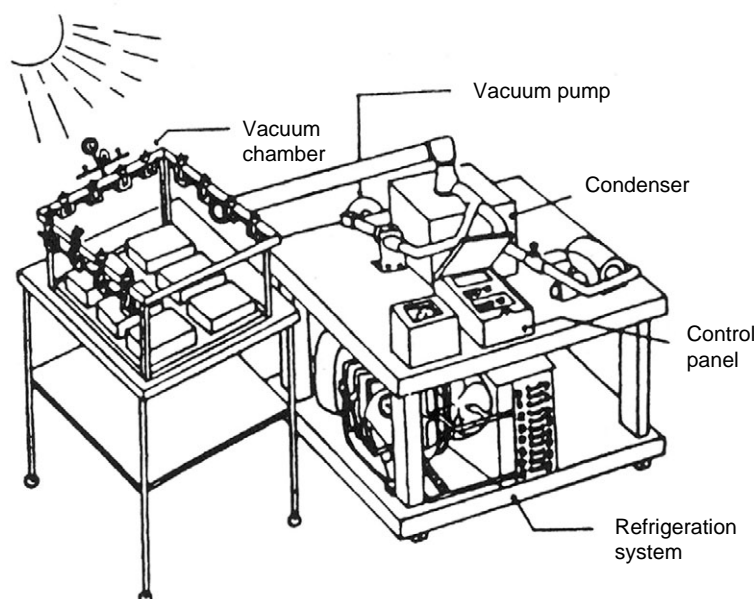


FIGURE 25.2 Components built for solar vacuum drying. (From Moy, J.H. and Kuo, M.J.L., *J. Food Process Eng.*, 8(1), 23, 1985. With permission.)

illustrated applications of solar-assisted osmotic dehydration systems for different production scales. It was observed that minimum twofold increase in the throughput of typical solar dryers was possible while enhancing the nutritional and organoleptic qualities.

A solar drying system consisting of eight flat-plate solar collectors was designed and constructed [21]. Each flat-plate solar collector had a gross area of 2.0 m², effective area of 1.86 m², and average heat-generating capacity of 18.6 MJ/d (at 50% efficiency). A dehydrator of 250 kg capacity was constructed for a fruit or a vegetable (apricot, grapes, persimmon, onion, chillies, etc.). Economic analysis showed that solar drying system is very economic for dehydration of fruit and vegetable.

For commercial success a solar dryer should be economically feasible. But, in general, solar energy systems are capital-intensive. In these dryers, although operating costs are low, large investments have to be made on equipment. The prime economic problem is to balance the annual cost of extra investment against fuel savings. Therefore solar drying could be economical only if the equipment cost is decreased or in the event of fuel cost escalation.

25.4.3 HOT AIR DRYING

Currently most of the dehydrated fruits and vegetables are produced by the technique of hot air drying, which is the simplest and most economical among the various

methods. Different types of dryers have been designed, made, and commercially used based on this technique.

In this method, heated air is brought into contact with the wet material to be dried to facilitate heat and mass transfer; convection is mainly involved. Two important aspects of mass transfer are the transfer of water to the surface of the material that is dried and the removal of water vapor from the surface. The basic concepts, various methods of drying, and different types of hot air dryers are discussed by various authors in review articles and books [1,2,5,22–24].

To achieve dehydrated products of high quality at a reasonable cost, dehydration must occur fairly rapidly. Four main factors affect the rate and total time of drying [23]: physical properties of the foodstuff, especially particle size and geometry; its geometrical arrangement in relation to air (crossflow, through-flow, tray load, etc.); physical properties of air (temperature, humidity, velocity); and design characteristics of the drying equipment (crossflow, through-flow, cocurrent, countercurrent, agitated bed, pneumatic, etc.). The choice of the drying method for a food product is determined by desired quality attributes, raw material, and economy.

The dryers generally used for the drying of piece-form fruits and vegetables are cabinet, kiln, tunnel, belt-trough, bin, pneumatic, and conveyor dryers. Among these, the cabinet, kiln, and bin dryers are batch operated, the belt-trough dryer is continuous, and the tunnel dryer is semicontinuous.

25.4.3.1 Cabinet Dryers

Cabinet dryers are small-scale dryers used in the laboratory and pilot plants for the experimental drying of fruits and vegetables. They consist of an insulated chamber with trays located one above the other on which the material is loaded and a fan that forces air through heaters and then through the material by crossflow or through-flow.

25.4.3.2 Tunnel Dryers

Tunnel dryers are basically a group of truck and tray dryers widely used due to their flexibility for the large-scale commercial drying of various types of fruits and vegetables. In these dryer trays of wet material, stacked on trolleys, are introduced at one end of a tunnel (a long cabinet) and when dry they are discharged from the other end. The drying characteristic of these dryers depends on the movement of airflow relative to the movement of trucks, which may move parallel to each other either concurrently or countercurrently, each resulting in its own drying pattern and product properties.

25.4.3.3 Belt-Trough Dryers

Belt-trough dryers are agitated bed, through-flow dryers used for the drying of cut vegetables of small dimensions. They consist of metal (wire) mesh belts supported on two horizontal rolls; a blast of hot air is forced through the bed of material on the mesh. The belts are arranged in such a way to form an inclined trough so that the product travels in a spiral path and partial fluidization is caused by an upward blast of air.

25.4.3.4 Pneumatic Conveyor Dryers

Pneumatic conveyor dryers are generally used for the finish drying of powders or granulated materials and are extensively used in the making of potato granules. The feed material is introduced into a fast-moving stream of heated air and conveyed through ducting (horizontal or vertical) of sufficient length to bring about desired drying. The dried product is separated from the exhaust air by a cyclone or filter. Jayaraman et al. [25] described a pneumatic dryer in which an initial high temperature (160–180°C for 8 min) drying of piece-form vegetables was done up to 50% moisture, resulting in expansion and porosity in the products that hastened finish drying in a conventional cabinet dryer besides significantly reducing rehydration times and increasing rehydration coefficients of the products (Table 25.3) [25].

25.4.4 FLUIDIZED BED DRYING

The fluidized bed type of dryer was originally used for the finish drying of potato granules. In FB drying, hot air is forced through a bed of food particles at a sufficiently high velocity to overcome the gravitational forces on the product and maintain the particles in a suspended (fluidized) state [22]. Fluidizing is a very effective way of maximizing the surface area of drying within a small total space. Air velocities required for this will vary with the product and more specifically with the particle size and density. A major limitation is the limited range of particle size (diameter usually 20 μm –10 mm) that can be effectively fluidized. The bed remains uniform and behaves as a fluid when the so-called Froude number is below unity.

TABLE 25.3
Process Conditions for High-Temperature, Short-Time Pneumatic Drying of Vegetables and Rehydration Characteristics of Products

Material	Moisture Content (%)				Optimum HTST Drying		Rehydration Time	Rehydration Coefficient
	Raw	Cooked/Blanched	HTST Treated	Final Dried	Temp. (°C)	Time (min)		
Potatoes	82.2	83.3	59.3	4.1	170	8	5	0.94
Green peas	71.1	72.5	38.3	3.4	160	8	5	1.06
Carrots	89.3	91.0	52.9	4.2	170	8	5	0.50
Yams	76.6	78.3	50.2	3.9	180	8	6	1.01
Sweet potatoes	73.6	78.6	53.8	5.3	170	8	2	1.06
Colocasia	80.2	83.3	54.2	4.9	170	8	2	0.98
Plantains, raw	80.8	83.3	58.8	4.6	170	8	4	0.97

Source: From Jayaraman, K.S., Gopinathan, V.K., Pitchamuthu, P., and Vijayaraghavan, P.K., *J. Food Technol.*, 17(6), 669, 1982. With permission.

The theory and food applications of fluidized bed drying have been discussed in many textbooks and articles [5,22–24,26,27]. Apart from the commercial drying of peas, beans, and diced vegetables, it is also used for drying potato granules, onion flakes, and fruit juice powders. It is often used as a secondary dryer to finish the drying process initiated in other types of dryers. It can be carried out as a batch or continuous process with a number of modifications.

The advantages of fluidized bed drying are high drying intensity, uniform and closely controllable temperature throughout, high thermal efficiency, time duration of the material in the dryer may be chosen arbitrarily, elapsed drying time is usually less than other types of dryers, equipment operation and maintenance is relatively simple, the process can be automated without difficulty, and, compact and small, several processes can be combined in an FB dryer [5].

Heat transfer in FB drying could be improved by increasing gas velocity. But, at higher velocities, the particles are transported out of bed and voidage in the bed increases, reducing the volumetric effectiveness of the equipment. From the viewpoint of good gas-to-solid contact, this is undesirable because most of the gas passes around the layers of particles without effective contact.

Another drawback of conventional fluidized bed drying is that the maximum gas velocity is closely related to the physical characteristics of the food particles such as shape, surface roughness, bulk density, and firmness. The maximum gas velocity controls the amount of heat delivered to the bed, since for foods there is usually a critical maximum gas temperature for processing.

The centrifugal fluidized bed (CFB) was designed [28,29] to overcome the limitations of piece size and heat requirements encountered in a conventional FB dryer by subjecting the food particles during fluidization to a centrifugal force greater than the gravitational force. This had the effect of increasing the apparent density of the particles and allowing smooth, homogenous fluidization. Smooth fluidization could be achieved at any desired gas velocity by varying the centrifugal force. The other advantages provided by CFB include increasing the gas velocity to provide improved heat transfer at moderate gas temperature without the problem of heat damage, and large pressure drops across the grid supporting the bed are not needed to obtain smooth fluidization. It was demonstrated to be effective for extremely high rate drying of high-moisture, low-density, sticky, piece-form foods.

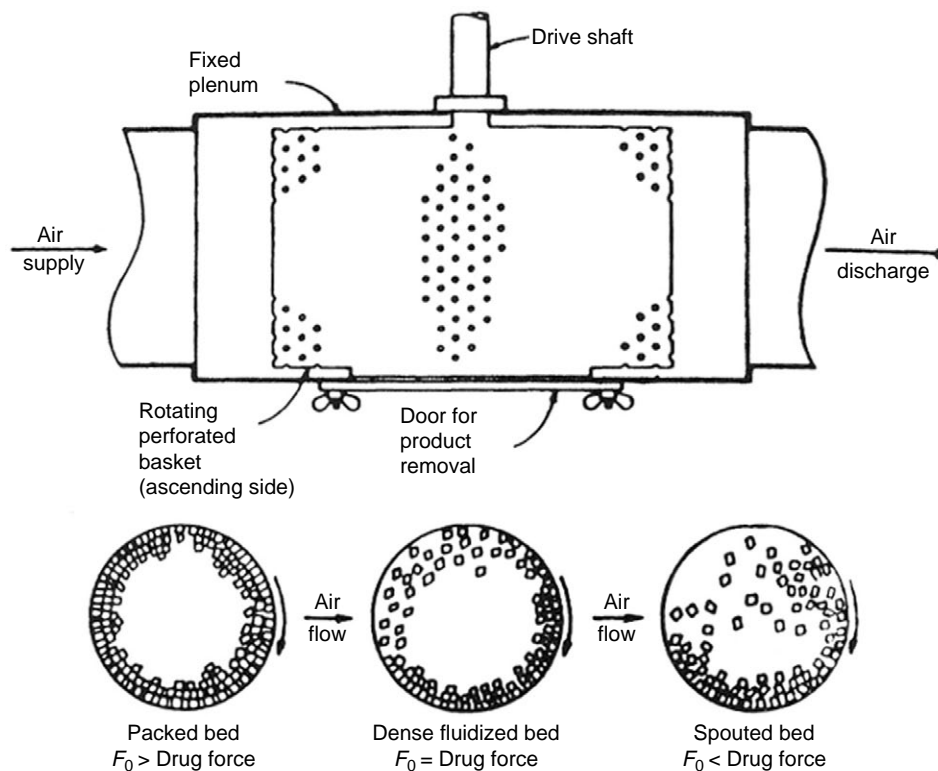


FIGURE 25.3 Modified design of centrifugal fluidized bed dryer allows for lower pressure drops and better heat economy. As the air velocity is increased, the degree of fluidization changes from packed to spouted. (From Brown, G.E., Farkas, D.F., and De Marchena, E.S., *Food Technol.*, 26(12), 23, 1972. With permission.)

A modified centrifugal fluidized bed dryer (CFBD) developed consisted of a cylinder with perforated walls, rotating horizontally about its axis in a high velocity, heated crossflow airstream (Figure 25.3) [29]. Piece-form product to be dried was fed into one end of the rotating cylinder, moved along the cylinder in almost plug-flow manner through the hot air blast, passing crossflow through the perforated walls, and discharged from the other end of the cylinder. On the downstream side (relative to the airflow) within the cylinder, the pieces were held as a fixed bed against the wall by the additive forces of frictional air drag and centrifugation. At high rpm or low air velocity, the centrifugal force on any particle was greater than the drag force of the entering airstream and each particle remained fixed in place. If the air velocity was increased or the rpm decreased, dense-phase fluidization was obtained on the upstream side of the bed because the drag force on the pieces was equal to or slightly greater than the opposing centrifugal force. If the air velocity was further increased, transport of the particles across the cylinder occurred as in a spouted bed. Centrifugal force obtained through cylinder rotational speed to give 3–15Gs allowed the use of air velocities up to 15 m/s or higher, many times greater than can be employed in conventional FBs.

Carrots, potatoes, apples, and green beans dried in this modified CFB at an air velocity of 2400 ft/min

and 240°F showed that a weight reduction of 50% could be achieved in less than 6 min for all items. In comparison with a tunnel dryer with a crossflow air velocity of 780 ft/min, 160°F temperature, and 2 lb/ft² tray loading, it was shown that average drying rate in a modified CFB (air velocity 2400 ft/min) was 5.3 times the crossflow value. This increase in drying rate (three times the theoretical value) was due to high efficiency of the air-to-particle contact achieved in the CFB.

A continuous CFBD was further designed (Figure 25.4) with a dryer surface of approximately 21 ft² in the form of a rotating perforated stainless steel cylinder (10-in. diameter and 100-in. long) with an open area of 45% and Teflon-coated inside [30]. The cylinder could be rotated at speeds up to 350 rpm ($F_c = 17.4 \times G$) through a belt drive and tilted between 0° and 6° from the horizontal to help control the residence time of material that is dried. Centrifugal fans with steam heaters enabled air temperatures up to 140°C.

Table 25.4 gives the performance data from trials for drying bell peppers, beets, carrots, cabbages, onions, and mushrooms using a CFBD [30,31]. Good continuous operation was achieved for a 1-h period. Feed rates and evaporation (kg/h) are given for a range of dryer sizes in Table 25.5 for cabbages, carrots, onions, and mushrooms [31].

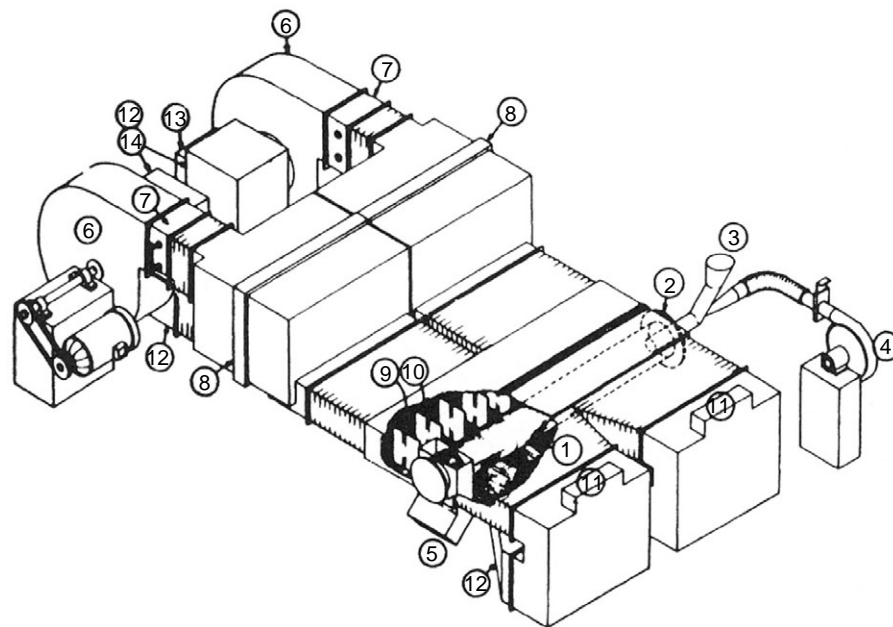


FIGURE 25.4 Isometric view of centrifugal fluidized bed drying system. (1, dryer cylinder; 2, drive pulley; 3, aspiration feeder; 4, feeder blower; 5, discharge chute; 6, air blower; 7, air discharge damper; 8, steam coil heater; 9, plenum; 10, air vent; 11, vent port; 12, recirculating duct; 13, make-up air; 14, blower intake). (From Hanni, P.F., Farkas, D.F., and Brown, G.E., *J. Food Sci.*, 41(5), 1172, 1976. With permission.)

TABLE 25.4**Operating Conditions for Drying Some Vegetables in Continuous Centrifugal Fluidized Bed Dryer**

Commodity	Feed Rate (kg/h)	Discharge Rate (kg/h)	Moisture (%)		Weight Reduction (%)	Temp. (°C)	Air Velocity (m/s)
			Feed	Discharge			
Bell pepper, diced	142	71	93.4	86.1	53	71	15.3
Beet, diced	133	74	84.6	74.5	40	99	15.3
Carrot							
Flaked	109	79	88.9	84.6	28	93	15.3
Diced	130–150	—	89.5	82.0	46	100–140	15.3
Cabbage, shredded	90–200	—	93.3	88.0	44	100–140	15.3
Onion, sliced	150–160	—	87.7	82.5	35	100–140	15.3
Mushroom, diced	230	—	95.3	91.3	48	100–140	15.3

Source: From Hanni, P.F., Farkas, D.F., and Brown, G.E., *J. Food Sci.*, 41(5), 1172, 1976; Cannon, M.W., *Food Technol. New Zealand*, 13(9), 28, 1978. With permission.

Whereas the CFBD can take the product to any degree of dryness, it is considered best suited for the rapid removal of moisture (30–50% weight reduction in about 5-min exposure) during the early stages of drying of piece-form vegetables as a pre-dryer, to be followed by a conventional tray or band dryer for later stages of evaporation in which the rate of moisture removal is governed by diffusion and high velocity is no longer advantageous. Incorporated upstream of the existing dehydration line, it increases overall output with a saving in floor space.

By using a whirling fluidized bed containing inert particles like glass beads, it was found feasible to dry coarse-size and sticky materials like diced potatoes and carrots [32]. A novel type of FB dryer, known as a toroidal fluidized bed, reported to be manufactured in the United Kingdom [24], could be used for a number of processes such as cooking, expanding, roasting, and drying. A high-velocity stream of heated air entering the base of the process chamber through blades or louvers that imparted a rotary

motion to the air created a compact, rotating bed of particles that varied in depth from a few millimeters to in excess of 50 mm. High rates of heat and mass transfer could be attained, resulting in rapid drying. This dryer could be utilized for a wide range of particle sizes and shapes of materials like peas, beans, diced potatoes, and carrots and operated on a continuous or batch basis.

25.4.5 EXPLOSION PUFFING

The technique of explosion puffing was initially developed to fulfill the objective of dehydrating relatively large pieces of fruits and vegetables that would reconstitute rapidly; the system would be operable at a cost comparable to conventional hot air drying. The method, adequately described and extensively reviewed in several articles [23,33], consisted of initially partially dehydrating the fruit and vegetable pieces, then imparting a porous structure by explosion puffing, and subsequently drying to a low

TABLE 25.5**Feed Rates and Evaporation (kg/h) for a Range of Continuous Centrifugal Fluidized Bed Dryer Sizes**

Dryer Size (m)	Cabbages Feed	Carrots Evaporation	Onions Feed	Mushrooms Evaporation	Feed	Evaporation	Feed	Evaporation
0.305 diameter × 2.13	133	58	130	54	156	44	231	106
0.50 diameter × 5.0	658	285	643	266	771	215	1143	425
0.65 diameter × 6.5	1263	546	1232	511	1478	412	2190	1003
0.80 diameter × 8.0	2130	921	2077	861	2491	696	3693	1694
1.00 diameter × 10.0	3719	1607	3628	1504	4352	1216	6451	2958

Source: From Cannon, M.W., *Food Technol. New Zealand*, 13(9), 28, 1978. With permission.

moisture content. Initial drying was required to reduce the moisture content to a level so that disintegration did not occur during explosion puffing. Since uniformity was essential for optimum results, an equilibration step was desirable after the partial drying. As an operational step integrated in hot air dehydration at moisture contents of 15–35%, explosion puffing created porosity in food pieces and speeded up hot air drying, modifying or eliminating diffusion controlled drying as the rate-controlling step. The case hardening problem was minimized so that processors could dry large pieces economically in shorter times, lessening browning potential. Also increased overall volume recovery on rehydration was reported compared with hot air drying. Batch models with output of 180 kg/h of 1-cm diced potatoes or carrots were designed and tested.

The gun used in batch model explosion puffing was essentially a rotating cylindrical pressure chamber that was fitted with a quick-release lid, and was heated externally. The rotational speed of the gun was fixed to give an optimal tumbling action of the charge. This

speed (33 rpm) was about 40% of the critical speed, that is, the speed at which the centrifugal and gravitational forces are equal and no tumbling takes place. In the gun, the pieces were exposed to 10–70 psig steam so that they were quickly heated and their remaining water was superheated relative to atmospheric pressure. When the pieces were suddenly discharged into the atmosphere, the rapid pressure drop caused some of the water within the pieces to flash into steam. The escaping steam caused channels and fissures, thus imparting a porous structure to the pieces. Commodities that were successfully dehydrated by this method include potatoes, carrots, beets, cabbages, sweet potatoes, apples, and blueberries.

A continuous explosive puffing system (CEPS) with 680-kg/h capacity was designed by separating the heating and puffing functions and successfully tested [34]. The three subassemblies that were unique to the system were the feed chamber, the heating chamber, and the discharge chamber (Figure 25.5). The use of CEPS resulted in better process control, improved product quality, and reduced labor costs.

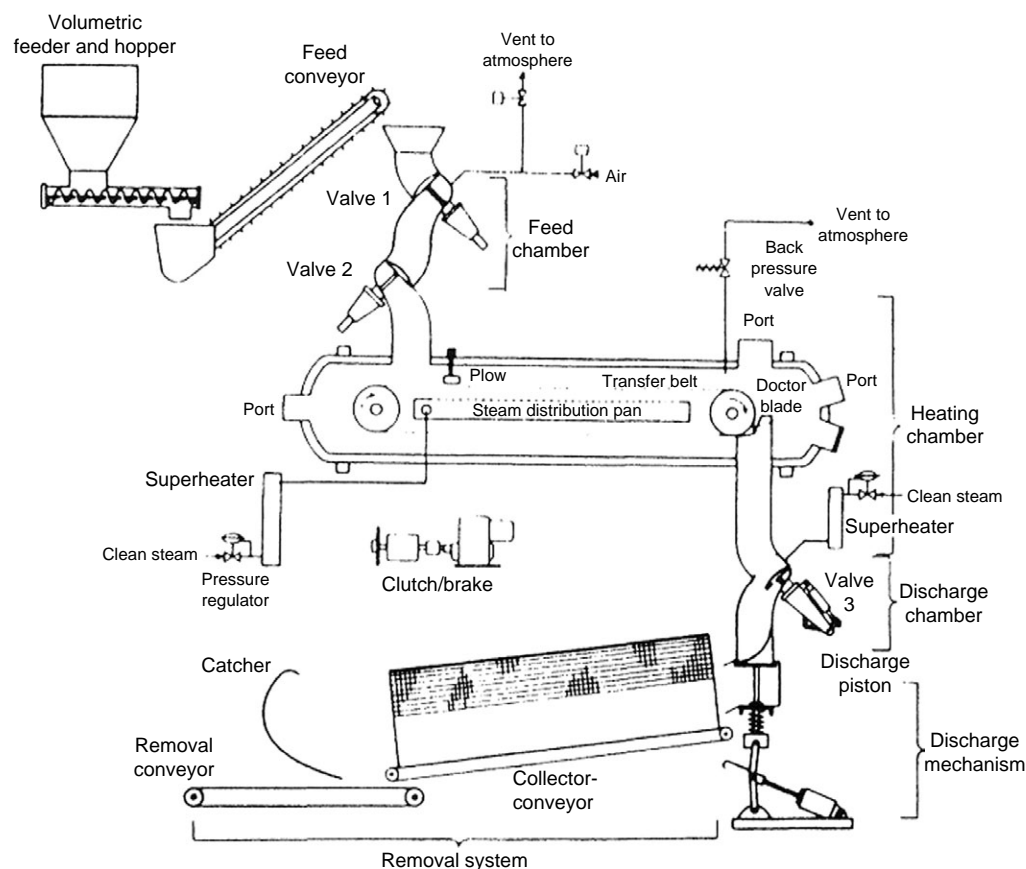


FIGURE 25.5 Schematic diagram identifying major components of continuous explosion puffing systems. (From Heiland, W.K., Sullivan, J.F., Konstance, R.P., Craig, J.C., Jr., Cording, J., Jr., and Aceto, N.C., *Food Technol.*, 31(11), 32, 1977. With permission.)

Once the system feed rate, feed moisture content, internal pressure, internal temperature, and discharge rate reached steady state, it operated with minimal care and needed only occasional operational adjustment [33].

Energy evaluation based on steam consumption showed a 44% reduction in steam consumption when a CEPS was used to dehydrate apple pieces as compared with conventional dehydration; this is attributed to the time saved for drying from 20% to less than 3% moisture. Process cost for EPS is reported to be similar to the cost of conventional hot air drying. Table 25.6 gives processing conditions (batch versus continuous) for a number of fruits and vegetables [35].

25.4.6 FOAM DRYING

The foam drying process is limited to specific products, such as fruit powders, for preparation of instant drinks. Techniques like vacuum puff drying, foam mat drying, microflake dehydration, and foam spray drying have been described elsewhere in this book. Among these, the foam mat drying process has received considerable attention.

Foam mat drying, originally developed by Morgan, involves drying thin layers of stabilized foam from liquid food concentrates in heated air at atmospheric pressure. Foam is prepared by the addition of a stabilizer and a gas to the liquid food in a continuous mixer. It can be dried in a continuous belt-tray dryer. Good quality powders capable of instant rehydration were made experimentally from tomatoes, oranges, grapes, apples, and pineapples [22].

Foam formation is the primary requirement of this process. Two characteristics required for foam stability are consistency and film-forming ability. Film-forming components used in the drying of fruits and vegetables are glyceryl monostearate, solubilized soya protein, and propylene glycol monostearate. Drying time and temperatures depend on the product that is dried; most fruit juices required about 15 min at 160°F to dry to about 2% moisture. Air velocity and humidity had no appreciable effect on the time required.

Foam mat drying has two definite advantages [36]. First, the use of foam greatly speeds up moisture removal and permits drying at atmospheric conditions in a stream of hot air in a short time. Second, though the product may be sticky at drying temperatures, it can be transferred to a cooling zone and crisped before it is scraped off the surface.

25.4.7 MICROWAVE DRYING

In microwave drying, heat is generated inside the food materials by the interaction of chemical constituents of food and radio frequency energy (915 and 2450 MHz). Use of this type of energy found its application in the finish drying of potato chips. Much of the work on the drying of fruits and vegetables utilizing microwave energy was described by Decareau [37].

The advantages of using microwave energy are penetrating quality, which effects a uniform heating of materials upon which radiation impinges; selective absorption of the radiation by liquid water; and capacity for easy control so that heating may be rapid if desired.

TABLE 25.6
Process Conditions for Explosive Puffing (Batch and Continuous) of Some Vegetables and Fruits

Commodity	Puffing Moisture (%)	Steam Pressure (kPa)	Temp. (°C)	Dwell Time (s)	Rehydration Time (min)
Potatoes	25	414	176	60	5
Carrots	25	275	149	49	5
Yams	25	241	160	75	10
Beets	20–26	276	163	120	5
Peppers	19	207	149	45	2
Onions	15	414	154	30	5
Celery	25	275	149	39	5
Rutabagas	25	241	160	60	6
Mushrooms	20	193	121	39	5
Apples	15	117	121	35	5
Blueberries	18	138	204	39	4
Cranberries	17–26	138	163	64	3
Strawberries	25	90	177	—	3
Pineapples	18	83	166	60	1
Pears	18	228	154	60	5

Source: From Kozempel, M.F., Sullivan, J.F., Craig, J.C., Jr., and Konstance, R.P., *J. Food Sci.*, 54(3), 772, 1989. With permission.

It can reduce drying time, particularly when the size of the piece is such that a conventional drying method is not feasible. However, the high cost per unit of energy compared with the conventional energy and the high initial cost of equipment limits its use for drying.

Microwave vacuum drying of concentrates of oranges, lemons, grapefruits, pineapples, strawberries, and others has been described [37]. One full-scale plant was in operation for the vacuum drying of orange and grapefruit juices, utilizing a 48-kW, 2450-MHz unit that dried 63° Brix orange juice concentrate to 2% moisture in 40 min.

The various modes in which microwaves are used in the industry comprised of booster (microwave-connection) and dryers (microwave vacuum dryers and microwave freeze dryers). Microwave vacuum drying of cranberries was investigated using laboratory-scale dryer operating either in continuous or pulsed mode [38]. Pulsed application of microwave energy was found to be more efficient than continuous application.

Microwave-assisted hot air drying of golden delicious apple and mushroom was examined [39] using a novel applicator to reduce the edge overheating effects. Microwave drying considerably reduced the drying time of potato as well as produced better quality dried product [40]. In another investigation [41] it was found that microwave drying of thin layer carrot resulted in substantial decrease in drying time and better quality product when dried at low power level.

Microwave drying was combined with spouted bed fluidization technique (MWSB) for the drying of frozen blueberries [42]. MWSB drying was characterized by a substantial reduction in drying time and an improved product quality compared to freeze-, tray-, and SB-dried samples.

25.4.8 SPRAY DRYING

The spray drying method is most important for drying liquid food products and has received much experimental study. Spray drying by definition is the transformation of a feed from a fluid state into a dried form by spraying into a hot, dry medium [43]. In general it involves atomization of the liquid into a spray (by a nozzle) and contact between the spray and the drying medium (hot air), followed by separation of dried powder from the drying medium (by a cyclone separator). Applicable to a wide range of products, there is no single, standardized design for the spray dryer common to all. Each product is treated individually and the dryer is designed to suit the product specifications. The principles and applications of this technique are well described in the literature [20–24,26,43].

The applications of spray drying to fruit and vegetable products are very limited. Fruit juices, pulps,

and pastes can be spray dried with additives. Special care must be taken to design the drying chamber as well because during postdrying, handling, and packing operations the products are both hygroscopic and thermoplastic. Fruits that have been spray dried include tomatoes, bananas, and, to a limited extent, citrus fruit, peaches, and apricots.

Tomato pulp is a typical example of a product that is very difficult to dry as the powder is sticky and poses a caking problem. A spray drying plant capable of producing a free-flowing product that on reconstitution compares favorably with tomato paste has been designed featuring a cocurrent drying chamber having a jacketed wall for air-cooling and a conical base. Cooling air intake is controlled to enable close maintenance of wall temperature in the range of 38–50°C. The paste is sprayed into the drying air entering the chamber at a temperature of 138–150°C.

A wide range of vegetables can be spray dried following homogenization and the powders can be readily used in dry soup mixes. As yet, there is limited interest in spray drying of vegetables though the drying process is not different from fruits and standard equipment can be used. Jayaraman and Das Gupta [44] spray dried a number of fruit juices in admixture with whole milk or yogurt.

Spray drying of a mixture of eight vegetable juices (tomato, cucumber, parsley, lettuce, beet, spinach, carrot, and celery juices) using 1% maltodextrin as additive was described [45].

25.4.9 DRUM DRYING

Drum drying is an important and inexpensive drying technique suitable for a wide range of products namely liquid, slurry, and puree. The material to be dried is applied as a thin layer to the outer surface of a slowly revolving hollow drum (made of iron or stainless steel) heated internally by steam [26]. The principle and types of drum dryers have been discussed by a number of authors [20,22,24,26,27].

The success of drum drying depends on the application of a uniform film of maximum thickness. The high rate of heat transfer is obtained by direct contact with the hot surface and the equipment may be used under atmospheric or vacuum condition [23]. It is mainly used for the manufacture of potato flakes. Its usefulness for dehydration of fruits (particularly fruits high in sugar and low in fiber content) is limited by the high temperature required. Thin sheets of very dry fruit are usually so hygroscopic that it has been necessary to overdry under severe heating conditions to compensate for the later pick up. The fruit was therefore usually heat damaged.

A pilot-plant, double-drum dryer modified to produce low-moisture flakes from a wide range of fruit purees has been described [46]. Products with a relatively high fiber content such as apples, guavas, apricots, bananas, papayas, and cranberries could be dried successfully without additives. Purees with a low fiber content such as raspberries, strawberries, and blueberries required the addition of fiber (low methoxyl pectin, up to 1%) to aid in the sheet formation at the doctor blade. The modification consisted of incorporation of variable-speed take-off rolls, cool airflow directed at the doctor blade area, and a ventilation system to remove saturated air from the area beneath the drums.

A process for manufacture of instant, drum-dried flakes from tropical sweet potato puree was evaluated using a Buflovac laboratory model atmospheric double-drum dryer internally steam heated at 35 psig [47]. The drums revolved at 1.73 rpm with a clearance between drums of 0.305 mm. It was found that pretreatment with α -amylase improved the drying characteristics of the puree.

A mathematical model was predicted for drum drying of mashed potatoes on the basis of primary process parameters such as drum speed, steam pressure, number of spreader rolls, wet and dry bulb temperatures, mash moistures, and drum dimension [48].

25.4.10 FREEZE-DRYING

Freeze-drying, which involves a two-stage process of first freezing of water of the food materials followed by application of heat to the product so that ice can be directly sublimed to vapor, is already a commercially established process. Sublimation from ice to water vapor can only be accomplished below the triple point of water, that is, at 4.58 torr at a temperature of approximately 32°F. Since the moisture removal does not pass through a liquid phase, the structure of the product remains in a more acceptable state. In addition, drying takes place without exposing the product to excessively high temperatures.

The advantages of freeze-drying are: shrinkage is minimized; movement of soluble solids within the food material is minimized; the porous structure of the product facilitates rapid rehydration; and retention of volatile flavor compounds is high. It has therefore proved to be the superior method of dehydration for many fruits. The major limitation to its commercial application is its very high capital and processing costs and the need for special packaging to avoid oxidation and moisture pick up. Industrial application includes some exotic fruits and vegetables, soup ingredients, mushrooms, and orange juice. Much of

the recent work is directed toward freeze-dried fruit juices and vegetables like spinach and carrots.

Essential components of a freeze dryer include the vacuum chamber, condenser, and vacuum pump. As in other forms of drying, freeze-drying represents coupled heat and mass transfer. For the analysis of this operation, Karel [26] considered three cases that represent three basic types of possibilities in vacuum freeze-drying: (a) heat transfer and mass transfer pass through the same path (dry layer) but in opposite directions; (b) heat transfer occurs through the frozen layer and mass transfer through the dry layer; and (c) heat generation occurs within ice (by microwaves) and mass transfer through the dry layer [26]. The principles and applications of freeze-drying are described in detail in many books and articles [22–24,27,49].

Another aspect that determines the structure of food materials, particularly fruit juices, during freeze-drying is the phenomenon of collapse. Freezing of food materials causes aqueous solution to be separated into two phases: ice crystals and concentrated aqueous solution. The properties of this concentrated aqueous solution depend on composition, concentration, and temperature. If during drying the temperature is very low, the mobility in the extremely viscous concentrated phase is so low that no structural changes occur during drying. But, if the temperature is above a critical level (known as the collapse temperature), mobility of the concentrated solution phase may be so high that flow and loss of original structure occurs. This is known as the phenomenon of collapse and was investigated in detail by several workers.

Atmospheric freeze-drying of several foods, including mushrooms and carrots, was investigated in a fluidized bed of finely divided adsorbent that combined adsorption and fluidization, achieving improved heat and mass transfer and shorter drying time than vacuum drying [50,51]. Products could be dried economically using very simple equipment.

Bell and Mellor [52] developed an adsorption freeze-drying process that depended upon the removal of water vapor by a desiccant rather than by refrigeration coils. The process consisted of a chamber in which the air pressure was reduced, a product rack to hold the samples, and a perforated container of desiccant that required regeneration. Defrosting, drying the chamber, and vacuum pretesting were not required because the inside of the chamber remained dry.

A combination of thermal and freeze-drying processes was tried on apple, potato, and carrot and was demonstrated [53] to be a promising technique in the production of high-quality dehydrated fruits and vegetables. A combined drying technology, initially by osmotic dehydration following by freeze-drying on

apple and potato was reported [54] to produce a high-quality product with lower freeze-drying times.

25.4.11 OSMOTIC DEHYDRATION

Osmotic dehydration is a water removal process that consists of placing foods, such as pieces of fruits or vegetables, in a hypertonic solution. As this solution has higher osmotic pressure and hence lower water activity, a driving force for water removal arises between solution and food, whereas the natural cell wall acts as a semipermeable membrane. As the membrane is only partially selective, there is always some diffusion of solute from the solution into the food and vice versa. Direct osmotic dehydration is therefore a simultaneous water and solute diffusion process [55]. Up to a 50% reduction in the fresh weight of the food can be achieved by osmosis. Its application to fruits and, to a lesser extent, to vegetables has received considerable attention in recent years as a technique for production of intermediate moisture foods (IMF) and shelf-stable products (SSP) or as a predrying (preconcentration) treatment to reduce energy consumption and heat damage in other traditional drying processes.

Some of the stated advantages of direct osmosis in comparison with other drying processes include minimized heat damage to color and flavor, less discoloration of fruit by enzymatic oxidative browning, better retention of flavor compounds, and less energy consumption since water can be removed without change of phase. However, products cannot be dried to completion solely by this method and some means of stabilizing them is required to extend their shelf lives.

Many workers have studied the different aspects of osmotic dehydration: the solutes to be employed, the influence of process variables on drying behavior, the opportunity to combine osmosis with other stabilizing techniques, and the quality of the final products. The osmotic agents used must be nontoxic and have a good taste and high solubility besides low a_w . Sugar in different concentrations is widely used. Common salt is an excellent osmotic agent for vegetables.

The quantity and the rate of water removal depend on several variables and processing parameters. In general it has been shown that the weight loss in osmosed fruit is increased by increasing the solute concentration of the osmotic solution, immersion time, temperature, solution-to-food ratio, specific surface area of the food, and by using vacuum, stirring, and continuous reconcentration. Also, to obtain the same a_w reduction, time tended to decrease exponentially as the temperature is increased.

Several models were proposed to show the effect of concentration of osmotic solution and temperature on the rate of water loss and gain of osmotic agent.

Thus, a model developed [56] for the calculation of osmotic mass transport data for potato and water activity to equilibrium in sucrose solutions for the concentration range 10–70% and solution/solids range 1–10 showed that, at equilibrium, there was an equality of water activity and soluble solids concentration in the potato and in the osmosis solution. A linear relationship existed between normalized solids content (NSC) and $\log(1 - a_w)$ and was given by

$$\text{NSC} = 6.1056 + 2.4990 \log(1 - a_w)$$

Another model developed [57] for solute diffusion in osmotic dehydration of apple based on solids gain divided by water content M as a function of rate constant K , time (t), and a constant A was given as $M = Kt + A$. A relationship was established in the form of $K = T^{1.40}C^{1.13}$, where rate parameter K is related to temperature T at different sucrose concentrations C . The average activation energy of the process was 28.2 kJ/mol.

The effects of solution concentration, osmosis time, and the osmosis temperature were studied in the osmotic dehydration of pineapple in sucrose solution [58]. The solute diffusion was analyzed by Magee's model. The effect of sucrose concentration C on rate parameter K was given by power law regression equation as $K = 4.15 \times 10^{-4}C^{1.51}$ at 20°C.

An empirical equation derived based on osmotic dehydration of apple slices could predict rate of osmosis F , that is, percentage of dehydration of any given fruit slices of specific size with time T , given the concentration of sugar (% B) and the temperature as follows [59]:

$$F = 31.8 - 0.307B - (0.56 - 0.016B)t - 2.10^{-9.26/B} - 1(T - 0.3)^{0.54} - 0.00425t$$

where F is the decrease in mass %, and was valid for $B = 60\text{--}75\%$, $t = 40\text{--}80^\circ\text{C}$, and $T = 0.5\text{--}4.5$ h.

Direct osmosis of different fruits at 70° Brix sugar at atmospheric and low pressure (about 70 mmHg) revealed higher drying rates with the latter. The addition of a small amount of NaCl to different osmotic solutions increased the driving force of the drying process.

Apple cubes submitted to HTST osmosis in sugar at 60–80°C for 1–20 min showed osmosis to be greatly accelerated by high temperature, as the water loss in apples after 1–3 min HTST osmosis was the same as that given by 2-h treatment at ambient temperature and HTST osmosis completely inactivated the enzymes.

Partial dehydration of fruits and vegetables by osmosis using various osmotic agents has been employed

before drying by other conventional methods, namely, hot air convection drying, high-temperature fluidized bed drying, vacuum drying, freeze-drying, and dehydrofreezing as a means of reducing processing time and limiting energy consumption besides improving sensory characteristics.

Osmotic dehydration has been utilized for developing intermediate moisture fruits stabilized solely by a_w control with added antimycotic preservative, as well as SSP with higher a_w stabilized by a combination preservation technique involving a_w and pH control plus heat pasteurization, due to simplicity of the operations involved, economy, and low-energy inputs.

25.4.12 HEAT PUMP DRYING

To improve the thermal economy and efficiency of conventional hot air dryer, use of heat pump technology was utilized for the development of heat pump dryer. In its simplest form, the heat pump dryer passes the drying air over the evaporator of a refrigeration system. This cools the air to below its dew point, condensing water vapor from the air stream. This cool air is then passed over the condenser over the refrigeration system to reheat the air to drying temperature. Most available heat pump dryers recirculate all the air, but nonrecirculating types are also available. Both types can be highly energy efficient [60].

The three major advantages of heat pump dryers are [60]:

1. Drying at low temperatures can improve quality
2. Higher energy efficiencies are achieved because both the sensible and the latent heat of evaporation are required
3. Drying conditions and therefore drying rate is unaffected by drying conditions

Against these advantages, a number of factors limit the application of heat pump drying. These include the use of electrical energy which is generally more expensive than other forms, higher capital cost and that the maximum drying temperature is limited to around 60°C to 70°C with currently used refrigerants.

Typical drying temperatures in a heat pump dryer are in the range 30°C–60°C. It is expected that drying by this technique would improve the retention of volatile flavor, reduce the color degradation as well as the loss of heat-labile vitamins [61].

25.4.13 ULTRASONIC DRYING OF LIQUIDS

The utilization of ultrasonic energy to remove water from dilute solution of nonfatty products was reported [62]. In this process the liquid is atomized

through a nozzle initially and then by cavitation using ultrasonic energy. An ultrasonic technique for drying of vegetables using a power ultrasound generator was reported [63]. In this technique high-intensity ultrasonic vibrations were used to investigate the drying of carrot slices and effect of this technique were compared with those of conventional drying and forced air-drying assisted by airborne ultrasonic radiation. Dramatic reduction in drying time was achieved maintaining the quality.

25.5 QUALITY CHANGES DURING DRYING AND STORAGE

25.5.1 LOSS OF VITAMINS (VITAMINS A AND C)

Fruits and vegetables are the major sources of vitamin C (ascorbic acid) and provitamin A (β -carotene) besides minerals. It is, therefore, quite understandable that to determine the efficacy of dehydration techniques scientists have primarily investigated and compared the effect such techniques have on these nutrients.

The effect of predrying treatments, dehydration, storage, and rehydration was studied [64] on the retention of carotene in green peppers and peaches during home dehydration. Carotene was completely retained in the case of green peppers. In peaches, 72.7% of the carotene was retained after predrying treatment, which decreased to 37.3% after dehydration. Retention of ascorbic acid during predrying treatment and dehydration depended on the nature of food. Thus, in the case of green peppers, most losses occurred during storage whereas dehydration was responsible for most of the loss in the dipped peaches.

In general, rapid drying retained a greater amount of ascorbic acid than slow drying. Thus vitamin C contents of vegetable tissue are greatly reduced during a slow sun-drying process, whereas dehydration, especially by spray drying and freeze-drying, reduced these losses. The effect of sun drying on the ascorbic acid content of 10 Nigerian vegetables showed that there was 21–58% loss depending on the nature of the vegetables [65].

Oxidative changes would be expected to be minimum in freeze-dried samples as freeze-drying is a low-temperature process operating under vacuum. A study [66] of the changes in quality of compressed carrots prepared in combinations of freeze-drying and hot air drying showed that value of ascorbic acid ranged from 15.97 mg/100 g for the totally air-dried samples to 33.39 mg/100 g for the totally freeze-dried samples (Table 25.7). In the case of carotenes also the totally hot air-drying treatment had

TABLE 25.7

Effect of Drying Treatment on Ascorbic Acid and α -Tocopherol of Dehydrated Carrots (mg/100 g Dry Weight Basis)^a

Treatment (% Moisture)	Ascorbic Acid	α -Tocopherol
Fresh	85.28	3.41
Totally freeze-dried	33.39 a	3.45 a
Totally air-dried	15.97 d	0.04 f
Freeze-dried (30%), mist plasticized (10%), air-dried	32.76 a	2.98 b
Freeze-dried (10%), air-dried	27.71 b	1.42 c
Freeze-dried (20%), air-dried	16.78 cd	1.13 d
Freeze-dried (30%), air-dried	16.38 cd	1.10 d
Freeze-dried (40%), air-dried	20.38 c	0.96 d
Freeze-dried (50%), air-dried	17.49 cd	0.55 e

^aMeans within columns followed by the same letter are not significantly different at the 5% level according to Duncan's multiple range test.

Source: From Shadle, E.R., Burns, E.E., and Talley, L.J., *J. Food Sci.*, 48(1), 193, 1983. With permission.

the lowest value (34.16 mg/100 g) and totally freeze-dried samples had the highest value (70.37 mg/100 g) (Table 25.8).

The effect of blanching, various drying methods (sun, vacuum oven, and hot air oven), and drying temperature (33–60°C) on ascorbic acid content of okra was investigated [67]. Blanching solution resulted in slight loss in ascorbic acid but led to more retention during dehydration. Vacuum dehydrated sample retained more ascorbic acid at each of the dehydration temperature than those from hot air oven. Vacuum microwave drying of carrot was compared to air-drying and freeze-drying on the basis of α - and β -carotene and vitamin C content. Total losses of α - and β -carotene during drying was 19.2% for air-drying and 3.2% for vacuum microwave drying samples. Loss of vitamin C content was substantial due to blanching [68]. The effect of blanching and drying methods on the β -carotene and ascorbic acid retention in three leafy vegetables, i.e., savoy beet, amaranth, and fenugreek showed [69] that the most suitable method for blanching was thermal treatment in water at $95 \pm 3^\circ\text{C}$ followed by potassium metabisulfite dip and drying at low temperature for the retention of ascorbic acid as well as β -carotene. The retention of ascorbic acid and β -carotene was reported to be 15.0%, 49.7% for savoy beet; 40.5%, 98.5% for amaranth, 54.6%, 91.5% for fenugreek after blanching, and 7.5%, 39.7%; 30%, 48.5; 49.7%; 85.1%, respectively after low-temperature drying.

TABLE 25.8

Effect of Drying Treatment on Carotene Content of Dehydrated Carrots (mg/100 g Dry Weight Basis)^a

Treatment (% Moisture)	α -Carotene	β -Carotene	Total Carotene
Fresh	14.4	52.06	66.20
Totally freeze-dried	15.66 a	54.71 a	70.37 a
Totally air-dried	6.67 e	27.50 f	34.16 f
Freeze-dried (3%), mist plasticized (10%), air-dried	10.61 d	40.47 e	51.08 e
Freeze-dried (10%), air-dried	12.81 b	49.40 b	62.21 b
Freeze-dried (20%), air-dried	11.73 c	44.49 d	56.22 d
Freeze-dried (30%), air-dried	11.42 cd	47.22 c	58.68 c
Freeze-dried (40%), air-dried	11.02 cd	44.89 d	55.91 d
Freeze-dried (50%), air-dried	10.52 d	40.23 c	50.81 e

^aMeans within columns followed by the same letter are not significantly different at the 5% level according to Duncan's multiple range test.

Source: From Shadle, E.R., Burns, E.E., and Talley, L.J., *J. Food Sci.*, 48(1), 193, 1983. With permission.

In general it is difficult to compare the losses in vitamins during dehydration because retention of vitamins depends on the nature of foods, predrying treatments given (sulfuring, blanching methods), and the conditions of drying (techniques, time, and temperature).

25.5.2 LOSS OF NATURAL PIGMENTS (CAROTENOIDS AND CHLOROPHYLLS)

Color is an important quality attribute in a food to most consumers. It is an index of the inherent good qualities of a food and association of color with acceptability of food is universal. Among the natural color compounds, carotenoids and chlorophylls are widely distributed in fruits and vegetables. The preservation of these pigments during dehydration is important to make the fruit and vegetable product attractive and acceptable. Both the pigments are fat-soluble although they are widely distributed in aqueous food systems.

Carotenoids are susceptible to oxidative changes during dehydration due to the high degree of unsaturation in their chemical structure. The major carotenoids occurring in food are carotenes and oxycarotenoids (xanthophylls).

Leaching of soluble solids during blanching had considerable effect on the stability of carotenoids of carrots during drying and subsequent storage [70]. Carotenoid destruction increased with increased leaching of soluble solids. Investigation of the effects of water activity, salt, sodium metabisulfite, and Embanox-6 on the stability of carotenoids in dehydrated carrots shows that carotenoid pigments were most stable at $0.43a_w$ and addition of salt, metabisulfite, and Embanox-6 helped in stabilizing carotenoids in dehydrated carrots (Table 25.9) [71].

Sulfur dioxide was found to have a pronounced protective effect on carotenoids of unblanched carrots during dehydration [72]. Dehydrated, sulfited, unblanched carrots contained about 2.9 times more carotenoids than dehydrated unblanched carrots that had not been sulfited (Table 25.10). Treatment with SO_2 gave additional protection to carotenoids of blanched carrots during dehydration and effectiveness of SO_2 increased with an increase in SO_2 content.

The importance of chlorophyll in food processing is related to the green color of vegetables. Many studies have been made on the changes of chlorophyll during processing and storage but little is known about the pigment behavior in low-moisture systems such as dehydrated vegetables. Generally, it was found that chlorophyll was quite stable in low-moisture systems. Degradation of chlorophyll depended on temperature, pH, time, enzyme action, oxygen, and light. The most common mechanism of chlorophyll degradation is its conversion to pheophytin in the presence of acid. Although the pathways of this degradative reaction are well-known, a method for its stabilization is not well-established.

Water activity has been shown to have a definite influence on the rate of degradation of chlorophyll in freeze-dried, blanched spinach puree [73]. At $37^\circ C$ and an a_w higher than 0.32, the most important mechanism of chlorophyll degradation was conversion to pheophytin. At a_w lower than 0.32, the rate of pheophytin formation in spinach was low. The rate of chlorophyll-*a* transformation was 2.5 times faster than chlorophyll-*b*. The study of the degradation of chlorophyll as a function of a_w , pH, and temperature in a spinach system during storage showed that even in the dry state the elimination of a magnesium atom and transformation of chlorophyll into pheophytin was very sensitive to pH changes [74]. Effect of temperature on the rate of chlorophyll-*a* degradation at water activity 0.32 and pH 5.9 is shown in Figure 25.6.

25.5.3 BROWNING AND ROLE OF SULFUR DIOXIDE

One obstacle always encountered by the food technologists in the dehydration and long-term storage of

dehydrated fruits and vegetables is the discoloration due to browning. Browning in foods is of two types: enzymatic and nonenzymatic. In the former, the enzyme polyphenol oxidase catalyzes the oxidation of mono- and ortho-diphenols to form quinones that cyclize, undergo further oxidation, and condense to form brown pigments (melanins). In the dehydration of fruits and vegetables, blanching destroys the causative enzymes and prevents subsequent enzymatic browning. Sulfur dioxide and sulfites act as inhibitors of enzyme action during preblanching stages. The presence of SO_2 retards browning of dehydrated fruits and vegetables, especially when the enzymes have not been heat-inactivated (e.g., freeze-dried products).

NEB, also known as Maillard reaction, describes a group of diverse reactions between amino groups and active carbonyl groups, leading eventually to the formation of insoluble, brown, polymeric pigments, collectively known as melanoidin pigments. The basic reactions that lead to the browning are well documented in the literature. These reactions are sometimes desirable but in many instances are considered to be deleterious not only due to the formation of unwanted color and flavor but also due to the loss of nutritive value through the reactions involving the α -amino group of lysine moieties and other groupings in proteins. It is a major deteriorative mechanism in dry foods and is sensitive to water content. It is influenced by the types of reactant sugars and amines, pH, temperature, and a_w .

The addition of sulfites during the predrying step is the only effective means available at present controlling NEB in the dried fruit and vegetable product. Sulfite is considered to be a safe additive to incorporate into fruit and vegetable products up to certain permissible limits. However, recently there are reports on the hypersensitivity of a few individuals to the ingested sulfite. Numerous attempts are therefore made to find alternative means to prevent browning reactions.

Among various treatments studied, such as addition of SO_2 , cysteine, $CaCl_2$, trehalose, manganese chloride, disodium dihydrogen pyrophosphate, oxygen scavenger pouch, and so on, the only ones that effectively retarded the formation of undesirable pigment in dried apples during storage were oxygen scavenging and sulfur dioxide [75]. Apples stored in oxygen scavenger packages darkened slower than those stored under regular atmospheric conditions, exhibiting a different initial induction period (Figure 25.7).

The effectiveness of sulfite in controlling the family of diverse reactions, leading to browning is probably due to the number of different reactions that sulfite can enter into with reducing sugars, simple carbonyls, α -, β -dicarbonyls, β -hydroxycarbonyls,

TABLE 25.10

Effect of Concentration of SO₂ on Carotenoid Content of Dehydrated Carrot of 5% Moisture Content during Storage at 37°C

Blanching Time (min)	Initial SO ₂ Content (μg/g)	Carotenoid Content after Dehydration (μg/g)	Carotenoids Remaining (%)				
			Storage Time (d)				
			60	120	180	300	440
0	0	464	68.0	51.1	43.0	36.2	33.1
0	1723	1296	87.5	76.5	69.4	62.6	55.5
1	2325	1360	92.5	85.0	79.4	69.0	62.0
2	2330	1350	88.7	79.4	71.1	61.7	55.0
5	0	1202	77.5	62.5	56.1	50.2	48.2
5	1584	1298	80.5	67.4	60.5	54.0	50.2
5	2357	1308	87.0	76.1	68.6	58.5	52.0
5	9621	1380	89.9	80.0	73.1	62.8	54.0

Source: From Baloch, A.K., Buckle, K.A., and Edwards, R.A., *J. Sci. Food Agric.*, 40, 179, 1987. With permission.

β-unsaturated carbonyls, and with melanoidins [76]. So far there is no practical substitute for SO₂ as a means of controlling NEB, although lowering pH, dehydration to very low water activity, separation of active species, and addition of sulfhydryl compounds might have limited applications [6].

25.5.4 OXIDATIVE DEGRADATION AND FLAVOR LOSS

The acceptability of dehydrated fruit and vegetable products is highly dependent upon their flavor attributes. Loss of desirable flavor is the limiting characteristic for most dehydrated products. The natural

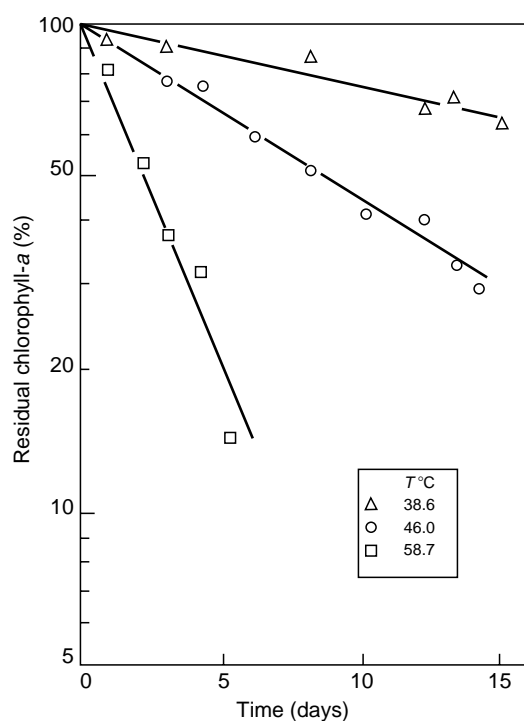


FIGURE 25.6 Degradation of chlorophyll-a in spinach as a function of temperature ($a_w = 0.32$; pH = 5.9). (From Lajolo, F.M. and Marquez, U.M.L., *J. Food Sci.*, 47, 1995, 1982. With permission.)

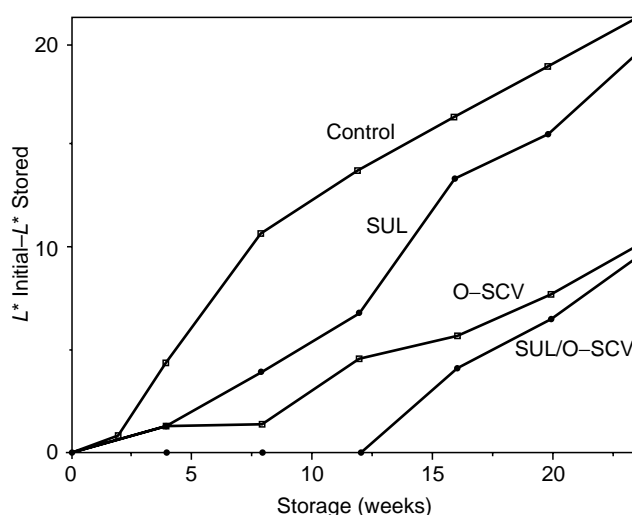


FIGURE 25.7 Effect of in-package oxygen scavenger on dried apple darkening during storage at 30°C (ΔL^* of 8 = observable change). (From Bolin, H.R. and Steele, R.J., *J. Food Sci.*, 52(6), 1654, 1987. With permission.)

flavor constituents are subjected to much variation and loss during predrying operations, drying, and storage. Conditions generally responsible for the destruction of natural flavors include rough handling, delay in processing, exposure to light, high temperature, and certain chemicals. Flavor retention is especially important in products in which the principal flavor constituents are volatile oils, as in onions. Flavor defects in dehydrated products were, however, not solely due to volatile losses. Chemical reactions, especially oxidation and NEB, greatly contributed to flavor deterioration.

In general, freeze-dried products had more preferable flavors than air-dried ones except in the case of onions, for which an air-dried product had a stronger flavor due to entrapment of volatile oils by shrinkage. Leeks and celery showed similar behavior.

Staling and off flavors developed during storage of both air-dried and freeze-dried vegetables. The degree of change was mainly related to temperature of storage and moisture content of the dried vegetables. Air-dried peas (6–7% moisture) developed off flavor at 15°C after 15–18 months. At about 20°C, shelf life was reduced to 9–12 months and at 37°C the period was 2–3 months. Comparatively, freeze-dried vegetables were much more sensitive to storage conditions because the highly porous texture allowed easy entry of air and stale flavor developed rapidly. For example, freeze-dried carrots developed off flavor after 1 month in air at 20°C. At 30°C the oxygen level had to be reduced to 0.1% to give a storage life of 6 months [77].

The absence of oxygen was essential for satisfactory storage of freeze-dried fruits and vegetables.

Excellent retention of fresh flavor quality was achieved in a series of freeze-dried foods of plant origin in zero oxygen headspace, using an atmosphere of 5% hydrogen in nitrogen with palladium catalyst [78]. Vegetable items took up oxygen chiefly as a function of pigment content. Those with a high carotene content (sweet potatoes, spinach, and carrots) underwent a fairly rapid uptake during the first 15–40 weeks and had consumed all available oxygen at the end of 1 year. Lesser-pigmented vegetables with a lower lipid content (green beans and potatoes) showed a slow, steady uptake. Two fruit items, peaches and apricots, displayed a very slow uptake, using only 30–50% of available oxygen during 1 year.

One of the major causes of degeneration of flavor in dehydrated potato products was the Maillard reaction. This aminocarbonyl reaction of reducing sugar and amino acid resulted in the formation of many volatile compounds. Thus, flavor deterioration in potatoes during the explosion-puffing step was attributed to NEB. In the puffing gun, potatoes at 30% moisture were subjected to a temperature condition conducive to NEB, which resulted in the formation of volatile aldehydes. On the other hand, dominant, rancid off flavor that developed during the storage of dried potato products was due to autoxidation of potato lipids [79], giving hexanal as a major volatile product. The use of BHA alone or with BHT effectively retarded the autoxidation of explosion-puffed potatoes, keeping oxidative off flavors below threshold levels for up to 12 months in storage as compared to 3 months for air-packed samples without antioxidant. The incorporation of a scavenger pouch packaging system (H_2 –palladium catalyst), although very

effective in antioxidative effect, was severely limited because of pinhole leaks.

25.5.5 TEXTURE AND RECONSTITUTION BEHAVIOR

The problem of hot air drying, which is still the most economical and widely used method for dehydrating piece-form vegetables and fruits, is the irreversible damage to the texture, leading to shrinkage, slow cooking, and incomplete rehydration. Many commercially dehydrated vegetables exhibit a dense structure with most capillaries collapsed or greatly shrunk, which affects the textural quality of the final product.

The possible causative factors suggested by different workers are loss of differential permeability in the protoplasmic membrane, loss of turgor pressure in the cell, protein denaturation, starch crystallinity, and hydrogen bonding of macromolecules. Texture of air-dried vegetables deteriorates during storage if the product is exposed to high temperature or if inadequately dehydrated. Even the freeze-drying technique has failed to produce an acceptable dehydrated product from celery. Damage generally occurred during freezing, drying, storage, and reconstitution.

Water removal affects many aspects of cell structure; histological studies were generally carried out to assess the membrane integrity. Pedlington and Ward [80], in studies on air-dried carrots, parsnips, and turnips, observed several changes, including a loss in the selective permeability of cytoplasmic membranes of cell responsible for maintaining turgidity and crisp texture of vegetables. They found loss of water to result in rigidity of cell walls and to their slow collapse by the stresses set up by shrinkage of neighboring cells.

Jayaraman et al. [81] studied the effect of sugar and salt to the texture of dehydrated cauliflower. They found that in treated, dehydrated florets there were 80% intact cells as compared with 0% in the untreated, dehydrated florets due to tissue collapse resulting in disruption of cell walls and loss of cell integrity. Khedkar and Roy [82] found a higher reconstitution ration in cabinet-dried raw mango slices as compared with sun-dried slices; this was due to less rupture of cells during cabinet drying (36.4%) than sun drying (67.3%).

Different dehydration techniques were tried to improve the rehydration behavior of dehydrated piece-form fruits and vegetables. Generally, it was observed that the greater the degree of drying, the slower and less complete was the degree of rehydration. Dehydration techniques used to improve the rehydration qualities of dehydrated fruits and vegetables include those aimed at reducing the drying time or involving use of additives like salt and polyhydroxy compounds such as sugar and glycerol as a predrying treatment.

Dehydrated carrots puffed and dried in a CFB unit absorbed 2 1/2 parts by weight of water and appeared completely rehydrated in 5 min whereas the unpuffed controls absorbed 1 1/2 parts and still had hard centers [29]. Jayaraman et al. found rehydration ratio, coefficient rehydration, and reconstitution time of HTST pneumatic-dried vegetables to be much superior to those of directly cabinet-dried samples [23].

The effect of additives on the rehydration qualities of dehydrated vegetables was studied by Neumann [83] and Jayaraman et al. [81]. A combined predrying treatment of sodium carbonate and sucrose (60%) produced the best rehydrated celery, with a rehydration percentage of 71% and the dices were well filled out with texture remaining tender to firm [83]. Similarly, a presoaking treatment in a combined solution of salt and sugar at 4°C for 16 h before cabinet drying markedly increased the rehydration percentage of cauliflower and reduced the shrinkage as compared with control without treatment [81].

The study of the rehydration ratios of forced air-dried compressed carrots after partially freeze-drying to different moisture levels showed the drying treatment significantly affected rehydration ratios in all cases [66]. The sample that was freeze-dried to 50% moisture, compressed, and then air-dried had the highest ratio and was the quickest to rehydrate. In comparison, the totally freeze-dried and hot air-dried compressed carrots showed much lower values of rehydration ratios. These observations were supported by scanning electron microscopy (SEM), which showed collapse of cellular structure and tissue coagulation to act as a barrier for rehydration.

Levi et al. [84] observed that pectin, one of the major cell wall and intercellular tissue components, played a significant role in the rehydration capacity of dehydrated fruits.

25.5.6 INFLUENCE OF WATER ACTIVITY

During the last three decades water activity, a_w , has played a major role in many aspects of food preservation and processing. It is defined as the ratio of the vapor pressure of water P in the food to the vapor pressure of pure water P_0 at the same temperature ($a_w = P/P_0$). Next to temperature, it is now considered as probably the most important parameter having a strong effect on deteriorative reactions. The effect of water activity was studied not only to define the microbial stability of the product but also on the biochemical reactions in the food system and its relation to its stability. It has become a very useful tool in dealing with water relations of foods during processing.

It is now well known that microorganisms cannot grow in the dehydrated food system when the water

activity range is less than or equal to 0.6–0.7, but other reactions, enzymatic and nonenzymatic (e.g., lipid oxidation, NEB, etc.) that cause change in color, flavor, and stability continue during processing and storage. Water activity has become the most useful parameter that can be used as a reliable guide to predicting food spoilage or to determine the drying end point required for an SSP.

The relationship between equilibrium moisture content and water activity, known as the sorption isotherm, is an important characteristic that influences many aspects of dehydration and storage. It can be constructed graphically or derived mathematically. The shape of the isotherm generally determines the storage stability of the dehydrated product. This concept is used to establish product specifications for the effective drying, packaging, and storage of foods.

Adsorption isotherms of potatoes were of sigmoid shape and were affected by drying method, temperature, and addition of sugar [85]. The freeze-dried product absorbed more water vapor than the vacuum-dried materials. The sorption isotherm prepared from fresh and freeze-dried Thompson seedless grapes indicated a hysteresis loop at both the upper and lower moisture level [86]. The isotherm sun-dried grapes were slightly lower than that of vacuum-dried grapes.

Both lipid oxidation and NEB are greatly influenced by a_w [87]. Autoxidation of lipids occurs rapidly at low a_w levels, decreasing in rate as a_w is increased until in the 0.3–0.5 range and increasing thereafter beyond $0.5a_w$. Most rapid browning can be expected to occur at intermediate a_w levels in the 0.4–0.6 range. Whether or not it is minimized at the lower or upper portion of this range depends significantly on the specific solutes used to poise a_w , the nature of the food (especially amino compounds and simple sugars that might be present), as well as the pH and a_w of the product. Interestingly, at a_w levels that minimize browning, autoxidation of lipids is maximized.

The kinetics of chlorophyll-*a* transformation was studied as a function of time at different water activities at 38.6°C (Figure 25.8) [74]. For $a_w > 0.32$ the most important mechanism of chlorophyll degradation was the transformation into pheophytin; this had a first-order dependence on pH, water activity, and pigment concentration.

Carotenoids in freeze-dried carrots were relatively more stable in the range of 0.32 to $0.57a_w$ [71]. The maximum stability was near $0.43a_w$ (corresponding to an equilibrium moisture content of 8.8–10%). Increase in the rate of carotenoid destruction was greater at lower a_w than at higher a_w .

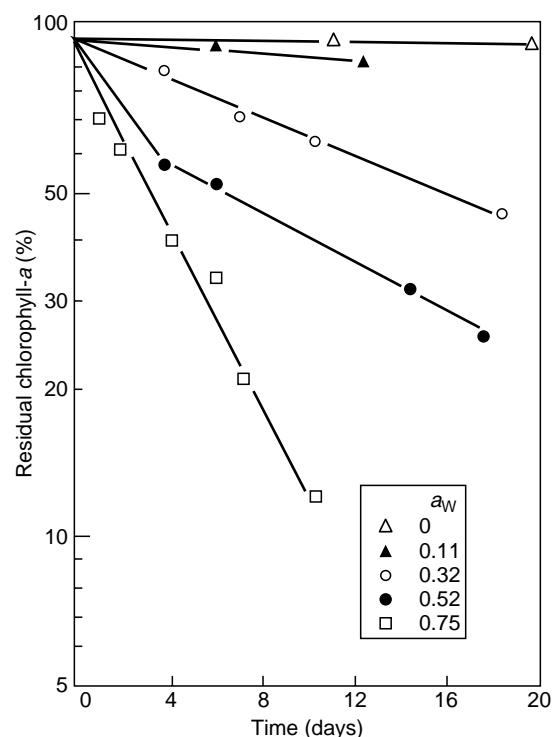


FIGURE 25.8 Degradation rate of chlorophyll-*a* in spinach as a function of time at different water activities (pH = 5.9; temperature 38.6°C). (From Lajolo, F.M. and Marquez, U.M.L., *J. Food Sci.*, 47, 1995, 1982. With permission.)

The kinetics of quality deterioration in dried onion flakes (NEB and thiosulfinate loss) and dried green beans (chlorophyll-*a* loss) were studied as a function of water activity and temperature and empirical equations and mathematical models developed that successfully predicted the shelf life of the dried products as a function of temperature and a_w (Table 25.11 and Table 25.12) [88]. Above the monolayer (a_w , 0.32–0.43) for onion, increasing moisture contents resulted in greater reaction rates for browning and thiosulfinate loss. Very little browning was observed over a storage period of 631 d at 20°C and $a_w = 0.33$, whereas all other samples stored at 30 and 40°C and $a_w = 0.43$ and 0.59 deteriorated to unacceptable levels within this time period. Similarly, in the case of green beans, the destruction of chlorophyll-*a* (pheophytinization) was found to be the principal factor responsible. The dried green beans were considered unacceptable when more than 30% loss of chlorophyll-*a* was observed the concentration at which the dull olive-green color began to predominate. Since conversion of chlorophyll-*a* to pheophytin is an acid-catalyzed reaction, the availability of water was essential and therefore a_w could be expected to influence the rate of chlorophyll loss.

TABLE 25.11
Actual (and Predicted) Shelf Life (Days) of Dried
Onion Flakes Based on Browning and Thiosulfinate
Loss at Different Temperatures

a_w	Browning			Thiosulfinate Loss		
	20°C	30°C	40°C	20°C	30°C	40°C
0.32	>631 (4778)	474 (472)	59 (63)	>631 (1619)	369 (306)	66 (55)
0.43	593 (600)	83 (69)	22 (21)	631 (585)	136 (139)	40 (38)
0.56	183 (190)	31 (33)	17 (17)	298 (288)	84 (82)	27 (29)

Source: From Samaniego-Esguerra, C.M., Boag, I.F., and Robertson, G.L., *Lebensml-Wiss. U.-Tech.*, 24(1), 53, 1991. With permission.

25.5.7 GLASS TRANSITION TEMPERATURE RELATED CHANGES

Glass transition is a second-order phase transition that occurs over the temperature range at which amorphous solid materials are transformed into viscous, liquid state [89]. The amorphous state of foods may result from a rapid removal of water from food solids that occur during such processes as extrusion, drying, and freezing. The temperature, water content, and time-dependent changes, which are the problems in manufacture and storage of powders and other low moisture foods, can be reduced by not exceeding their critical values based on T_g determination [90]. The T_g can be applied in evaluating proper temperature and

TABLE 25.12
Actual (and Predicted) Shelf Life (Days) of Dried
Green Beans Based on Chlorophyll-a Loss at
Different Temperatures

a_w	Temperature (°C)		
	20	30	40
0.32	>637 (962)	273 (282)	86 (84)
0.43	478 (452)	143 (146)	45 (38)
0.56	150 (148)	61 (56)	25 (26)

Source: From Samaniego-Esguerra, C.M., Boag, I.F., and Robertson, G.L., *Lebensml-Wiss. U.-Tech.*, 24(1), 53, 1991. With permission.

humidity conditions of agglomeration and in reducing quality changes occurring with dehydration.

The collapse of the dehydrating foods during freeze-drying, stickiness of the product during spray drying, caking and agglomeration of the powders during processing, and storage are some of the properties that are related to glass transition temperature.

del Valte et al. [91] studied the relationship between shrinkage during drying and glass rubber transitions of apple tissue. Their work demonstrated that infusion of sugar during osmotic dehydration at high solute concentration brought about some protection against shrinkage. This was reflected by a 20–65% increase in volume of samples treated with 50% sucrose and maltose solutions as compared to air-dried control. However, reported data did not indicate that structural collapse could be reduced by diminishing the difference between drying temperature and glass transition temperature. Dried samples remained in the rubbery state and shrunk during subsequent storage.

25.5.8 MICROBIOLOGICAL ASPECTS

Drying is the oldest method of preserving food against microbiological spoilage. Since presence of water is essential for enzymic reactions, the removal of water prevents these reactions and the activities of contaminating microorganisms present. Removal of water increases the solute concentration of the food system and thus reduces the availability of water for microorganisms to grow. There is a lower limit of water activity for specific microorganisms to grow; for complete microbiological stability, water activity of the system should be below 0.6.

Drying, however, is also an effective means of preserving microorganisms in a viable state, even though their numbers may be reduced and a proportion sublethally damaged [92]. Survival during and after drying will depend upon the physicochemical conditions experienced by microorganisms, such as temperature, a_w , pH, preservatives, oxygen, and so on. The survival of food spoilage organisms may give rise to problems in a reconstituted food item, but survival of foodborne pathogens must be viewed much more seriously.

With a view to minimize organoleptic changes in foods during drying, time and temperatures are kept as short and as low, respectively, as feasible. The process of drying, whether by freeze-drying, hot air drying, solar drying, or by high temperature (e.g., spray or drum drying) is not *per se* lethal to all microorganisms and many may survive. The more heat-resistant organisms are the more likely survivors (e.g., bacterial spores, yeasts, molds, and thermophilic bacteria). Thus there is a strong possibility for microbial growth,

including pathogens, before the a_w of the product falls below the critical level for each organism.

Vegetables, because of their greater proximity to soil and lower acidity and sugar content as compared with fruits, predominately have more bacterial populations. A majority of the species has been found to be common for soil- and waterborne bacteria of the genera *Bacillus* and *Pseudomonas*. Some workers have found other types of bacteria such as coliforms and bacteria of the genera *Achromobacter*, *Clostridium*, *Micrococcus*, and *Streptococcus* from different dehydrated vegetables.

Factors that influence markedly the microbial population of dehydrated vegetables include the microbial quality of fresh produce; the method of pretreatment of the vegetables (peeling, blanching, etc.); the time elapsed between preparation of the vegetables and start of the dehydration process; the time involved in the dehydration of the vegetables; the temperature of dehydration; the moisture content of the finished product; and the general level of sanitation in the dehydration plant [93]. Blanching, if sufficient to inactivate enzymes, would reduce the contamination of the fresh produce to an insignificant figure. Reduction in total count during blanching was found to be greater than 99.9%.

Coliforms and enterococci are commonly used as indicators of unsanitary conditions in food processing. Clarke et al. [94] isolated enterococci from 18 out of 35 dehydrated vegetable samples. They found coliforms in 18 and enterococci and coliforms in 15 samples. Statistical analysis showed a positive correlation between number of enterococci and coliforms. The predominant species recovered from enterococci was *Streptococcus faecium* (60%) and from coliforms was *Aerobacter* (56%).

Fanelli et al. [95] surveyed a number of commercially available vegetable soups and found that the maximum total number of bacteria was less than 50,000 per gram and the mean and median total bacterial numbers were very low. The numbers of coliforms, yeasts, molds, and aerobic species were also low.

Dehydrated onion, which is an important commercial flavoring ingredient, is not blanched before dehydration. Its microbiology was therefore extensively investigated. Total plate count (TPC) was less than 100,000 per gram in 76% of the slices from the belt dryer and only 52% of the sample in the tunnel-dried product [96]. In both cases, the average bacterial spore count was 12,000 per gram. Many workers have variously reported the presence of *Bacillus*, *Pseudomonas*, *Aerobacter*, *Lactobacillus*, and *Leuconostoc* species [97]. Exposure to ethylene oxide gas was found to be effective in reducing the relatively high

TPC, but future application of this gas is in doubt because of its toxic hazards. Alternatively, the application of gamma radiation at a level 0.2–0.4 Mrad was suggested to sterilize onion powder without any detrimental effect.

25.5.9 FACTORS AFFECTING STORAGE STABILITY

The shelf life of dehydrated fruits and vegetables depends on many deleterious reactions, which in turn depend on the specific nature of the food materials, storage conditions, and nature of packaging. The undesirable changes that occur are due to off flavors, browning, and loss of pigments and nutrients as enumerated above. Knowledge of the causes of these reactions is highly necessary to improve the shelf life of the dehydrated products.

Villota et al. [98], in their review on the storage stability of dehydrated foods, discussed the factors mainly responsible for deterioration, that is, moisture, storage temperature and period, oxygen, and light. They compiled the literature data on storage stability of several dehydrated products, which included dehydrated fruits, vegetables, and fruit and vegetable powders, based on method of drying, additional treatment, storage conditions, time required for appearance of earliest defects, and the state of other factors at times of unacceptability.

Moisture content is a very important parameter influencing the stability of dehydrated foods. It has been suggested that the optimal amount of water for long-term storage corresponds in most dehydrated foods to the Brunauer–Emmett–Teller (BET) monolayer value. On the other hand, items such as freeze-dried spinach, cabbage, and orange juice were reported to be more stable at a zero moisture content, whereas items like potatoes and corn had maximum stability at the monomolecular moisture content. It appeared that optimal moisture content could not be predicted with precision on the basis of theoretical considerations.

Another important factor affecting storage stability of dehydrated foods is temperature and period of storage. Generally, the storage stability bears an inverse relationship to storage temperature, which affects not only the rate of deteriorative reaction (enzyme hydrolysis, lipid oxidation, NEB, protein denaturation), but also the kind of spoilage mechanism.

It is well established that elimination of oxygen by packing in an inert atmosphere such as nitrogen contributes to extending the storage stability of many dehydrated products. However, in certain products like spray-dried powders, in which a large surface area is exposed to air during processing, some entrapment of oxygen occurs in the final product and

packing under inert atmosphere results in a very little improvement. Storing in zero oxygen headspace, using an atmosphere of 5% hydrogen in nitrogen with a palladium catalyst, is reported to result in superior quality retention. Further, since oxidation of lipids and vitamins like ascorbic acid, riboflavin, thiamine, and vitamin A and loss of pigments such as carotenoids and chlorophyll are initiated or accelerated by light, adequate packaging needs to be provided to protect such dehydrated foods from light.

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26 Drying of Herbal Medicines and Tea

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26.1 INTRODUCTION

Herbal medicine is part of human civilization. It has been used in China for nearly 5000 y. One of the oldest and most important documents is the Egyptian Ebers papyrus (ca. 1550 BC), which includes more than 700 prescriptions using natural products such as caraway, coriander, garlic, linseed, peppermint, figs, fennel, anise, poppy, and castor oil [1]. *Shen Nong's Materia Medica* was compiled by ancient Chinese in about 200 BC, which described the properties and usages of 365 types of Chinese medicines in three categories. In ancient Greece there was a guild of rhizomatists or root collectors, who gathered, prepared, and sold medicinal plants. The Greek botanist and physician, Dioscorides (AD 40–90), compiled the first systematic description of 579 plants and their 4700 medicinal uses and modes of action. His work titled *De Materia Medica* was of central importance to European medicine until the 17th century [1]. Ayurveda is the principal traditional medical system of

India, Pakistan, Nepal, and Sri Lanka, which has also influenced medicine in Tibet, Burma, and Malaysia.

No doubt that synthetic drugs have played a vital role in the enhancement of human living standards during the past century; herbal medicine has regained its momentum once again in recent years. Long-term ailments cannot typically be cured by injection or consumption of a single medicine. Instead of focusing only on curing an illness, people are paying more attention nowadays to improve the whole body immune system so as to prevent the attack of diseases. Besides their traditional pharmaceutical usage, herbs have also become one of the important sources for drug discovery and production. Herbal medicine plays an important role in healthcare in many regions of the world. The combined sales of herbal medicine products in major markets around the world exceeded US\$12 billion in 1994, [Table 26.1](#) [2]. According to the report from Herbal Medical Database Ltd., the US market for herbal medicines has been estimated to

TABLE 26.1
Global Market for Herbal Medicine Products in 1994

Region	Annual Herbal Product Sales, in Millions US\$
European Union	6000
Rest of Europe	500
Asia	2300
Japan	2100
N. America	1500

be worth US\$ 58 billion in 1999 with an annual growth rate of 25%.

There are more than 1000 types of herbs in use around the world as medicines, spices, flavors, etc. On the basis of their biological complexity, they can be classified as algae, fungi, liverworts and mosses, ferns and fern allies, seed-bearing plants, and higher flowering plants [3]. Depending on their physical properties, one can further classify them as follows: sticky, aromatic, powdery, oily, and lustrous. This latter classification determines the way the herbs are handled. The sticky herbs usually contain significant amount of sugar, e.g., *Radix Asparagi*, *Rhizoma Polygonati*. The aromatic herbs are known by their special scents. Typical aromatic herbs are: *Herba Schizonepetae*, *Herba Menthae* (also known as mint), *Herba Elsholtziae*, and *Rosae banksiae*, etc. The powdery herbs contain large amounts of starch such as *Rhizoma Dioscoreae* or rhizome of common yam. *Radix Angelicae sinensis* and *Rhizoma Ligustici chuanxiong* are two oily herbs. For lustrous herbs, such as *Radix Platycodi*, *Rhizoma Alismatis*, and *Astragalus membranaceus* the brightness of their surfaces is quite important [4].

The quality of the herbs depends very much on the contents of active ingredients. It is known that heating or thermal drying, if not carried out properly, can cause a significant loss of the active ingredients. The drying methods described below therefore should be adopted with care. For a specific herb species, its traditional way of drying should be analyzed carefully before any alternative methods are employed. The comparison of different drying methods on the quality of herbs has to be carried out based on physical and chemical analyses. Various countries have established standard testing methods for herb examination with known indicators for given species, for example, *Species Systematization and Quality Evaluation of Commonly Used Chinese Traditional Drugs* [5] and *Handbook of Composition and Pharmacological Action of Commonly Used Traditional Chinese Medicine* [6] are available in China.

A large assortment of dryers can be used for drying herbs and medicinal plants. The reader is referred to [Chapter 1](#) of this handbook for information on selection criteria for dryers.

26.2 DRYING OF HERBS AFTER HARVESTING

26.2.1 HERBS IN SMALL QUANTITY

Herbs are very delicate plants. Their efficacy depends on species, parts, planting location, harvest time, drying method, and storage. For a specific species, the medicinal effect may derive from seeds, flowers, leaves, stems, or roots, or all of these. Herbs were traditionally harvested in small quantities from the mountains or fields by individuals. Besides the active ingredients, fresh plant material contains a high portion of water. Leaves and flowers usually lose up to 85% of their weight on drying. The fresh plant material is spread out in thin layers (or in certain cases hung up in bunches) and kept in a dry, well-ventilated place. Tubers and roots take longer to dry than flowers and leaves even though the former are often cut up into pieces. The selection of proper drying temperature is vital. Too high a temperature may cause loss of active ingredients. On the other hand, too low a temperature may actually accelerate decomposition by promoting enzyme activity within the plant itself or even microbial attack, particularly for sugar-containing substances.

Herbs are usually harvested when the flowers are just coming into bloom, as they are then richest in aromatic oils. As a guideline, one should pick herbs just after the dew has gone and discard any yellow or damaged herbs. Herbs should be handled with care and only rinsed to remove obvious dust or soil. Herbs may be dried in a ventilated room or in a barely warm oven, leaving the oven door open. The temperature should not exceed 34°C. The following procedure is suggested: lay the herbs on wire cooling racks covered with muslin, cheesecloth, or nylon net. When dry, the herbs are brittle and crumbly. Put the dried herbs into storage jars, preferably of tinted glass, and cover with a plastic screw cap. Appearance of signs of condensation inside the jar, which implies that the herbs are incompletely dried and hence should be returned to the drying cupboard or oven. Long-stemmed herbs may be dried by hanging them in a warm, dry, airy place for a few days. They may be tied in small bunches in a loose fashion. Cover the bunches with dark paper if direct exposure to sunlight may occur [3]. It should be pointed out that some herbs are dried under direct sunshine, for example,

Caulis Ephedrae, *Radix Polygalae*, *Radix Astragali*, *Radix Glycyrrhizae*, Shiitake mushrooms, etc. Microwave oven has been used recently for drying garden herbs. For microwave drying, place a single layer of herb leaves between double thickness of paper towels. Dry them for 1–2 min on a medium to high setting (or half of the power), depending on the thickness of the leaves. Flip over and repeat for one more minute [7]. Pretests with grass are suggested if one is not familiar with the microwave oven drying practice.

Catalytic Drying Technologies has introduced a technology capable of significantly reducing the moisture content of a wide range of agricultural products, quickly and without causing product damage or generating regulated emissions. Tests have proven this technology capable of reducing moisture content from as high as 60% to levels between 3.4 and 4.4%. The catalytic infrared system (burning gas without flame) dries rice quickly and uniformly in a continuous 3-h cycle. Low operating temperatures minimize energy consumption; patented material handling and agitation technology maximize efficient and uniform drying [8]. The heat pump dryer has been investigated and found to be useful in herb drying. It dehumidifies the drying air, heats it, and leads it back for recirculation. This type of dryer can work at low air temperature; thus color and active substances remain intact. Another feature of the heat pump dryer is its characteristic low energy consumption [9,10]. Silica sand drying can be used when the herbs are not used as cooking recipes. Freeze drying can also be used if it is important to retain the color of the herbs and the active ingredients are highly heat-sensitive. For a small quantity, simply place the fresh herbs inside the freezer, which allows dehydration to take place via sublimation. Freeze drying is an expensive process recommended only for high-value, low-volume products.

26.2.2 DRYING OF TEA LEAVES

Tea is an ancient crop that has been cultivated for thousands of years. Its exact origins have not been confirmed by historians. There are many theories about the discovery of tea. According to tradition, tea was discovered by the Chinese Emperor, Shen Nong in 3000 BC. The first records of tea date from

the 4th century AD in China. Tea leaves are probably one of the earliest and most consumed herbs. They were initially found to have the property to counteract the poisonous effects derived from other herbs. Nowadays, tea has become one of the everyday drinks next to coffee and cocoa. The main tea producers are India, China, Sri Lanka, Kenya, Malawi, and Indonesia. There are more than 1500 teas to choose from, originating from more than 29 different countries [11]. The majority of the teas can be classified as green tea, oolong tea, and black tea. Each requires a different process and hence a different drying system is required. The quality of dried tea depends to a great extent on the drying system and operating conditions used.

26.2.2.1 Green Tea

Green tea is drunk mainly in China, Japan, and some parts of South America. Figure 26.1 shows the process of a typical green tea production before final sorting and packaging [12]. The initial water content of fresh tea leaves after plucking is around 75–78% (wet basis) in spring and 65–70% (wet basis) in autumn [13]. The withering process is accomplished by laying the tea leaves on trays or racks in the shade at 25–30°C for a few hours depending on how wet the leaf is. This process prepares the tea leaves for rolling without losing juice. Meanwhile, the water content falls to about 50% (wet basis). The kill out process can be achieved by pan frying, steaming, or baking. This process is designed to arrest the enzymatic reaction, oxidation in particular. Pan drying usually employs a wok at a surface temperature of 400–470°C (preferably 430–460°C) to reduce the leaves' moisture content to 10–15%. The drying step starts with using hot air at 110–120°C to evaporate water in a layer of tea leaves about 20-mm thick. In practice, air temperature up to 150°C can be used. It is recommended to have the dried tea leaves tempered to prevent the edges of leaves from “crisping.” Then the next drying starts with the wok temperature at 150–160°C and drying time of 30–40 min until moisture content reaches about 20%. Afterwards, the wok surface temperature is reduced to 80–100°C and drying continues for 60–90 min to have the moisture content of 9–10%. Finally, the wok surface temperature is dropped to

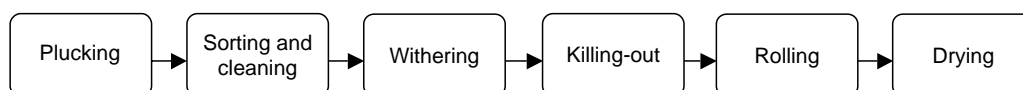


FIGURE 26.1 Production of green tea.

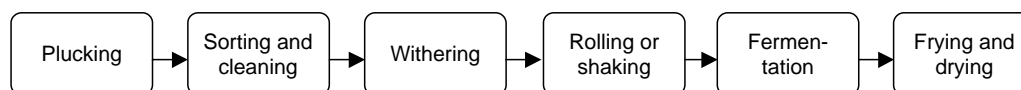


FIGURE 26.2 Production of oolong tea.

60°C and drying carried out for 60–90 min until the final moisture content of 4–5% is reached [13].

26.2.2.2 Oolong Tea

The process of oolong tea production is shown in Figure 26.2. The freshly plucked shoots from *Camellia sinensis* are spread out thinly over a cleared area of flat ground, which is usually covered with a mat or a towel to keep leaves from contact with the earth. The shoots are wilted in air under the sun for 30–60 min, depending on the temperature. The leaves are then taken indoors to wither further at ambient temperature for a few hours. This withering process can also be assisted by heated air at <40°C to shorten the process time. During this process, the leaves are gently agitated by hand every hour. This process causes the edges of the leaves to turn red [14]. During this process, there is about 20% decrease in moisture content of the tea leaves. This step is accompanied by “fermentation” or curing to produce the unique aroma and color found in oolong teas. To arrest the enzymatic reaction so as to stop the leaves from getting dark, the leaves are dried for 5–7 min at a temperature slightly lower than that required for “killing-out” in green tea production. Afterwards, the tea leaves are dried in a few steps in heated air with rolling in between the drying steps. The first hot air drying in conveyer belt or an oven is usually achieved at a temperature around 125°C with the depth of leaves 10–20 mm for 7–10 min. At the end of this step the moisture content is about 40–45% (wet basis). The second drying step is completed with air temperature of 90–100°C for another 7–10 min until the moisture content nearly reaches its equilibrium level. A third drying step may be used at air temperature of 80–90°C and a bed depth of 20–30 mm for 20 min. The second and third drying steps may be combined using air temperature of 80–90°C. The depth of leaves can be a bit higher depending on the location of the tea plants [13].

26.2.2.3 Black Tea

The process of black tea production is shown in Figure 26.3. Tender young growth is picked by hand from *Camellia sinensis*. Any surface water on the leaves and shoots is allowed to dry on racks for 10–20 h to bring down the internal moisture content so as to make the leaf more pliable for the next step. The light, medium, and heavy withering can be adopted with the moisture content decreases in 10, 15, 20% (wet basis), respectively. The selection of degree of withering depends on the downstream processes and the final product specification [13]. The leaves are bruised to allow the fermentation process to begin. Cutting is done at this point if necessary. The rolling process will last until the leaves turn shining dark red like a bright copper penny. The leaves are allowed to ferment by placing thin layers on a tray in a shady location for 2–3 days before drying [14]. If a fermenter is used for the cut leaves, the temperature should be controlled around 32°C and time within 90 min. When moist air of 20–26°C is blown through the leaves, the fermentation time should be below 60 min [13]. The drying can be accomplished in an oven set at 110–120°C for 12–16 min to obtain 18–25% water content in the leaves of depth 15–20 mm. Further drying is achieved by having oven temperature of 90–95°C for another 12–16 min to the final moisture content of 5–6% (wet basis). The higher temperature step is necessary to stop the enzymatic fermentation reaction and seal the flavors inside the leaves.

26.2.3 SOME TYPICAL DRYERS FOR TEA LEAVES

26.2.3.1 Multitray Oven

This type of dryer is usually designed for specialty tea processing. It is an oven with about four layers of trays. The drying surface area is 1–2 m². Electrical heating or heated air can be the source of drying energy. The wet tea leaves are placed on the top layer tray. The product is obtained in the bottom layer tray. The moving of the trays is accomplished manually

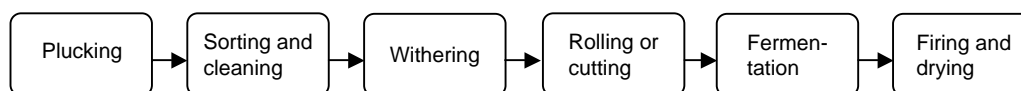


FIGURE 26.3 Production of black tea.

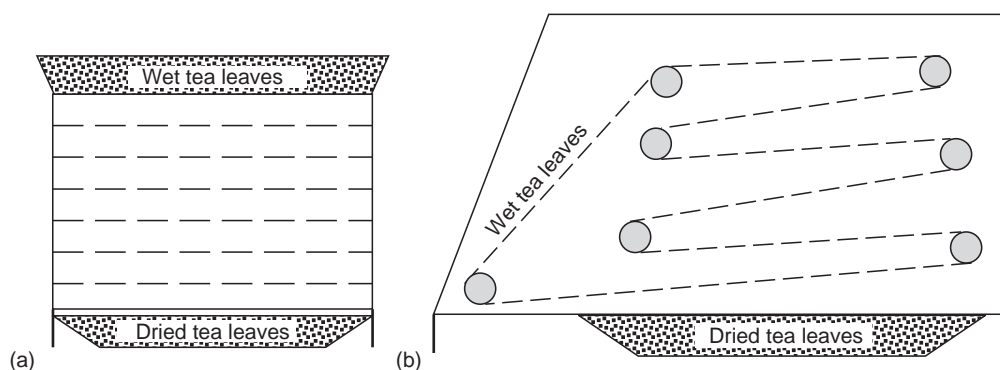


FIGURE 26.4 Louver-type oven: (a) manual operation; (b) continuous.

usually from top to bottom as drying proceeds. To arrest the fermentation quickly, heated air can be blown individually into each layer [13].

26.2.3.2 Louver-Type Oven

This dryer is similar to that of multitray oven. The main difference is that there are usually six layers of louver plates. The size of this dryer can vary according to the capacity requirement. Again the wet leaves are loaded at the top. As drying proceeds, the louvers are flipped manually to turn the leaves to the next layer below. The flipping is done in alternative directions so as to accomplish a more uniform drying. Heated air is blown from one side of the dryer. The dried product is collected at the bottom of the oven. This dryer can also operate in the continuous mode. The wet tea leaves are fed onto the continuous moving louver plates. The louver plates make five turns inside the drying chamber before sending out the dried product to the bottom. While moving, the louvers are flipped according to the schedule to have a more uniform drying. This type of dryer can have a drying area of 50 m². A schematic diagram of this type of oven is shown in Figure 26.4 [13].

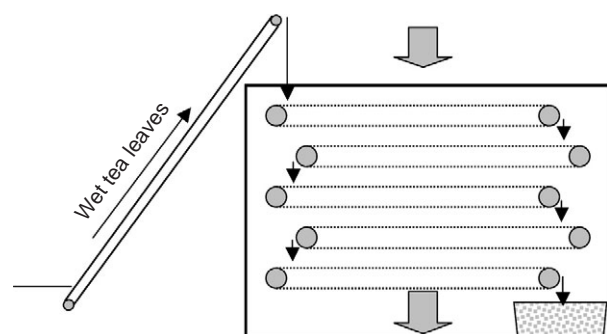


FIGURE 26.5 Conveyor belt dryer. (From Hu, J., *Drying of tea*, in *Modern Drying Technology*, Eds. Y.K. Pan and X.Z. Wang, Chemical Industry Press, Beijing, 1998, pp. 788–809.)

26.2.3.3 Conveyor Belt Dryer

Figure 26.5 shows how a conveyor dryer operates. Instead of louver plates, a net belt is used in this dryer. Superheated steam is used as the drying medium. The drying area can be 20 m². It is good for drying small pieces of herbs and vegetables. Electrical heating can also be used as an auxiliary source of thermal energy.

26.2.3.4 Fluidized Bed Dryer

Figure 26.6 shows the schematic layout of a fluidized bed dryer [13]. This type of dryer is good for drying cut leaves in black tea production. It can give a uniform drying in a relatively short period of time. Mobile bed in combination with vibrating fluidized bed may also be used for this purpose. A fluid bed dryer consists of a grid through which hot air is forced out. Fermented leaf with not too high moisture content is suspended by the air streams, resulting in fluidization. The air also acts as a carrier of the particles through the dryer, making the bed of leaves move forward until the dried leaves are discharged. To facilitate fluidization of wet fermented leaf particles, the grid can be vibrated with half-amplitude of several millimeters and frequencies in the range of 10–20 Hz using metal springs to make a vibro-fluidized bed dryer, Figure 26.7 [15]. This type of dryer for tea leaves has become very popular over the past three decades. It is claimed to produce less fines and result in gentler handling of the fragile leaves resulting in a better quality product.

The fluidized bed dryer essentially consists of a drying chamber, plenum chamber, dust collectors, and flow control dampers. The drying chamber normally consists of three drying zones and one cooling zone. Fermented leaf is loaded on the grid plate of the drying chamber. The top of the drying chamber is closed and two sets of centrifugal exhaust fans

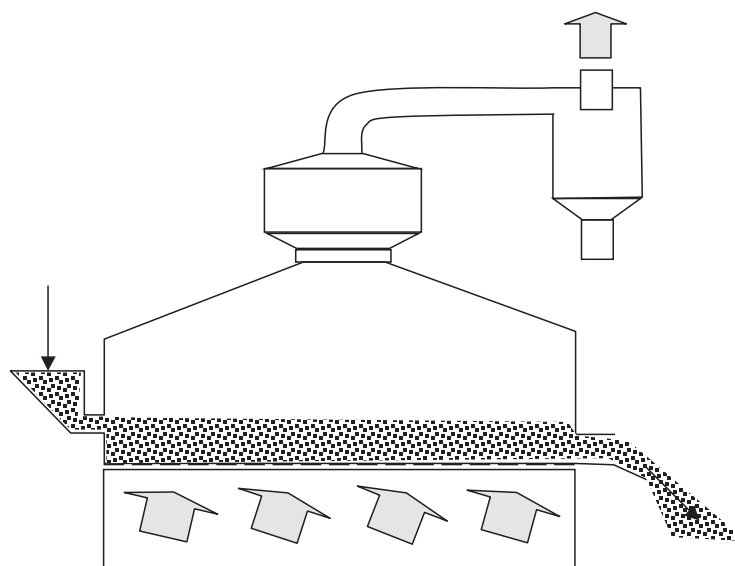


FIGURE 26.6 Fluidized bed dryer. (From Hu, J., Drying of tea, in *Modern Drying Technology*, Eds. Y.K. Pan and X.Z. Wang, Chemical Industry Press, Beijing, 1998, pp. 788–809.)

provided with cyclones—one for refiring and the other for dust extraction. In the first zone of the dryer, the air flowrate is the maximum to remove the very high moisture content of the fermented leaf rapidly. As the moisture content is reduced, the density of the particles is also decreased. These lower moisture content leaves tend to move away from the feed end and are replaced by a new load of fermented leaf particles. When the tea leaves are fully dried, they are expelled into a cooling chamber where ambient air is introduced [15]. The hybrid dryer consisting of tray dryer and fluidized bed dryer may also be used to reduce the temperature of the hot air in the fluidized bed and also to save on energy consumption.

26.2.4 DRYING OF GINSENG ROOTS

Ginseng is the most commonly used medicinal plant in Asia. Recently, this ancient, cultivated plant has been rediscovered in the Western world as a remedy with numerous and diverse benefits. Traditionally, ginseng was dug from the forest in the mountains. It

is a very delicate herb that usually takes 6 y or much longer time to grow. In fact, high-quality roots should be around 15 y or older. The harvested ginseng roots are usually sun-dried [16].

Because of its economic values, ginseng roots are now cultivated in China, Korea, North America, and even Australia. The majority of ginseng root consumed in the world today first needs to be dried. The fresh root moisture content is around 72% (wet basis). The dried ginseng has a moisture content of approximately 8–12% (wet basis) [17]. Good quality dried roots should have final moisture content from 5.5 to 7.5% depending on the size of the roots [18]. Drying is critical for quality ginseng root. Heat damage or fungal development can occur from too short or too long a drying process. Roots should be sized and similar sizes dried together. The drying of ginseng root can be achieved by placing them on wire racks or bamboo baskets and leaving them under the sun as done traditionally. It will take a few weeks for the roots to dry. They are more often dried inside by stacking the wire racks in a room with temperature

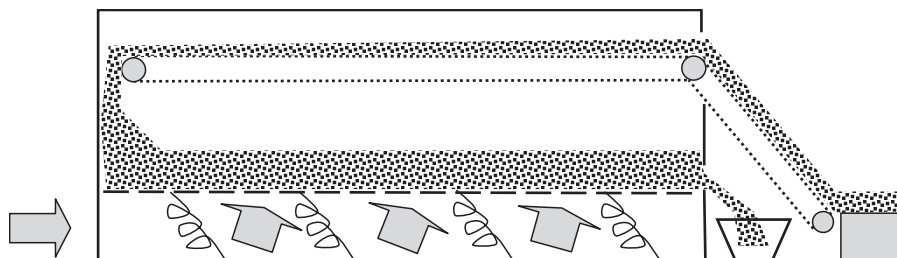


FIGURE 26.7 Vibro-fluidized bed dryer. (From Hu, J., Drying of tea, in *Modern Drying Technology*, Eds. Y.K. Pan and X.Z. Wang, Chemical Industry Press, Beijing, 1998, pp. 788–809.)



FIGURE 26.8 A shed dryer for ginseng roots. (From Brun, C.A., Ginseng, Cooperative Extension Program, Washington State University, 2004.)

maintained around 38°C. This drying process will take about 2–3 weeks to complete. The dried product will have a brittle exterior and creamy-white interior [17]. The rough quality control is carried out by snapping tests. A modified tobacco kiln dryer can be used for large volumes of ginseng. Drying cabinets can also be employed with the air circulation and dehumidification systems equipped.

Temperature management is important in ginseng drying. Too low a temperature will result in the green coloring or molding, whereas too high a temperature will give a product that is too hard (not easy to snap) and of a brown color indicating the formation of caramelized sugars. The danger of brown roots developing is greater toward the end of drying. It is not recommended to accelerate the drying by increasing the drying temperature at the last stage of drying. Generally, drying is begun with temperatures at least above 30°C and temperatures below 38°C are used after root wilt. Using higher temperatures at the beginning of the cycle will reduce the drying time but it has to be ensured that the root temperatures throughout the dryer are always within the safe limit. Three or four thermometers should be placed inside the drying shed to monitor the temperatures. The floor temperature should also be monitored where the value is the lowest. Heating of the shed can be achieved by natural gas or electricity. Electric heated walls have the advantage of not introducing any odor from burning of natural gas.

Figure 26.8 shows a photo of a shed dryer for ginseng roots. The drying history of different sized ginseng roots is illustrated in Figure 26.9.

Large root refers usually to those of larger than 17 mm in diameter. The diameters of medium-sized roots are from 12 to 17 mm. The small roots are those of 7–12 mm in diameter. Any root with diameter below 7 mm is classified as a fiber [18].

In a shed-type dryer the moisture laden, cooler air will sink to the lower levels of the room. Thus, the upper trays will tend to dry faster. To maintain a uniform drying, either the upper trays are loaded with large roots or the trays are rebuilt during the course of drying to move the upper trays down and lower trays up. During the first week of drying, the air inside the room should be refreshed every 10–15 min using an exhaust fan. The capacity of a shed dryer can vary from 125 to 200 kg fresh root per sq. m. If higher

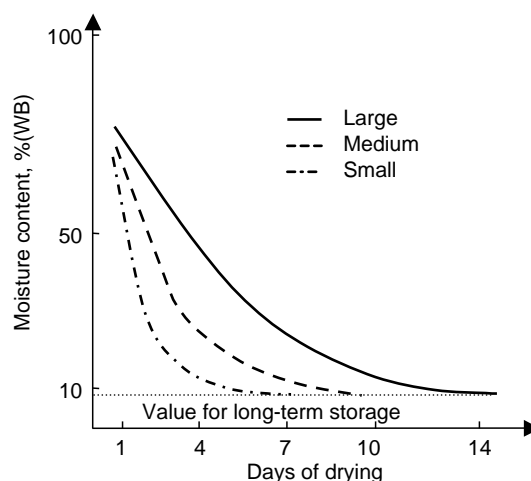


FIGURE 26.9 Hypothetical drying curves of ginseng roots.

TABLE 26.2
List of Experimental Data of Some Herbs Dried by Convective Flow Heated Air

	Name	Allowed T , °C	Actual Maximum T , °C	Ingredient Loss, %	Maximum ΔT , °C	Air Velocity, m/s	T_{air} , °C	Initial Moisture Content, kg Water/kg Solids	Mean Drying Rate, g/kg-min
1	<i>Cortex Mori Raicis</i>	70–80	79	–14.9	20	1.1	70 → 80	0.773	9.9
2	<i>Fructus Aurantii seu Ponciri</i>	50–60	63	1.4	11	0.7	20 → 75 → 84 → 68 → 47	0.75	7.77
3	<i>Herba Caricis phacotae</i>	70–80	67	100	20	1.0	70	1.00	7.70
4	<i>Herba Menthae</i>	Ambient	68.5	100	35	1.3	70	0.80	6.8
5	<i>Polyporus umbellatus</i>	70–80	69.6	16	7	0.9	20 → 70 → 54	0.88	6.76
6	<i>Radix Trichosanthis</i>	70–80	71.2	66.7	28	0.9	70	0.87	6.7
7	<i>Cortex Moutan Radicis</i>	50–60	68	2.7	20	1.0	70	0.60	6.5
8	<i>Radix Paeoniae alba</i>	70–80	69	60	16	1.1	70	0.70	6.5
9	<i>Citrus reticulata</i>	50–60	66	0	10	0.5	20 → 74 → 26	0.28	6.1
10	<i>Curcumae aromatica</i>	70–80	68	61.8	14	0.9	70	0.90	5.42
11	<i>Astragalus membranaceus</i>	70–80	68	0	15	0.8	50 → 70 → 30	0.53	5.36
12	<i>Cortex Magnoliae Officinalis</i>	70–80	72	2.8	20	0.5	50 → 76 → 70	1.03	5.22
13	<i>Atractylodes macrocephala</i>	70–80	72	2.8	20	0.5	50 → 70 → 88	0.74	4.77
14	<i>Cortex Phellodendri</i>	70–80	49.1	17.4	16	1.2	50	0.40	4.6
15	<i>Radix Platycodi</i>	70–80	68	33.6	18	0.8	40 → 74 → 68	0.29	4.5
16	<i>Rhizoma Ligustici chuanxiong</i>	50–60	48	22	6	0.8	57 → 50	0.923	4.5
17	<i>Semen Arecae</i>	70–80	69	0.5	28	1.3	50 → 70 → 68	0.80	4.4
18	<i>Radix Saposhnikoviae divaricatae</i>	70–80	69	41.3	20	0.8	52 → 72 → 66 → 69 → 50	0.34	4.3
19	<i>Radix Glycyrrhizae</i>	50–60	49	24.2	14	1.0	54 → 50 → 45	0.44	4.09
20	<i>Rhizoma Dioscoreae</i>	70–80	69.8	16.6	21	0.7	70	0.81	4.14
21	<i>Rheum Officiale</i>	70–80	70.2	43	19	0.95	50 → 80 → 53 → 62	1.12	3.86
22	<i>Radix Salviae</i>	70–80	73.4	33.7	20	0.8	60 → 67 → 62 → 40	0.64	3.78
23	<i>Radix Angelicae sinensis</i>	50–60	59	29.1	15	1.0	50 → 59.8	0.75	3.77
24	<i>Caulis Lonicerae</i>		71	28.6	36	1.2	70	0.72	3.72
25	<i>Atractylodes chinensis</i>	50–60	49.9	0	12	0.8	50	0.356	3.69
26	<i>Radix Isatidis</i>	70–80	43.9	59.2	8	0.8	55 → 59 → 56	0.29	3.1
27	<i>Rehmannia glutinosa</i>	50–60	62	11.4	20	0.92	50 → 71 → 50 → 60	1.08	2.74

TABLE 26.3
The Experimental Data of Drying of *Radix Glycyrrhizae*

	Temperature, °C			Drying Rate, g/kg·min		Drying Time, min
	Heated Air	Maximum at the Bottom Surface	Maximum Difference between Top and Bottom	Maximum	Average	
1	40	33.5	7.8	10.5	5.6	70
2	50	46	15	11.5	6.2	61
3	70	63	28	12.4	8.6	48
4	30 → 77 → 60	49	7.6	10.0	6.9	55

Air flow velocity = 1 m/s, initial moisture content = 0.44, final moisture content = 0.08, depth of bed = 30 mm, convective flow of heated air drying.

loading rates are required, a better-controlled drying room should be used.

Electrohydrodynamic (EHD) drying has been reported to work well for drying of root or sliced root materials [19]. This technique employs high voltage DC and sharp cathode to generate a corona to accelerate the air movement close to the surface of the wet material [20]. Because this drying technique requires only slightly heated air for drying, it gives very high quality of dried material. It is claimed as high quality as freeze dried materials are obtained [21,22]. The economics of this novel drying process are not clear as yet. For additional details, the reader may refer other sections of this handbook.

26.3 DRYING OF HERBAL MEDICINES IN SMALL PIECES

Before the herbs can be used as herbal medicines, further treatment is necessary. The dried herbs are usually collected by a herbal pharmaceutical company where they are washed or rewetted, cut, dried, and packaged for sale. The rewetting step is necessary to prepare the herbs for the cutting process. Drying is obviously very important in small pieces herbal medicine processing. Table 26.2 lists the effect of convective air drying on 27 types of herbs in small pieces, which were placed on a screen as heated air was blown through [23]. The drying chamber used in these tests was 1.03 m × 0.85 m × 0.85 m in cross-section. The weight loss of the material was obtained from the readings of an electronic balance. The depths of the beds of herbs were about 30 mm. Nine temperatures were monitored in the fixed bed of the herb pieces located at the center planes of the rectangular box. The maximum temperature in the material and also the maximum temperature difference in the materials are listed. Most of the drying was

conducted over 90 min with the final moisture content of 0.08 kg water/kg solids. The initial load of the wet material was about 400 g. The loss of active ingredients can be quite high even though the operation was conducted at a temperature supposed to be safe for these herbs [23]. More experiments by these authors show that neither the air flowrate nor the initial moisture content affect the quality of dried herb pieces. The quality of the dried herbs depends on the drying technology and drying conditions with temperature being a key parameter to control. It is anticipated that for some herbs the currently listed values of allowed temperature are simply too high.

In order to maintain a high drying rate and a low temperature of herb pieces, it is suggested to use staged drying with variation of temperatures of heated air. For example, in drying of 400 g of *Radix Glycyrrhizae* from 0.44 to 0.08 kg water/kg solids, it took 70, 61, 48, and 55 min, respectively, at temperatures of 40, 50, 70, and (30 → 77 → 60)°C. For the staged drying, the heating from 30 to 77°C takes 28 min and the cooling to 60°C takes about 27 min. Table 26.3 gives the temperature variation within the materials. The depth of the fixed bed is also a parameter to consider as demonstrated in the drying of *Cortex Mori Raicis*. When the bed depth is increased from 25 to 50 mm, the drying time is 2.7 times longer and within the material, the maximum temperature differences increased from 10 to 18°C at a drying temperature of 66°C [23]. The optimal depth for the experimental facilities described previously is about 30 mm.

When infrared drying is used, the loss of active ingredients is even higher for some herbs as shown in Table 26.4 [23]. In this drying experiment, the IR radiates sideways toward the fixed bed of herb pieces. The depth of the bed is 30 mm. The initial

TABLE 26.4
The Experimental Data of Infrared Drying of Herbs

	Name	Temperature, °C			Active Ingredients Tested/Method	Loss, %
		Allowed	Actual	Heating Plate		
1	<i>Cortex Mori Raicis</i>	80	78	292	Water extracts/standard	12.3
2	<i>Fructus Aurantii seu Ponciri</i>	60	62	279 → 212, $K_1 = 0.76$	Synephrine/HPLC	9.5
3	<i>Herba Caricis phacotae</i>	80	82	300	Phenylethyl alcohol/HPLC	100
4	<i>Herba Menthae</i>	Ambient	75	200 → 150, $K_1 = 0.28$	Menthol/GC	100
5	<i>Polyporus umbellatus</i>	60	70	293 → 229, $K_1 = 0.13$	<i>Polyporus umbellatus</i> polysaccharides/UV	6.0
6	<i>Radix Trichosanthis</i>	80	84	386	Trichosanthin/UV	27.8
7	<i>Cortex Moutan Radicis</i>	60	68	282 → 175, $K_1 = 0.74$	Paeonol/HPLC	0
8	<i>Radix Paeoniae alba</i>	80	79	289 → 250, $K_1 = 0.31$	Albiflorin/HPLC	65
9	<i>Citrus reticulata</i>	60	62	275 → 130 → 62, $K_1 = K_2 = 0.05$	Hesperidin/HPLC	40.0
10	<i>Curcumae aromaticae</i>	80	88.9	280	Camphor/GC	39.3
11	<i>Astragalus membranaceus</i>	80	85	278	Water extracts/standard	17.6
12	<i>Cortex Magnoliae Officinalis</i>	80	72.6	295	Magnolol/GC	12.4
13	<i>Atractylodes macrocephala</i>	80	88.9	360	Water solubles/UV	11.7
14	<i>Cortex Phellodendri</i>	80	67	360	Berberine/HPLC	17.8
15	<i>Radix Platycodi</i>	80	75	387	Platycodside/standard	28.6
16	<i>Rhizoma Ligustici chuanxiong</i>	60	70	374 → 219, $K_1 = 0.81$	Water extracts/standard	3.8
17	<i>Semen Arecae</i>	60	78	351 → 220 → 170, $K_1 = K_2 = 0.23$	Arecolime/standard	22.7
18	<i>Radix Saposhnikoviae divaricatae</i>	80	79.4	377	Mannitol/GC	65.9
19	<i>Radix Glycyrrhizae</i>	60	62	386	Glycyrrhizic acid/HPLC	17.9
20	<i>Rhizoma Dioscoreae</i>	80	78	296 → 241, $K_1 = 0.33$	Allantion/UV	10.25
21	<i>Rheum Officinale</i>	80	98	278 → 248 → 217, $K_1 = K_2 = 0.22$	Rhammoside/HPLC	64
22	<i>Radix Salviae</i>	80	75	281 → 264, $K_1 = 0.71$	Water extracts/standard	11.9
23	<i>Radix Angelicae sinensis</i>	60	64	338 → 286 → 243, $K_1 = K_2 = 0.13$	Ferulic acid/HPLC	55.6
24	<i>Caulis Loniceriae</i>		79.5	360	Chlorogenic acid/HPLC	0
25	<i>Atractylodes chinensis</i>	60	71	388 → 275, $K_1 = 0.55$	Alcohol extracts/GC	0
26	<i>Radix Isatidis</i>	80	77.7	360	Indirubin/UV	38.3
27	<i>Rehmannia glutinosa</i>	60	81	370 → 300, $K_1 = 0.89$	Water extracts/standard	0.08

K = heating time at T /total drying time.

mass of the wet material is about 400 g. It is interesting to find also that some herbs remain active even though drying was carried out a temperature exceeding the allowed values such as *Polyporus umbellatus*, *Radix Trichosanthis*, *Radix Glycyrrhizae*, *Rhizoma Ligustici chuanxiong*, etc. A comparison of the ingredient losses of the herbs dried by heated air and IR is given in Table 26.5. Besides the material temperature, the way in which the heat is supplied and the moisture is taken away from the drying chamber definitely plays a role in the quality of the dried herbs. When infrared drying is combined with vibro-fluidized bed, Figure 26.10, the drying rate as well as the quality of the dried product is improved significantly [23]. Table 26.5 lists 27 herbs.

For information of herbs and their active ingredients, readers are referred to an on-line reference [24].

The typical dryers found in drying of herbs in small pieces are oven, drying house, tunnel dryer, vacuum dryer, and fluidized bed dryer. The descriptions of these dryers can be found in the other chapters of this handbook [23,25].

26.4 CONCLUSIONS

Herb drying includes drying after harvest and drying before processing. The drying after harvest usually involves small volumes, thus the household equipment can also be used. Either heated air or ambient drying

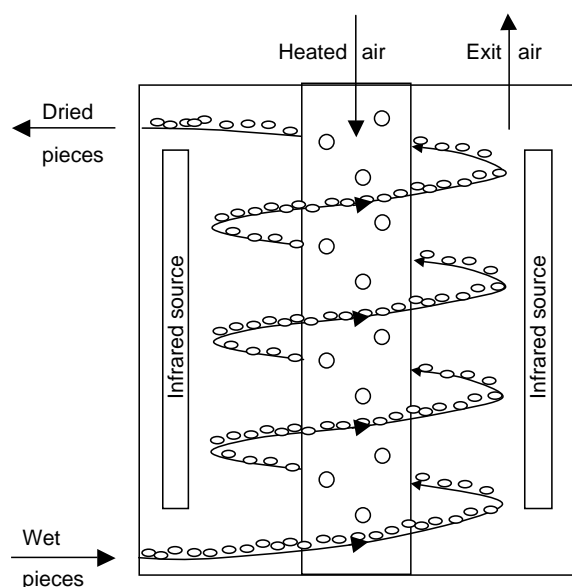


FIGURE 26.10 Vibro-fluidized bed dryer with IR heating. (From Chu, Z.D., *Drying of small pieces of Chinese medicine*, in *Modern Drying Technology*, Eds. Y.K. Pan and X.Z. Wang, Chemical Industry Press, Beijing, 1998, pp. 810–841.)

can be used. Freeze drying, microwave drying, or silica sand drying can be used depending on the value and sensitivity of the herb. For large quantity of herbs, such as tea leaves or ginseng, drying requires industrial-scale operation. The quality of the dried products is important to control and thus governs the drying conditions. For the drying of small pieces of herb medicines, temperature control is important in retaining the active ingredients. The selection of drying techniques is also important in affecting the quality of the dried herbs. Uniform drying of the small herb pieces can give better quality of the products. The reader may refer to other chapters of this handbook for an in-depth discussion of industrial-scale drying e.g., fluidized bed, freeze drying, microwave drying, etc.

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TABLE 26.5
Comparison of Ingredient Loss between Heated Air and IR Drying

	Name	Actual Maximum T , °C		Ingredient Loss, %	
		Heated Air	IR	Heated	IR
1	<i>Cortex Mori Raicis</i>	79	78	–14.9	12.3
2	<i>Fructus Aurantii seu Ponciri</i>	63	62	1.4	9.5
3	<i>Herba Caricis phacotae</i>	67	82	100	100
4	<i>Herba Menthae</i>	68.5	75	100	100
5	<i>Polyporus umbellatus</i>	69.6	70	16	6.0
6	<i>Radix Trichosanthis</i>	71.2	84	66.7	27.8
7	<i>Cortex Moutan Radicis</i>	68	68	2.7	0
8	<i>Radix Paeoniae alba</i>	69	79	60	65
9	<i>Citrus reticulata</i>	66	62	0	40.0
10	<i>Curcumae aromaticae</i>	68	88.9	61.8	39.3
11	<i>Astragalus membranaceus</i>	68	85	0	17.6
12	<i>Cortex Magnoliae Officinalis</i>	72	72.6	2.8	12.4
13	<i>Atractylodes macrocephala</i>	72	88.9	2.8	11.7
14	<i>Cortex Phellodendri</i>	49.1	67	17.4	17.8
15	<i>Radix Platycodi</i>	68	75	33.6	28.6
16	<i>Rhizoma Ligustici chuanxiong</i>	48	70	22	3.8
17	<i>Semen Arecae</i>	69	78	0.5	22.7
18	<i>Radix Saposhnikoviae divaricatae</i>	69	79.4	41.3	65.9
19	<i>Radix Glycyrrhizae</i>	49	62	24.2	17.9
20	<i>Rhizoma Dioscoreae</i>	69.8	78	16.6	10.25
21	<i>Rheum Officinale</i>	70.2	98	43	64
22	<i>Radix Salviae</i>	73.4	75	33.7	11.9
23	<i>Radix Angelicae sinensis</i>	59	64	29.1	55.6
24	<i>Caulis Lonicerae</i>	71	79.5	28.6	0
25	<i>Atractylodes chinensis</i>	49.9	71	0	0
26	<i>Radix Isatidis</i>	43.9	77.7	59.2	38.3
27	<i>Rehmannia glutinosa</i>	62	81	11.4	0.08

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27 Drying of Potato, Sweet Potato, and Other Roots

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27.1 INTRODUCTION

For many centuries potato, *Solanum tubersum*, has served as the primary food source for people in many parts of the world. The potato, with a total production in excess of 300 million tons, is one of the major food crops grown in a wide variety of soils

and climatic conditions. Many cultivars of potatoes are grown, however, a few cultivars account for most of the potatoes produced. These differ in time of maturity, yield, appearance, disease resistance, marketing, and processing quality. The majority of potatoes are used as table food these days, frequently consumed in different forms of processed potato

products. In developed countries, more than 50% of potatoes produced are consumed as processed products. The processed products may be in the form of chips, granules, flakes, power, dice, strips, powder, etc. [1].

The chemical composition of potato depends on many factors including genetic features, age and maturity, as well as environmental conditions, i.e., climatic and soil conditions, weather and growing conditions, such as fertilization, pesticides, diseases, etc. Table 27.1 shows the chemical composition of potato tuber. The potatoes are a considerably richer source of nutrient than energy. Starch is the basic component of potato dry matter, contributing the major amount of energy obtained by potato consumption. Potato contains some nonstarch polysaccharides, which constitute dietary fiber. Potato protein is of superior importance among all the nutrients because of its high biological value. Potato also contains valuable minerals, such as iron and magnesium, and essential vitamins, such as vitamin C and several of B vitamins [2]. The levels of calories, protein, minerals, and vitamins in raw and processed potato products are shown in Table 27.2.

TABLE 27.1
Chemical Composition of Potato Tubers

Component	Content (%)	
	Average	Range
Dry matter	23.7	13.1–36.8
Starch	17.5	8.0–29.4
Reducing sugars	0.3	0.0–5.0
Total sugars	0.5	0.05–8.0
Crude fiber	0.71	0.17–3.48
Pectic substances	—	0.2–1.5
Total nitrogen	0.32	0.11–0.74
Crude protein	2.00	0.69–4.63
Amide nitrogen	—	0.029–0.052
Amino acid nitrogen	—	0.065–0.098
Nitrates	—	0.0–0.05
Lipids	0.12	0.02–0.2
Ash	1.1	0.44–1.87
Organic acids	0.6	0.4–1.0
Ascorbic acid and dehydroascorbic acid (mg/100 g)	10–25	1–54
Glycoalkaloids (mg/100 g)	3–10	0.2–41
Phenolic compounds	—	5–30

Source: From Leszczynski, W., in: *Potato Science and Technology*, Lisinska, G. and Leszczynski, W., Eds., Elsevier Science Publishing Co., Inc., New York, 1989, pp. 11–128.

27.2 PREPROCESSING

27.2.1 PEELING

Peeling is one of the most important steps in potato processing. The yield of finished product depends on effectiveness, efficiency, and method of peeling. In recent years, waste disposal problems have become acute in the processed industries including potato processing. Because the peeling process generates more waste than all of the other potato-processing operations combined, it is important when designing and selecting peeling systems to give proper considerations [3].

The potatoes must be thoroughly washed to remove all mud, dirt, and sand. Potatoes are often washed in machines fitted with cylinder brushes or studded rubber rolls that vigorously scrub the potatoes as they are tumbled and sprayed with water. Washing also assists in sprout removal. Equipment using large amounts of water for washing is now a concern to processors because of the need to recycle the water to meet required water pollution controls. Generally, brush washers use much less water than barrel washers or deepwater washers.

Potatoes are peeled by the use of heat, chemicals, and mechanical abrasive methods. The most common industrial processes are caustic (lye) peeling, steam peeling, and abrasive peeling. The selection of method depends upon the type and variety of products to be produced and the anticipated capacity of the plant. Abrasive peeling is commonly used in the potato chip industry, where minimal peeling is required, and in the canned potato industry, where abrasive action is utilized to shape the potatoes. Abrasive peelers are available either as batch-type units or continuous machines. In steam peeling, potatoes are subjected to steam pressure to rapidly heat and soften the peel and underlying surface tissue. The steam pressure is then suddenly released, causing vaporization of moisture in the heated surface tissue, which further loosens the peel. The peel is then removed either with barrel-type washers or by dry scrubbers. Steam peelers are available in both batch and continuous types. Caustic or lye peeling of potatoes combines the chemical and thermal effects for peeling. In this process, washed potatoes come in contact with hot dilute lye solution followed by washing with high-pressure water sprays. Sometimes combination of lye and steam peeling is also used for peeling of potatoes. The efficiency of lye peeling method can further be improved using an infrared heating treatment of potatoes after potatoes are immersed in more dilute caustic solution. The potato surfaces are exposed to infrared radiation at 900°C for 90 s. This accelerates the chemical activity, thereby utilizing

TABLE 27.2
Composition of Potatoes and Potato Products per 100 g Edible Portion

Product Type	Water (%)	Food Energy (cal)	Protein (g)	Fat (g)	Carbohydrate (g)	Calcium (mg)	Iron (mg)	Thiamin (mg)	Riboflavin (mg)	Niacin (mg)	Ascorbic acid (mg)
Raw	79.8	76	2.1	0.1	17.1	7	0.6	0.10	0.04	1.5	20
Baked in skin	75.1	93	2.6	0.1	21.1	9	0.7	0.10	0.04	1.7	20
Boiled in skin	79.8	76	2.1	0.1	17.1	7	0.6	0.09	0.04	1.5	16
French-fried	44.7	274	4.3	13.2	36.0	15	1.3	0.13	0.08	3.1	21
Fried from raw	46.9	268	4.0	14.2	32.6	15	1.1	0.12	0.07	2.8	19
Hash-brown after holding night	54.2	229	3.1	11.7	29.1	12	0.9	0.08	0.05	2.1	9
Mashed, milk and fat added	79.8	94	2.1	4.3	12.3	24	0.4	0.08	0.05	1.0	9
Canned, solids and liquid	88.5	44	1.1	0.2	9.8	—	0.3	0.04	0.02	0.6	13
Dehydrated mashed, granules, water, milk, fat	78.6	96	2.0	3.6	14.4	32	0.5	0.04	0.05	0.7	3
Frozen, cooked, hash-browned	56.1	224	2.0	11.5	29.0	18	1.2	0.07	0.02	1.0	8
Frozen, french-fried, heated	52.9	220	3.6	8.4	33.7	9	1.8	0.14	0.02	2.6	21
Frozen, mashed, heated	78.3	93	1.8	2.8	15.7	25	0.6	0.06	0.04	0.7	4
Potato chips	1.8	568	5.3	39.8	50.0	40	1.8	0.21	0.07	4.8	16
Potato flour	7.6	351	8.0	0.8	79.9	33	17.2	0.42	0.14	3.4	19

Source: From McCay, C.M., McCay, J.B., and Smith, O., in: *Potato Processing*, 4th ed., Talburt, W.F. and Smith, O., Eds., An AVI Book by Van Nostrand Reinhold Company, Inc., New York, 1987, pp. 287–332.

the caustic more completely and reducing caustic consumption [3].

27.2.2 BLANCHING

Peeled potatoes are cut in the form of slices, dices, and strips before blanching. After potatoes are cut in desirable form, they are blanched by heating, either in steam or hot water (93 to 100°C). Blanching destroys or inactivates enzymes; otherwise, potatoes may darken during dehydration and develop off-flavors and off-odors during storage. Blanching also serves to reduce microbial contaminations and affects the way the dehydrated product reconstitutes. Degree of blanching has a very marked effect on the texture and appearance of finished product as well as on the way the potato tissue dehydrates and reconstitutes. The potatoes are sulfided immediately after blanching using sodium sulfite, sodium bisulfite, sodium metabisulfite, or combinations. Sulfite protects the product from nonenzymatic browning or scorching during dehydration and increases the storage life of the product under adverse temperature conditions.

27.3 DATA ON USEFUL PROPERTIES

Data on physical, thermal, and thermodynamic properties of potato, sweet potato, and other roots are

presented in Table 27.3 through Table 27.11. Such property data are required in product development, process design, and quality control. The data are presented in tabular form. Various prediction models are also compiled for different processing conditions.

27.4 DRYING METHODS AND CONDITIONS

Dehydration of potato dates back to as early as 200 AD when it was cultivated in the mountainous areas of Peru [1]. Today the drying process is considered one of the most important unit operations in the production of a variety of commercial potato products including granules, flakes, diced dehydrated potatoes, and potato starch and flour. The drying process in potato processing can be the end-stage operation (for dehydrated slices, dices, strips, flakes, starch, flour, etc.), accompany-end-stage operation (for granulates, instant products), or mid-stage operation (for chips and fresh fries as predrying before frying in modern technologies). Various drying methods are employed to achieve desired product. In this section, different drying methods and operating conditions used in the processing of potatoes are described. Details on the drying equipment can be found in other chapters of this handbook.

TABLE 27.3
Moisture Diffusivity and Activation Energy for Moisture Diffusivity

Drying Methods	Moisture Content (Dry Basis kg/kg)	Temperature (°C)	Moisture Diffusivity $\times 10^{10}$ (m ² /s)	Activation Energy (kJ/mol)	Ref.
<i>Potato</i>					
Air-drying	0.15	65	2.0		40
Air-drying	—	65	14		41
Air-drying	—	60–80	2.4–2.6		42
Air-drying	—	65	9.0	16.3	43
Air-drying	0.03–5.0	60–100	2.8–53	17.0	44
Air-drying	—	31	0.6–1.6		45
Air-drying	0.05–1.50	30–70	0.2–4.2	30–108	46
Air-drying	—	30–90	1.1–4.5		47
Air-drying	—	40	8.8–12		48
Air-drying	0.10–1.00	65	4.4	52.2	49
Air-drying	—	60	1.8		50
Infrared drying	3.39–4.96	23.5–55	0.6–17.3		18
Air-drying	—	30	0.03–0.2		51
Freeze-drying	—	30	1.3–3.2		51
Freeze-drying	0.01–0.10	30	0.083		52
Puff drying	—	30	0.20–0.70		51
<i>Sweet potato</i>					
Air-drying	0.10–3.5	328	3.7–4.35		53

TABLE 27.4
Relationships Showing Moisture Diffusivity as a Function of Temperature and Moisture Content for Potato

Parametric Model	Ref.
$D(m, T) = a_0 \exp\left(-\frac{a_1}{m}\right) \exp\left(-\frac{a_2}{T}\right)$ $a_0 = 1.29 \times 10^{-6}$, $a_1 = 7.25 \times 10^{-2}$, $a_2 = 2044$ $0.03 < m < 5.0$, $333 < T < 373$	44
$D(m, T) = a_0(m) \exp\left(-\frac{a_1 \exp(-a_2 m) + a_3}{T}\right)$ $a_1 = 168$, $a_2 = 15.4$, $a_3 = 30$	46
$D(m, T, L) = a_0 + a_1 m + a_2 T + a_3 L$ $a_0 = -7.015 \times 10^{-11}$, $a_1 = -2.11 \times 10^{-10}$, $a_2 = 2.57 \times 10^{-11}$, $a_3 = 7.96 \times 10^{-11}$ $0.3 < m < 3.39$, $296 < T < 328$, $2.2 < L < 11.0$ (L in mm)	18

27.4.1 CONVENTIONAL DRYING

Cabinet, tunnel, rotary, drum, or conveyor dryers have normally been used at the industrial scale to produce various potato products. The following description of potato drying is based on the type of dried product.

27.4.1.1 Diced Potatoes

Practically all diced potatoes are produced on conveyor (or apron) dryers, although a few tray and tunnel dryers still are in use. Due to ease of automation in conveyor dryers, the material handling costs are reduced significantly. The conveyor dryer is designed appropriately in stages so that the conditions of the air (i.e., temperature, humidity, quantity, and speed) can be adjusted in a desirable manner as to result in the highest quality of dried product. The most

TABLE 27.5
BET and GAB Monolayer Values

Product	T (°C)	M	K	Y	A/D	Ref.
Potato	19.5	0.075			D	54
	20.0	0.074			A	
	25.0	0.052			A	
	30.0	0.057			A	
	40.0	0.061			A	
	60.0	0.048			A	
	80.0	0.037			A	
Potato		0.051–0.078				55
Potato starch		0.066				
Potato dice		0.060				56
Potato (freeze-dried)	20	0.066	0.849	19.10	A	57
Potato sliced	25	0.083	0.774	8.50	A	
Potato vacuum dried	25	0.130	0.700	1.39	A	
Potato vacuum dried	25	0.091	0.725	8.62	D	
Potato	40	0.052	0.830	13.73	A	58
	50	0.048	0.820	13.41	A	
	60	0.036	0.860	21.18	A	
	70	0.029	0.900	17.75	A	
	40	0.059	0.840	12.52	D	
	50	0.061	0.820	8.89	D	
	60	0.053	0.840	8.57	D	
	70	0.057	0.800	4.37	D	
Potato (Pentland Dell)	30	0.061	999.7	0.86	A	59
	45	0.068	10.2	0.77	A	
	60	0.031	990.1	0.88	A	
	30	0.089	14.45	0.78	D	
	45	0.074	15.68	0.81	D	
	60	0.046	19.99	0.84	D	
Mushroom	25	0.058			A	54

TABLE 27.6
Sorption Data (Average) of Different Peanut Varieties at 25°C

a_w	Moisture Content (% db)			
	Flour Rummer	Tobaldo	Colorado	Roata
0.577	5.6	6.0	5.8	5.5
0.751	8.8	9.0	8.8	8.8
0.803	10.6	10.6	10.7	10.0
0.842	12.9	13.0	12.8	12.6
0.903	17.6	18.2	18.0	17.4

Source: From Bianco, A.M., Boente, G., Pollio, M.L., and Resnik, S.L., *J. Food Eng.*, 47, 327, 2001.

common dryer for diced potato is the two-stage dryer. The first stage of the dryer is divided into two sections. Potato dice bed height and the size of diced potato strongly influence the drying rates. The bed height is maintained constant along the conveyor for proper distribution of drying air and good product quality. Typical bed height is in the range of 7.5 to 15 cm. The temperature of the drying air in the first section of the first stage is in the range 93–127°C, in the second section of first stage 71–105°C, and in the

TABLE 27.7
Net Heat of Sorption

Product	M_w	T Range (°C)	A/D	Q (kJ/kg)	Ref.
Potato	0.05	10–80	A	562.3	61
	0.10	10–80	A	310.5	
	0.15	10–80	A	166.8	
	0.060	10–80	D	1933.3	
Potato	0.080	30–60	D	1555.6	62
	0.100	30–60	D	1283.3	
	0.120	30–60	D	883.3	
	0.140	30–60	D	644.4	
	0.160	30–60	D	527.8	
	0.180	30–60	D	461.1	

second stage 55–82°C. The moisture content of 6–7% (wb) for dried potato is sufficient for proper storage. A further extension of the storage time will require reduction of moisture content to 3–4% (wb), which can be achieved by long-time sorption drying using sorption agents such as calcium oxide.

A belt-trough dryer has been used to produce very high-quality diced potato [4,5]. Only vacuum drying or freeze-drying can produce a more porous product than products dried using a belt-trough dryer. The dryer consists of an endless, closely

TABLE 27.8
Data and Prediction Model for Thermal Conductivity

Product	X_w (% wb)	Apparent Density (kg/m ³)	T (°C)	Thermal Conductivity (W/mK)	Ref.
<i>Potato variety</i>					
Katahdin	81.4	1040	25.5	0.533	63
Russet burbank	82.9	1040	24.8	0.571	
Kennebec	82.4	1050	25.0	0.549	
Monona	83.6	1040	24.6	0.547	
Norchip	81.2	1050	25.9	0.533	64
Potato	83.5	—	25.0	0.563	
	83.5	—	75.0	0.622	
	83.5	—	105.0	0.639	
	83.5	—	130.0	0.641	
$k = 0.1445 + 0.389X_w$ W/mK; $T = 30$; $0 < X_w < 1.0$					65
Potato variety: Bintje					
$k = 0.624 + 1.19 \times 10^{-3}T$ W/mK; $293 < T < 358$; $X_w = 79.8\%$					66
Potato variety: Bintje					
<i>Sweet potato</i>					
Beauregard	—	995	25	0.481	67
Hernandez	—	996	25	0.536	
Jewel	—	1003	25	0.597	

TABLE 27.9
Freezing Point

Product	X_w	T_f (°C)	Ref.
Potato (white)	0.778	−1.70	68
Sweet potato	0.685	−1.90	68
$T = -14.46 + 49.19X_w - 37.07X_w^2$			69

woven metal-mesh conveyor belt supported between two horizontal rolls with a great deal of slack so that it hangs freely (Figure 27.1). The potato pieces are turned and agitated as they travel whereas hot air is blown up through the bed. The airflow provides an air cushion to potato pieces but does not fluidize them. The standard belt-trough dryer, which has a bed 1.2 m wide and 3 m long, evaporates 450 kg of moisture per hour [4]. In this dryer, the moisture level achieved in individual diced potato is nearly the same. The belt-trough dryer has proven to be capable of making dehydrated potato of better quality than those produced by conventional air-dryers.

27.4.1.2 Potato Strips

Potato strips are usually dried in concurrent rotary dryers of 1 to 3 m diameter and 5 to 15 m length. The inlet and outlet air temperatures are about 500 and 100°C, respectively. The fresh potato strips are

TABLE 27.10
Specific Heat

Product	X_w	Specific Heat (kJ/kg K)	Ref.
Potato (raw)	0.75	3.515	70
Potato (boiled)	0.80	3.640	
Potato (dried)	0.061 (27–60°C)	1.715	71
Potato(dried)	0.080	1.925	
Mushroom (fresh)	0.90	3.933	70
Mushroom (dried)	0.30	2.343	
$C_p = 4.180 \times 0.406 + 1.46 \times 10^{-3}T$ $+ 0.203M_w - 2.49 \times 10^{-2} M_w^2$			72
$0 < M_w < 4.13$ and $40 < T < 70$; Desiree variety of potato			
<i>Sweet potato</i>			
Beauregard	995 ^a	3.726	67
Hernandez	996 ^a	3.677	
Jewel	1003 ^a	3.499	
Sweet potato (dried)	0.076	2.050	71

^aApparent density (kg/m³).

TABLE 27.11
Thermal Diffusivity

Product	T (°C)	Thermal Diffusivity $\times 10^7$ (m ² /s)	Ref.
Potato (whole)	−6 to 26	1.77	73
Potato (flesh)	20	1.48	74
Potato (Excel)	24 to 91	1.17	75
Potato (Irish)	−18 to 27	1.23	
Potato (Pungo)	—	1.31	
Potato (several)	25	1.70	
<i>Sweet potato</i>			
Beauregard	25	1.30	67
Hernandez	25	1.46	
Jewel	25	1.65	

prepared by the slicer similar to one used in the sugar industry. The potato strips are dried to a final moisture content of 12% (wb). A significant portion of the dried potato strip is utilized as livestock feed since it is the most economical method of preserving potato. The details of rotary dryer can be found in Kelly [6].

27.4.1.3 Potato Flakes

Potato flakes are dehydrated using drum dryers. Before drying, the potatoes go through washing, peeling, slicing, precooking, cooling, and mashing [7]. Mashed potatoes are then mixed with several additives (emulsifier, chelating agent, milk solids, antioxidants, and sulfur dioxide) to improve the texture and extend the shelf life of the products; they also influence the drying process.

Generally, single-drum and double-drum dryers are used to produce potato flakes (Figure 27.2). The most frequently used cylinders are 0.6–1.25 m in diameter and 0.9–2.2 m in length (double drums), or 2.2–3.2 m in length (single drum). All drum dryers are equipped with several (up to 6) applicator rolls. The space between the applicator rolls allows steam to escape and accumulate mash in the trough between the roll and the dryer drum surface. The temperature of the cylinders reaches 140°C. Saturated steam is used as a heating medium. Mineral oil or gas is used as a heating medium to achieve higher temperatures in the range of 250°C. Such a high temperature allows higher drum speeds, which are typically in the range 2 to 8 rpm. The dry product output is directly proportional to drum speed but inversely proportional to potato sheet density. The moisture content of dry potato flake is in the range 6–7% (wb). Infrared radiation or impinging jets may be used to enhance drying rates.

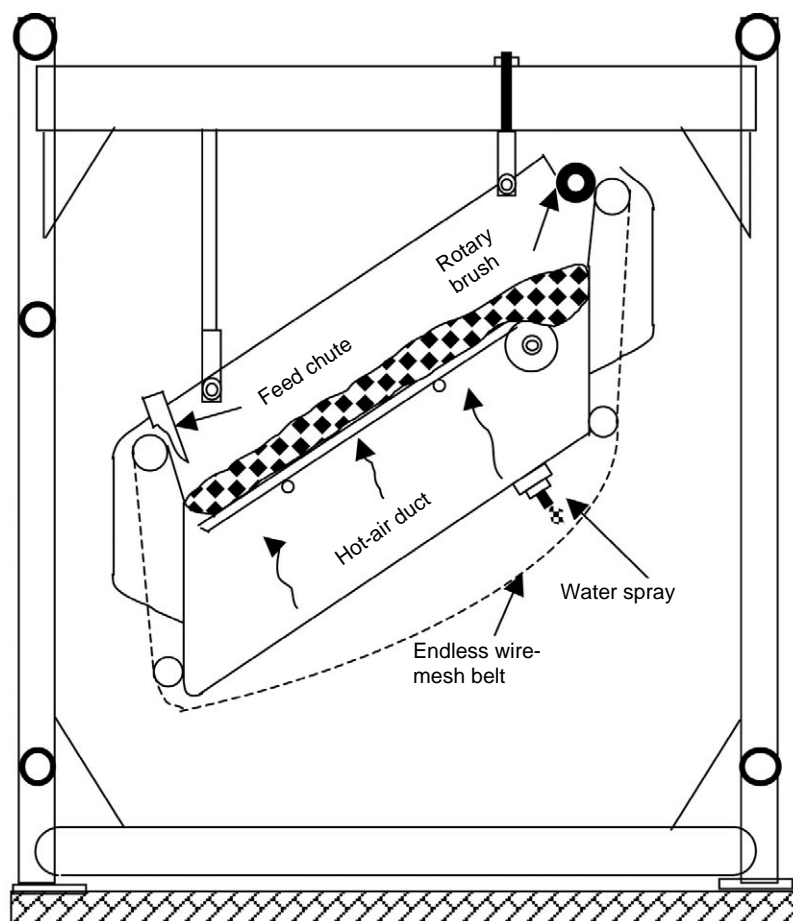


FIGURE 27.1 Schematic cross section of belt-trough dryer. (From Sokhansanj, S. and Jayas, D.S., in: *Handbook of Industrial Drying*, Mujumdar, A.S., Ed., Marcel Dekker, New York, 1987.)

Drum dryers are economical in operation, usually requiring a maximum of 1.3 and a minimum of 1.1 kg of steam per kilogram of evaporated water (76 to 90% efficiency). Technical data on drum dryers applicable in potato flake manufacturing have been presented by Moore [8]. Many improvements have been made by the manufacturers of potato flake dryers over the years. For example, internal pressure has been increased from 100 to 125 psig to achieve increased production rate. Internal design of the drying cylinder has been improved to extend uniform heat transfer to the edge of the dryer. Hydraulic dampers have been installed to reduce wear and improve smoothness of operation. Rod and handwheel devices to control end-board tension have been replaced with air cylinders [7].

27.4.1.4 Potato Starch

The manufacture of potato starch is a simple and easy process based on thorough comminution of potato tubers, separation of starch from potato pulp, its purification and dehydration. All the processing

steps before drying have been described in detail by Lisinska and Leszczynski [9] and Treadway [10].

Starch milk obtained in the last stage of refining in hydrocyclones or centrifuges dewatered by continuous rotary vacuum filters. Extraction of water under vacuum results in 36–38% water content of the starch. After dewatering, starch is transported to the drying section in which moisture level is reduced to 18–20% by supplying hot dry air. The finished product consists of 98–98.5% starch on dry basis. Since moist starch granules begin to swell (gelatinize) at 45°C, the temperatures of moist starch are maintained below 35°C during drying. The final drying of moist starch is carried out in flash (pneumatic) dryers. Such dryers consist of a system of vertical pipes (large in diameter) whereby starch is dried and transported by the air supplied. The air with a velocity of 10 to 20 m/s and temperatures of 160–165°C is mixed with starch. Mixing is more efficient due to the extension of the pipe. Water from the wet starch rapidly evaporates in contact with the hot air. Due to the high evaporation heat of the water removed from the starch, starch

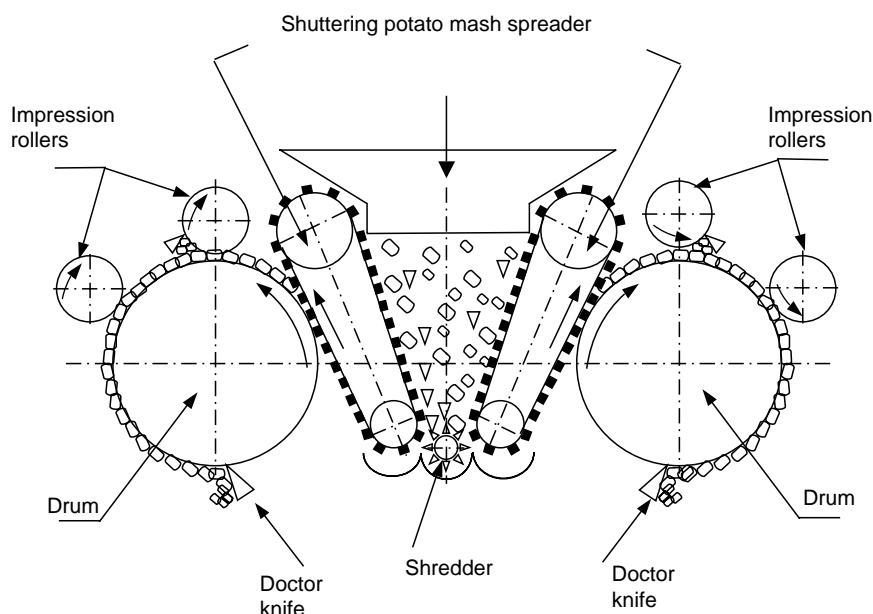


FIGURE 27.2 Scheme of Foerster double-drum potato flakes dryer. (From Lisinska, G. and Leszczynski, W., *Potato Science and Technology*, Elsevier Science Publishing Co., Inc., New York, 1989, pp. 281–346.)

granules are not heated to temperatures higher than 40°C. The entire process of starch drying takes 2 to 5 s.

One of the important parts in potato starch technology is utilization of secondary products such as extracted potato pulp (pomace) and the soluble constituents of the protein water (fruit water). Potato pulp, after dewatered mechanically to a dry matter content of about 25%, is dried in drum or flash dryer using combustion gases at 500–600°C. The dry product of relatively high-feeding value contains about 13% of water. Also fruit water after various processes with centrifuging at the end can be dried by one of the above-mentioned methods or by sonic or TorusDisc drying [11].

27.4.1.5 Potato Granules

Potato granules are dehydrated and single cells or aggregates of cells of potato are dried to about 6–7% (wb) moisture content. The standard commercial procedure used for the production of potato granules is known as “add-back” process (Figure 27.3). In this process, cooked potatoes are partially dried by adding back enough previously dried granules to give a moist mix, which after holding can be satisfactorily granulated to a fine powder. In the manufacturing of potato granules, precautions are taken to (1) minimize rupture of the potato cells and (2) produce satisfactory granulation. Hence all manufacturing operations are carried out as gently as possible.

Drying of granulated moist mix is performed in one or two stages to about 12–13% (wb) moisture

content. Airlift dryers are used in a single-stage operation. The dryer consists of a vertical tube with upward flow of hot air. The moist feed enters at the bottom and drying occurs as the product is rising in the tube. The airlift dryer operates at relatively low air velocities (7.5–10 m/s), with minimum damage to the potato cells. The two-stage dehydration of potato granules is accomplished in a pneumatic dryer (to remove surface moisture) followed by a fluidized bed dryer-cooler. The temperatures of inlet and outlet air in the pneumatic dryer are 175 and 80°C, respectively. The product leaves pneumatic dryer at moisture content of about 17–19% (wb). A two-chamber fluidized bed (with drying in the first chamber and cooling in the second chamber) is used in the second stage to dry potato granules up to 10–12% (wb) moisture content. The temperature of the product leaving the second chamber is obtained in the range of 32–43°C.

The dried granules of 10–12% (wb) moisture content are screened using meshes of different sizes: 6, 40, and 105 mesh. The largest particle fraction from 6 mesh goes for stock feed and small particle fraction passed from 105 mesh undergoes the final step of rehydration. In between particle fraction is utilized at add-back step and also for stock feed.

The drying of fine potato granules is carried out in a fluidized bed dryer-cooler where the moisture content is reduced to 6–8% (wb). A multichamber fluidized bed (or a vibrofluidized bed) or continuous-type fluidized bed dryer is used in the final drying of potato

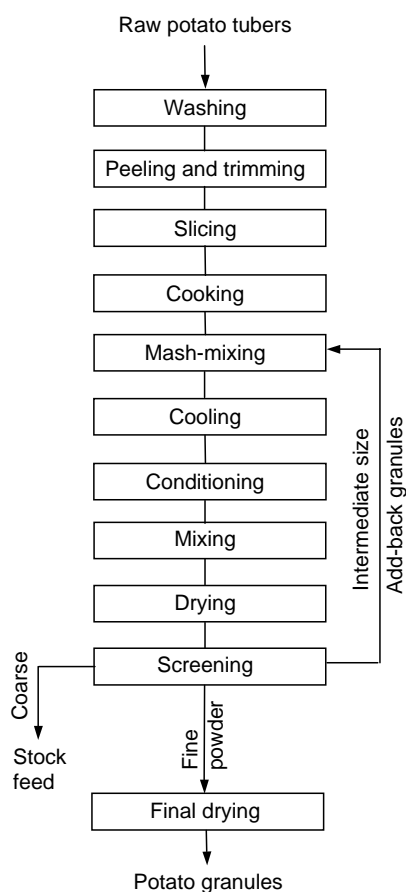


FIGURE 27.3 Scheme of the add-back (solid back-mixing) process for manufacture of potato granules. (From Talburt, W.F., in: *Potato Processing*, 4th ed., Talburt, W.F. and Smith, O., Eds., an AVI Book by Van Nostrand Reinhold Company, Inc., New York, 1987, pp. 1–9.)

granules. The residence time of potato granules in this dryer is about 10 to 30 min.

27.4.1.6 Potato Chips and French Fries

In potato chips manufacturing, raw sliced potatoes are partially dried before frying to reduce the frying time and oil uptake. The raw sliced potatoes are dried on a perforated, revolving drum, on a vibrating mesh belt. The excess surface moisture of about 4% is removed during drying.

In some chips manufacturing, drying process is the finishing step or an alternative frying method. For this, raw sliced potatoes may be partially fried and then subjected to tunnel heat (at 121°C). to remove the excess moisture. The procedure results in lighter colored chips. Finish drying time depends up on the frying time; for drying in hot air at 120°C it is about 1 min (frying time 1 min 40 s) or 4 min (frying time 1 min 20 s).

In the production of frozen French fries, the surface moisture of blanched potato strips is usually removed before frying by blowing warm air over them. This is done to reduce the load on the fryer and to minimize the rate of hydrolytic breakdown of the fat.

27.4.2 OSMOTIC DEHYDRATION

Osmotic dehydration does not allow a product of low moisture content to be considered shelf stable. Consequently, an osmotically treated product is further processed by air, vacuum, or freeze-drying methods. Osmotic preconcentration followed by fluidized bed drying of potato pieces to produce dehydrated quick-cooking potato was studied by Ravindra and Chattopadhyay [12]. They found that a solution of 50% sugar and 10% salt at 47°C for 4 h was optimum for osmotic preconcentration. The moisture content of potato cubes was reduced from 84 to 60% (wb). Drying at 140°C for 10 min at 5.3 m/s followed by thin-layer drying at 50–60°C and 0.75 m/s for about 7 h was found to be optimal. Fluidized bed drying lowered the moisture content to 30% (wb) and it was further reduced to 5% (wb) during thin-layer drying.

27.4.3 VACUUM, FREEZE, AND HEAT PUMP DRYING

Vacuum drying involves subjecting food material to a low pressure and a heating source. The vacuum allows the water to vaporize at a lower temperature and thus foods can be dried without exposure to high temperature. Freeze-drying utilizes a high vacuum to remove water from a solid phase (ice) to a vapor phase without going through a liquid phase. Since the material remains frozen and drying takes place at low temperature, no heat damage occurs. Heat pumps are known to be energy efficient when used in conjunction with drying. Heat pump dryers can employ a wide range of drying conditions (temperature –20 to 100°C and relative humidity 15 to 80%) to produce better quality dried products [13]. Although product quality obtained by vacuum or freeze-drying is superior, both methods are expensive and they are not used on industrial scale for potato products. Heat pump drying is also not applied as widely as it should or could be due to higher initial and operating costs.

27.4.4 MICROWAVE AND INFRARED DRYING

The use of microwave energy has been of growing interest in recent years. The interest is due to short start-up times, volumetric heating, and reduced processing times, making microwave an attractive source of thermal energy. Although numerous studies on

microwave drying of potato have been reported in the literature, microwave drying has not been employed extensively on industrial scale. Microwave drying has been evaluated for final stages of drying (between 30 and 6%) of different potato products. For example, raw potato slices of 0.21 mm thick can be dried to 2–3% final moisture content within about 4 min in a microwave field. Increasing the thickness of raw slices to 0.32 mm leads to cooking effects and also a reduction in the drying rate [14]. Simultaneous air and microwave drying of potato slices can overcome this disadvantage but this type of potato-processing equipment needs further industrial development to make it cost-effective.

Microwave finish drying of potato chips has found practical industrial application [14]. A 25-kW oven can process about 320 kg/h of chips. The quality of the product (color, texture, etc.) is reported to be good whereas oil absorption was about 5% less compared to conventionally processed chips. Similar quality advantages have been reported in other studies [15,16]. This suggests that application of microwave energy to finish drying process of potato chips seems to have potential for use.

Far infrared (FIR) drying also has several advantages over conventional drying [17]. Afzal and Abe [18] studied FIR drying of potato slabs with radiation intensity varying from 0.125 to 0.500 W/cm². They observed that drying rates increased with increasing thickness of potato slab due to a lower activation energy for moisture desorption. Later they demonstrated that FIR drying of potato is a falling rate process and drying rates were dependent on the radiation intensity level [19].

27.5 QUALITY OF DRIED PRODUCTS

Potato is sensitive to heat and mechanical stress; thus thermal degradation of potato products leads to losses in quality (color, structure, shape, etc.) as well as nutrient content (nutrient value, vitamins, protein damage, etc.). The quality of dried potatoes is not only affected by drying methods and conditions but also by preprocessing steps such as peeling, blanching, etc. The losses in quality of dried potato products can be minimized by both pretreatment (adding some chemicals) before drying as well as optimizing the drying conditions. Most of the information related to quality of dried potato available in the literature is based on the experiments carried out at the laboratory scale.

27.5.1 OPTICAL PROPERTY

Color (as an optical property) is a very important quality attribute of dried potato products and is

subjected to appreciable changes during drying. One of the most serious color problems is darkening that occurs in potatoes during the production of dehydrated potato products. Enzymatic oxidation and nonenzymatic (Maillard or browning) reaction are most frequent causes for discoloration. The manifestations in dehydrated potato dice are reddish-brown discolored pieces in the dry state and dark discolored soggy brown centers in the reconstituted state. It is a limiting factor in determining maximum piece size and drying rate. It has been shown that the rate of color change increases as temperature increases and air humidity decreases during conventional and vacuum drying [20]. Most industries involved in the manufacturing of dehydrated potato products use sodium bisulfite as a source of sulfur dioxide to prevent nonenzymatic browning in dehydrated potatoes. Addition of sodium acid pyrophosphate (at a level of 0.1% based on potato solid) before drying prevents discoloration in potato flakes.

The use of time-varying air temperature profiles (sinusoidal and square wave) during drying of potato pieces in a two-stage heat pump has shown to reduce color change. Choy and coworkers [13] observed that the overall color change in potato was minimum when subjected to a higher varying dry temperature (square waveform 35°C, amplitude 5°C, peak to valley 10°C, cycle time 60 min, drying time about 300 min). Other drying conditions were relative humidity 18.9%, absolute humidity 0.0087 kg/kg dry air, and air velocity 2.4 m/s. The percentage reduction in overall color change for potato was 87%. The study suggested that time-varying temperature drying in a heat pump dryer process has potential for drying potato products. Krokida et al. [21] studied the effect of microwave drying on some quality properties of dehydrated potato. The study showed that microwave and microwave–vacuum drying of potato resulted in reduced color change compared to air-drying. The color change of potato during drying can be minimized by various types of pretreatments. Sulfite treatment, blanching, osmotic and microwave pretreatments have shown to reduce the color change significantly during convectional drying of potato [22].

27.5.2 PHYSICAL PROPERTIES

Drying methods and conditions, sample size, and shape have been shown to influence various physical characteristics including structure, density, porosity, shrinkage, and rehydration of dried potato products. These are important properties characterizing the texture and the quality of the dried potato product.

Light microscopic studies of potato slabs (4.5 × 2.0 × 1.0 cm) during air-drying have revealed that

shrinkage occurs first at the surface and then gradually moves to the bottom with increase in drying time. The cell walls become elongated. The degree of shrinkage at a low drying temperature (40°C) was greater than at high temperature (70°C). The surface structure of potato slab is damaged in 1 h drying, whereas the inner structure is apparently intact. As drying proceeds, cracks are formed in the inner structure. When the interior structure finally dries and shrinks, the internal stresses pull the tissue apart. The dry material then contains numerous holes. Shrinkage affects the density and porosity of dried potato. The density at a given moisture content decreases with increasing drying temperature [23].

McMinn and Magee [24] studied the air tunnel drying (temperature 30 to 60°C, velocity 1.5 m/s) of potato. The results of their study indicated that internal porosity of potato cylinders first decreased with drying time then it increased up to 10% at the end of drying. The volume shrinkage exhibited a linear correlation with respect to moisture content. The rate of shrinkage during low-temperature drying was greater than at high temperature and coefficient of rehydration increased to about 4% with increasing drying temperature from 30 to 60°C.

27.5.3 NUTRITIONAL PROPERTIES

Dehydrated potato products contain the same chemical constituents as fresh potatoes, though their amounts vary. The amount of dextrins in the dehydrated products is larger than in fresh potatoes. The amount of sugars is less in dehydrated products than that in the dry matter of fresh potato. The level of cellulose decreases in the dehydrated products, by 100% on average, as compared to its content in the dry matter of fresh potato. The total number of pectic constituents changes slightly during processing. A small amount of fat in fresh potato disappears completely from dehydrated products. The protein content in dehydrated products decreases as compared to the dry matter content of fresh potato tubers. The ash content in the dehydrated products is also lower than that in raw potato. The vitamin content in dehydrated potato products is very low in comparison with fresh potato (Table 27.12).

27.6 SWEET POTATO AND OTHER ROOTS

Sweet potato is an important crop in many parts of the world and is cultivated in more than 100 countries. As a world crop with 119 million tonnes production per year, it ranks seventh from the viewpoint of total production. It is fifth on the list of the developing countries as the most valuable food crops and

TABLE 27.12
Changes in Chemical Composition of Potato (Russet burbank) after Processing

Constituents ^a	Potato Products		
	Raw	Granule ^b	Flake ^b
Total solids (g)	20.8	14.2 (68)	16.9 (81)
Starch (g)	15.0	11.8 (79)	15.2 (101)
Total sugar (mg)	350	170 (49)	169 (48)
Reducing sugars (mg)	108	54.0 (50)	42 (37)
Total nitrogen (mg)	295	180.8 (61)	203 (69)
Protein ($N \times 6.25$) (g)	1.84	1.13 (61)	1.26 (3.5)
Free amino acid (mg)	12.5	7.1 (57)	6.8 (54)
Ascorbic acid (mg)	10.3	3.2 (31)	2.94 (29)
Riboflavin (mg)	0.014	0.006 (43)	0.003 (21)
Niacin (mg)	1.44	0.59 (41)	0.57 (40)
Thiamine (mg)	0.080	0.004 (5)	0.018 (23)
Magnesium (mg)	21.8	11.6 (53)	11.2 (51)
Calcium (mg)	9.4	7.1 (76)	5.4 (57)
Potassium (mg)	443	216 (49)	166 (37)
Sodium (mg)	16.9	17.8 (107)	28.4 (71)
Copper (mg)	0.11	0.03 (43)	0.04 (57)
Manganese (mg)	0.15	0.06 (40)	0.06 (40)
Zinc (mg)	0.19	0.14 (79)	0.15 (79)
Iron (mg)	0.37	0.22 (59)	0.30 (81)

^a100 g serving of each product.

^bValues in parentheses are the percentage of the constituents in raw potato.

Source: From Lisinska, G., in: *Potato Science and Technology*, Lisinska, G., Leszczynski, W., Eds., Elsevier Science Publishing Co., Inc, New York, 1989, pp. 234–280.

in monetary terms, it ranks 13th globally in the production value of agricultural commodities. Like other roots and tubers, the sweet potato has a high moisture content of about 70%. The dry matter in sweet potato consists of average 70% starch, 10% total sugars, 5% total protein, 1% lipid, 3% ash, 10% total fiber and rest 1% vitamins, organic acids, and other components. Sweet potatoes are not only a moderately good energy source, but also contain significant quantities of water-soluble vitamins ascorbic acid and thiamin, besides supplying part of the daily requirement for riboflavin and niacin. They also contain relatively higher amounts of pyridoxine, folic acid, and pantothenic acid [25].

27.6.1 DRYING OF SWEET POTATO

In traditional practice, the roots, which may or may not be peeled, are sometimes cooked but more often directly cut into pieces and spread out in the sun to dry. The dried chips or slices may be only an

intermediate stage in the final production of other products such as flour, snacks, starch, or alcohol. In some cases, fresh sweet potatoes are soaked in 8–10% salt solution for an hour before cutting and drying. This practice is reported to inhibit microbial growth during drying [26]. On a laboratory or commercial scale, sweet potatoes are often treated with a solution of sodium metabisulfite to inhibit enzymic browning, discoloring finished product.

27.6.1.1 Sun and Solar Drying

In many countries including China and India, thousands of tonnes of sweet potatoes are dried every year in the form of chips or slivers by traditional sundrying. The major part of this dried product is then sent on to starch or alcohol factories for further processing. Sundrying has various drawbacks including poor control of energy input and product quality, as well as frequent contamination by microorganisms, dust, and insects. Methodology for the improvement of traditional sundrying to produce a higher quality product has been developed. Solar dryers with temperature control, airflow regulation, and other technical innovations have been designed to minimize the effects of climatic changes, increase sundrying efficiency, and improve product quality. A solar dryer, developed by the International Potato Center, Lima, Peru, can reduce the moisture content of cooked and shredded sweet potatoes by about 45% in 18 h. Reorientation to direct sunlight the following day dries the product to about 10% moisture. A similar type of solar dryer has been developed by the University of Hawaii, Manoa, which can function directly with sunlight, indirectly with solar-heated air, or in a combined direct and indirect mode. The unit can also make use of biogas during temporary unfavorable climatic conditions. May and Chi [27] suggested partial drying of sweet potato slices by immersion in a recirculating concentrated sugar solution of 60–65% for several hours before they are solar-dried.

27.6.1.2 Conventional Drying

Natural convectional dryer has been used in Philippines to produce dried sweet potato. These dryers utilize heat from burning the wastes. About 50 kg of sweet potato chips of high quality are dried for 7–8 h. Another form in which sweet potato is dried and eaten as a dessert or snack is as an edible leather also known as fruit roll or crush. In this preparation, the flesh of sweet potato is cooked, mashed, and sieved, mixed with 0.5% (w/w) carboxymethyl cellulose (a binder), 200 ppm sodium bisulfite, and 7% (w/w) sugar and formed into a sheet 1-mm thick,

which is oven-dried at 55 to 75°C and 10–17% moisture level. Samples dried at 75°C were chewier than those dried at lower temperatures. Samples were crisper and crunchier when dried at 55 or 65°C. A laboratory-scale high-temperature short-time (HTST) pneumatic dryer has been fabricated in India for the pretreatment before conventional drying of sweet potato cubes [28]. Dried cubes produced by this technique were of high quality in terms of porosity, texture, and rehydration characteristics. Peeled, cubed, and blanched sweet potatoes were subjected to HTST pneumatic drying followed by conventional cabinet drying to 5% moisture.

27.6.1.3 Industrial Drying

The production of high-quality flakes and other dried sweet potato products may entail many unit operations at preprocessing and drying stages. Preheating at preprocessing stage reduces enzyme discoloration (polyphenolic oxidation) and peeling time. Steam peeling is often used for removal of the skin. Peeling with lye involves a 5–6 min exposure to 20% lye solution for cured roots, and 3–6 min exposure to a 10% solution for freshly harvested roots, at 104°C [29]. Peeling with superheated steam followed by flash cooling by direct injection of cold water into the peeler chamber increases the yields of peeled sweet potato.

Flakes: After preprocessing, the roots are steamed, crushed, and fed directly to a single-drum dryer heated to 120°C. The dry film is then broken up into flakes. In the United States, the preprocessed roots are pureed in a pulper and screened through 0.8 mm mesh to remove fibrous material. The puree is heated to 74–85°C using steam injection to gelatinize the starch and activate amylolytic enzymes. This processing step results in a consistently textured final product. Flash heating subsequently inactivates the enzymes and completes the cooking. The puree is then dried to 2–3% moisture on steam-heated drum. Sodium acid pyrophosphate or citric acid may be added to the puree before drying to control nonenzymic browning to reduce discoloration tendency of the reconstituted flakes.

Granules: In Japan, sweet potato granules are produced by the add-back process similar to that of potato granules. In this process, roots are peeled, steamed, crushed (moisture content 68%), dry granules added back (moisture content reduced to 35–45%), conditioned, granulated, and flash-dried. The granules can be used in croquettes or snack foods [25].

Diced, slices, or strips: The sweet potatoes are cut into the requisite shapes and sizes. The cut pieces are then blanched and dried into cabinet or tunnel dryers at 40–70°C with co- or countercurrent air circulation.

Chips: For the manufacture of chips, sweet potatoes are peeled, boiled, or steamed for 30 min, mashed, mixed with pea or cornstarch, salt, and monosodium glutamate; mixture is pressed into paste and transformed into rectangular shape. The strips are boiled or steamed to gelatinize the starch, cooled, and cut into slices of 1–2 mm thick. The slices are then dried and packed as semifinished product. The semifinished dried chips are fried before consumption.

Starch: Sweet potato has the advantage of remaining as a high starch-producing crop with 30 and 49% greater starch yield than rice, corn, or wheat, respectively, under the same conditions. In China, at the former level process, sweet potatoes are cut up by hand, crushed by machine, and sieved to remove waste such as peel and fiber. The starch is left to settle in a tank and then removed. Water is added to the starch, which is once more sieved and settled. Wet starch is transferred to jute bags and allowed to drip-dry for 1 d, and air-dried outside at ambient temperatures [25].

27.6.1.4 Quality Changes

Sweet potato contains highly unsaturated lipids, which are susceptible to oxidation. Sweet potato flakes stored in air leads to strong hay-like off-odor, loss of color, and undesirable decline in nutritional value. The highly unsaturated nature of beta-carotene and other carotenoids leads to their rapid oxidation and consequent partial loss in flakes unless they are stored in a reduced oxygen environment. The long storage of dried sweet potato products will require packaging in material, which excludes oxygen as well as moisture. Darkening of uncooked slices and excessive browning during frying results in discoloration in the chips made from some cultivar. Blanching, in boiling water or steam or in a solution of sodium acid pyrophosphate, or dipping in sulfite solution helps to prevent chip discoloration.

27.6.2 DRYING OF MUSHROOMS

Among the cultivated mushrooms, shiitake mushroom (*Lentinus edodes*) is mainly dried and consumed mostly by Orientals. The common mushroom (*Agaricus*) is mainly canned and consumed by Western people. In Japan, consumption of dried shiitake is always 20% higher than that of fresh shiitake. The Japanese consumption of shiitake per capita is about the same as that of *Agaricus* in European or North American countries [30].

Sun and conventional drying: Some mushrooms (*Boletus luteus* in Chile) are predried in Sun to about 15–20% moisture, held for a short time in temporary warehouses, and then taken to a central drying plant.

The mushrooms are then fumigated, cleaned, and sorted to remove dirt and foreign materials. The mushrooms are sliced, spread on tray, and tunnel dried. The drying starts at low air temperature of 49°C, which is gradually increased to a final drying temperature of 71°C and moisture is reduced to 10% [30].

White jelly fungus (*Tremella fusiformis*) grown in China are dried and consumed as a drug or precious food. Drying is usually done only by sun or by hot air. The hot air-drying process takes up to 8 h and the temperature is first held at 50°C, then gradually decreased until it reaches 40°C. Control of temperature and a strong airflow are very important for the quality of dried product. The dried white jelly mushrooms weigh only about 6–8% as much as fresh ones.

Industrial drying: Commercial drying of mushroom involves soaking of mushroom in 0.05–0.1% NaHSO₃ for 5–10 min before cutting into slices to retard discoloration. Unblanched mushroom slices dried at 40–50°C result in light-colored products. Drying temperature above this could cause browning of the product. Light-colored dried mushroom can also be obtained by drying at 40–45°C at the beginning and then raising the temperature gradually to about 60–70°C for 10 to 12 h. The most suitable moisture content of mushroom for storage at room temperature is in the range 6–8%. Freshly harvested shiitake are dried in cabinet dryers to maintain good flavor and the luster of the cap. The mushrooms are dehydrated starting at 30°C and the temperature is then increased to 2°C/h until a temperature of about 50°C is reached (usually in 12–13 h). Finally they are heated to 60°C and held for 1 h to enhance the flavor and bring out the luster of the cap [30].

Combined microwave and hot air drying experiments have shown to greatly improve the structure and bulk volume of dried mushrooms. However, the geometry and dielectric properties of mushrooms are such that potential for overheating of the center hampers the application of this technology [31]. This was overcome by osmotic pretreatment with sodium chloride solution and subsequently dried by application of combined microwave and hot air drying. The results showed that the mushrooms were heated more homogeneously. The rehydration properties and porosity of dried mushrooms were also improved.

Quality and nutritional values: Mushrooms contain large amounts of flavor and aroma producing 5'-ribonucleotides. Common mushroom (*Agaricus*) contains about 50 mol and fresh shiitake mushroom from 182 to 235 mol of 5'-nucleotides per 100 g fresh weight as compared to vegetables, which contain 1 to 10 mol of these compounds per 100 g fresh weight. The process of drying not only increases the shelf life of mushrooms but also enhances the flavor with a

unique taste. Guanylic acid has been identified as the main constituent of the good taste of shiitake extract. The increase of guanylic acid content is due to the decomposition of ribonucleic acids by ribonuclease during cooking at 60 to 70°C. Lenthionine (C₂H₄S₅) has been identified in the aroma of shiitake mushrooms [30]. Studies have shown that the drying temperature and time affected the retention of lenthionine [32]. They proposed two-stage mechanism of formation and destruction of lenthionine during drying. The first stage was formation of lenthionine and other substances simultaneously; then lenthionine was changed to other derivatives. At the second stage, the mechanism that produced other substances was terminated but the mechanisms that produced lenthionine and the one that changed lenthionine to other derivatives were continued. It was found that the optimal condition was maintained at 70°C and 9 h and it retained the maximum amount of lenthionine whereas the moisture content of dried mushrooms was 13% (db). The Chinese believe that shiitake is effective in the preservation of cerebral hemorrhage. The Japanese researchers found that shiitake has ability to remove serum cholesterol and it has antitumor activities [30].

Loch-Bonazzi et al. [33] evaluated the quality of mushrooms dried by various processes such as hot air, vacuum, freeze-drying in an adsorbent fluidized bed and vacuum freeze-drying. Results of their study indicated that vacuum freeze-dried products were of superior quality in terms of density, rehydration capacity, and color. They also reported that most flavor compounds disappear during preparation and freezing steps. Microwave–vacuum drying of mushroom has been explored by Pappas et al. [34]. Their experimental results indicated that microwave–vacuum drying exhibits superior drying performance as well as improved rehydration characteristics.

27.6.3 DRYING OF PEANUTS

Moisture content is the most critical factor in harvesting, drying, storing, and marketing of peanuts. Peanuts are normally between 35 and 50% moisture content at digging. Molds that produce aflatoxin grow best in peanuts between 12 and 35% moisture content and between 26 and 38°C. Peanuts are artificially dried to a moisture content to reach a safe storage level of 7–8%. For marketing, the desired moisture content of shelled peanuts is about 7.5%. Some producers field-dry peanuts to 9% moisture to avoid mechanical drying. This practice is unprofitable because of the higher losses associated with dry vines, stems, and pods. In Oklahoma, farmers have used solar dryers (3 m wide, 1.8 m deep, 7.8 m long) for

peanut drying. The dryers constructed with wire mesh sides and bottoms (0.9 m from ground) allowed the air to circulate through entire load. The drying time was 2–3 days for a batch of about 500 kg of freshly harvested peanuts [35].

Mechanical drying of preanuts is usually carried out in bin dryers. Peanuts are spread evenly over the drying bin. The optimal drying air temperature for peanuts is 35°C. The relative humidity of drying air should not be less than 55%. Higher capacity fan is employed to ensure an airflow rate of approximately 900 m³ air/h/m² drying area. Increasing temperature to 43°C increases the shelling operation. In case peanuts are harvested above 35% moisture content, the height of peanuts bed is kept around 0.6 m. The moisture content of peanuts is frequently checked during drying as peanuts approach the 12% moisture level. This can be done by electronic moisture meters. The heat is cut off when the peanuts reach 10–10.5% moisture but fan is operated continuously until peanuts are cooled, since drying will continue until an equilibrium is reached between the kernel and shells. No portion of batch should dry less than 7% and more than 10% [35].

Quality considerations: In order to produce good quality peanuts, the moisture removal rate is maintained at 0.5% per hour. Peanuts dried too rapidly or stored at low moisture content will have a high percentage of split and bald kernels when shelled whereas peanuts dried too slowly or stored at high moisture content are more subjected to mold growth and aflatoxin production. High-temperature drying is the major cause of bad flavors in peanuts [35].

27.6.4 DRYING OF GINSENG

Ginseng is a herbaceous perennial plant in the *Araliaceae* family, which has been used for thousands of years to cure or prevent a large number of ailments. One of its most recognized roles is as an adaptogen, a herb that strengthens body's defences against stress. Ginseng is native to China, Korea, Russia, and some areas of North America. It contains vitamins A, B-6, and the mineral zinc, which is claimed to help in the production of thymic hormones necessary for proper functioning of the defence system. The primary active ingredients of ginseng are more than 25 saponin glycosides called "ginsenosides." It is noteworthy that the American ginseng is different from the Asian variety in both physical and chemical characteristics.

The principal measures of quality of ginseng root are the contents of ginsenosides and polysaccharides, which are a mixture of complex sugars. The quality of the root depends on the production practices followed by the growers as well as postharvest processing,

e.g., drying of the root [36]. Root age as well as size and shape are said to affect the ginsenoside content. The quality is said to be best at the root age of 6–7 years for Korean ginseng. There are also some varieties of the Korean ginseng, e.g., white and red ginseng. According to Chang et al. [37], red ginseng is among one of the most expensive agricultural products, fetching some US\$3000 for a 600 g package of ten red ginseng roots for the premium heaven grade to just under US\$400 for the low end “good” grade in Korea (figures for 2000).

Raw ginseng upon harvesting must be processed and dried in a short time to prevent spoilage. The postharvest-processing steps include cleaning, trimming of small lateral roots, washing, peeling (optional), drying, grading, and packaging. The market value depends on color, size, maturity, and shape. It is important to dry the roots carefully to obtain high-value-added ginseng.

The initial moisture content of ginseng is about 70–80% wet basis. The optimal final moisture content for storage is about 12–13% wet basis [37]. Diametral shrinkage of about 20–30% has been reported as a result of drying. Typically the roots are dried in convectional dryers using air temperatures of 32 to 38°C. Park et al. [38] proposed a new combined infrared (IR) and convectional dryer for ginseng roots and found that the presence of infrared affected the color of the root but did not affect the final quality of the product in terms of its chemical content. Freeze-dried root gives the highest quality but it is a slow and expensive process. The ginsenoside content may be reduced by over 25% by drying the root using 44°C air instead of 38°C. Internal color also darkens with increase in drying air temperature. As all the moisture is internal and hence in the falling drying rate period it is not necessary to use high airflow rates in the dryer. It is important to expose all drying surfaces uniformly to the drying medium to avoid over- or underdrying of the roots in the batch.

27.7 SUMMARY

An overview is provided for commercial drying practices used in the manufacturing of various potato-based products. Drying practices for other roots such as sweet potato, mushroom, peanuts, ginseng, and tobacco have also been mentioned briefly. In general, the technology related to convectional drying is well established whereas other methods are still under investigation. Microwave drying is used mostly as finish drying operation. Osmotic drying may find its place as a pretreatment method for reducing drying load on conventional dryers and for improving color of final product.

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28 Osmotic Dehydration of Fruits and Vegetables

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28.1 INTRODUCTION

Water as a main constituent of most foods affects food stability, microbial as well as chemical, and is responsible for the consumer perception of many organoleptic attributes, i.e., juiciness, elasticity, tenderness, and texture. It is generally accepted that it is not the quantity of water in food but its thermodynamic state that is responsible for its influence on food stability and texture. The thermodynamic state of water in food is expressed by its activity, which is 0 for absolutely dry material and 1 for pure water. The lower the water activity the more stable is the food, and the texture changes from juicy and elastic to brittle and crunchy.

The lowering of water activity can be achieved in two ways, either by addition of humectants or by removal of solvent (i.e., water). The first way is mostly

unacceptable by consumers as it needs large amounts of sodium chloride, sugars, or polyols to be added to food. Moreover, this way is limited by nutritional and toxicological restraints. The other way is energy intensive, hence the final product is expensive.

The use of osmosis allows both ways of decreasing water activity in food to be applied simultaneously. The permeability of plant tissue is low to sugars and high molecular weight compounds; hence, the material is impregnated with the osmoactive substance in the surface layers only. Water, on the other hand, is removed by osmosis and the cell sap is concentrated without a phase transition of the solvent. This makes the process favorable from the energetic point of view. The flux of water is much larger than the countercurrent flux of osmoactive substance. For this reason the process is called osmotic dehydration or osmotic dewatering.

The food produced by this method has many advantageous features:

- It is ready to eat and rehydration is not needed.
- The amount of osmoactive substance penetrating the tissue can be adjusted to individual requirements.
- The chemical composition of the food can be regulated according to needs.
- The mass of the raw material is reduced, usually by half.

The osmotic dehydration does not reduce water activity sufficiently to hinder the proliferation of microorganisms. The process extends, to some degree, the shelf life of the material, but it does not preserve it. Hence, the application of other preservation methods, such as freezing, pasteurization, or drying is necessary. However, processing of osmotically dehydrated semi-products is much less expensive and preserves most of the characteristics acquired during the osmosis.

28.2 THE NATURE OF OSMOTIC DEWATERING

28.2.1 OSMOTIC PRESSURE

The thermodynamic state of water in solution is characterized by water interactions with solute. Because each molecule has its internal energy and interactions also need energy each substance of the solution is in the defined energetic state. This state referred to one mole of the substance is called the chemical potential.

Chemical potential is a function of concentration, temperature, and pressure. Under isothermal conditions it is solely determined by concentration and pressure. The increase of solute concentration decreases the chemical potential of a solvent, which can also be expressed by its activity according to the following relationship:

$$\mu_w = \mu_{ow} + RT \ln a_w \quad (28.1)$$

where μ_w —chemical potential of water
 μ_{ow} —chemical potential in a standard state
 R —gas constant
 T —absolute temperature
 a_w —water activity coefficient

The interaction of two systems in different energy states is manifested by the energy exchange. This exchange proceeds until the equilibrium state is achieved, which is the state in which chemical potentials of two systems are the same.

The equilibrium state, under isothermal condition, can be achieved by the change of either concentration

or pressure. Excess pressure needed to reach the state of equilibrium between pure solvent and a solution is called osmotic pressure and is expressed by the formula:

$$\Pi = -\frac{RT}{V} \ln a_w \quad (28.2)$$

where V is molar volume of water.

As water is the solvent in foods, the above equation can be simplified to

$$\Pi = -4.6063 \cdot 10^5 T \ln a_w \quad (28.3)$$

Osmotic pressure is related to molar mass of the solute; the smaller the mass, the higher will be the pressure at the same concentration. Electrolytes show higher osmotic pressure than nonelectrolytes because each ion affects the chemical potential of a solvent.

Relationship between concentration and osmotic pressure is shown in Figure 28.1. Osmotic pressure has an inhibitory effect on microorganisms. Most bacteria, yeasts, and moulds do not proliferate at $\Pi > 12.7$ MPa, $\Pi > 17.3$ MPa, and $\Pi > 30.1$ MPa, respectively. Hence, the shelf life of foods can be regulated by the osmotic pressure of the solution in the material.

Difference in osmotic pressure of two systems is a motive power for mass transfer, if the systems are separated by a semipermeable membrane, i.e., the membrane is permeable to solvent and impermeable to solute molecules. This phenomenon is utilized in osmotic dewatering of fruits and vegetables.

28.2.2 THE STRUCTURE OF PLANT TISSUE

A plant cell can be simply pictured as a unit consisting of two main components: the cell wall and the protoplast (Figure 28.2). The cell wall is permeable to water and low molecular weight compounds and is not a

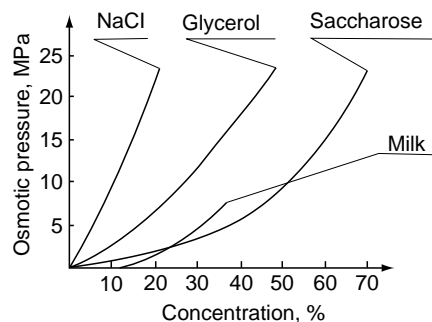


FIGURE 28.1 Relationship between concentration and osmotic pressure of solution.

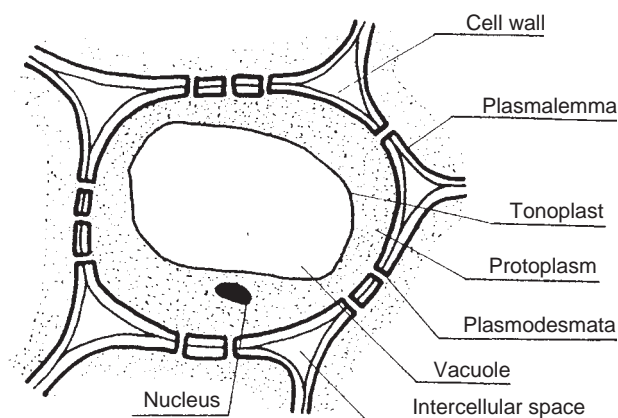


FIGURE 28.2 Plant cell (simplified).

barrier in solute transport from and to the cell. The cell wall is perforated and the channels are filled with thin strands of protoplasm, assuring the contact between protoplasts of neighboring cells. These strands of protoplasm are called plasmodesmata. The diameter of the strands is 20–70 nm and the average contact area can be estimated as $0.2 \text{ m}^2/\text{m}^2$ of the cell wall [1].

The protoplast is composed of protoplasm enclosed in a membrane called plasmalemma, vacuoles, and other structural elements such as the nucleus, plastids, and so on. The plasmalemma is a protein–lipid layer that regulates the contact between the protoplast and the environment. It is 7.5–10 nm thick [2], permeable to water, and selectively permeable to other substances. Protoplasm is a colloidal solution of proteins and lipoproteins in water. The vacuole is suspended in protoplasm and is enclosed in a membrane called the tonoplast. It contains a solution of minerals, sugars, and other organic compounds in water.

Most cells have dimensions between 10 and 100 μm . Depending on their function they are loosely or closely packed in a tissue. Cells are highly specialized. A group of cells that is designated to play a special role in a plant is called tissue.

In general, three types of tissues are recognized. Epidermal tissue forms the outermost layer of cells that are thick walled and covered, in many cases, with cuticle containing a waxy substance known as cutin. Parenchymatous tissue forms the essential part of organs and serves to produce and store nutritional substances. The cells are predominantly large, thin walled, and highly vacuolated. Usually, parenchyma cells are loosely arranged in the tissue and intercellular spaces are formed. The volume of intercellular spaces depends on the kind of plant and its part. Leaves have large intercellular spaces whereas roots are little porous. For example, intercellular spaces in

potato tuber occupy about 1% of total volume, whereas in apple fruit this volume is as large as 20%. Intercellular spaces form a continuous system of channels that is filled with air.

A particular type of tissue is the vascular one. It contains xylem and phloem, which form bundles. Xylem is present in elongated cells with perforated end walls that no longer contain viable protoplasm. Xylem, in other words, forms open dead vessels that provide a way of transportation for minerals and water from roots to other parts of a plant. Phloem is present in elongated viable cells that have sieve end plates. Phloem translocates a solution of sugars, amino acids, and other nutritious substances.

A solution in a vacuole has an osmotic pressure that pushes protoplasm and plasmalemma toward the cell wall. The protoplast is tightly pressed to the cell wall and the cell is in a turgor state. The difference between the osmotic pressure in the cell and in its surroundings is called the turgor pressure.

28.2.3 MASS TRANSFER IN OSMOTIC PROCESS

If the cell and the surroundings have the same osmotic pressure then turgor pressure is zero and the system is in thermodynamic equilibrium. Osmotic pressure of the surroundings lower than that of the cell causes transfer of water into the cell. The cell swells, but the rigid cell wall limits the extent of swelling. A cell placed in a hypertonic solution (osmotic pressure higher than that of the cell) will lose water. The dehydration of a protoplast causes decrease of its volume and, in consequence, detachment of plasmalemma from the cell wall. This process is called plasmolysis (Figure 28.3). As the cell wall is permeable the volume between the cell wall and plasmalemma fills with the hypertonic solution.

Osmotic dehydration occurs on a piece of material and not on a single cell. Hence, it should be assumed that the piece exists in all kinds of plant tissue. As a rule, a skin is removed from the raw material; therefore, epidermal cells and cuticle are absent in most cases. A piece of fruit or vegetable thus will contain parenchymatous and vascular tissue and intercellular spaces, as well.

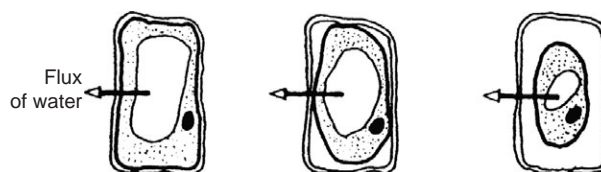


FIGURE 28.3 Plasmolysis.

From the process point of view, a plant material can be considered as a capillary-porous body that is divided internally in numerous repeating units. Some capillaries and pores are filled with a solution, whereas others are empty (i.e., contain air). Most capillaries and pores are open. Repeating units can exchange water between each other.

The internal structure of a body is not a homogeneous one as far as transport of water is considered. Cell walls are built from microfibrils, and intermicrofibrillar spaces are some 10 nm in cross section [3]. These spaces are large enough to allow water, ions, and small molecules to pass through them. As cell walls are interconnected in the tissue, a continuous matrix capable of transporting water and small molecules is formed. This continuum is called the apoplast. In a majority of cells, protoplasm of neighboring cells is interconnected through plasmodesmata and another continuous network is formed. The system of protoplasts and connecting plasmodesmata is widely known as symplast. Because plasmodesmata permit the passage of solutes [1], they undoubtedly permit the passage of water also. The apoplast and symplast as networks capable of transporting water are separated from each other by plasmalemma. Each vacuole is enclosed in tonoplast and they are not interconnected. Hence, vacuoles form a discontinuity in the system.

Two ways of water transport in a plant have been recognized: apoplastic and symplasmic (Figure 28.4). It is generally agreed that the cell walls provide the major pathway of water movement in plant material. The ratio of volume flows in the apoplastic and symplasmic (vacuole-to-vacuole) pathways is of the order of 50:1 in leaf tissue [1]. For the root cortex, the ratio is lower.

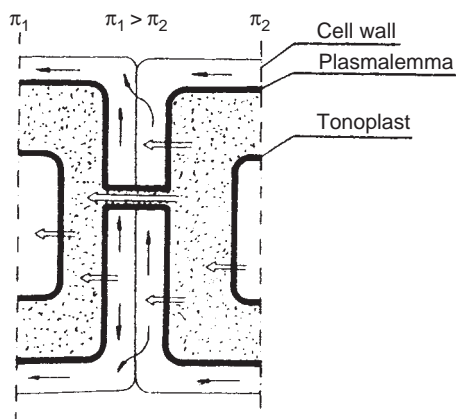


FIGURE 28.4 Apoplastic and symplasmic transport of water.

The capillary and porous system of the body exists in vascular tissue and intercellular spaces. Xylem forms an open conduit of relatively low hydraulic resistance that is filled with diluted mineral solution. Phloem exists in cells with a width ranging from 10 to 70 μm and a length from 100 to 500 μm in dicotyledons [4]. Their turgor is around 2 MPa (beetroot is 1.83 MPa) with a pressure gradient of 0.02–0.03 MPa/m [5,6]. As phloem transports substances of very different molecular weight, shape, charge, and surface activity along with water, it is presumed that the mechanism is an osmotically driven solution flow [6].

The intercellular system of channels has the volume dependent on the kind of tissue. In potatoes, it occupies 1–3%, whereas in beetroot 25% of volume is attributed to cell walls and intercellular spaces [7].

There is no doubt that all these structures of the transport system in the plant tissue will participate in the process of osmotic dehydration.

Contacting plant tissue with the hypertonic solution, a sequence of mass transfer processes can be envisaged (Figure 28.5) as follows:

- In intercellular spaces a capillary suction will occur. The channels will fill in with the hypertonic solution and the gas phase will be compressed or pushed out until the equilibrium state will be achieved.
- Xylem and phloem containing solutions of lower osmotic pressure than the hypertonic solution will be penetrated by the osmoactive substance by diffusion. Osmotic pressure flow can also take place.
- Cell walls in contact with hypertonic solution will lose water due to diffusion and osmotic flow. Osmoactive substance will penetrate cell walls by diffusion.
- Change of osmotic pressure in xylem and phloem and the dewatering of the cell walls will initiate the symplasmic movement of water in the material. The dehydration of the cells will take place and plasmolysis will be induced.

The sequence presented above also suggests the kinetics of the osmotic dehydration process. As long as the mass transfer processes are not strongly dependent on the symplasmic pathway, the water flux will predominate over the osmoactive substance flux. This is due to osmotic pressure flow, which will reduce the countercurrent diffusion of osmoactive substance but it will not strongly affect the diffusive flux of water as the self-diffusion of water in a solution is of the same order of magnitude as that for solute. When plasmolysis occurs and the hypertonic solution fills in the volume between cell walls and plasmalemma, the

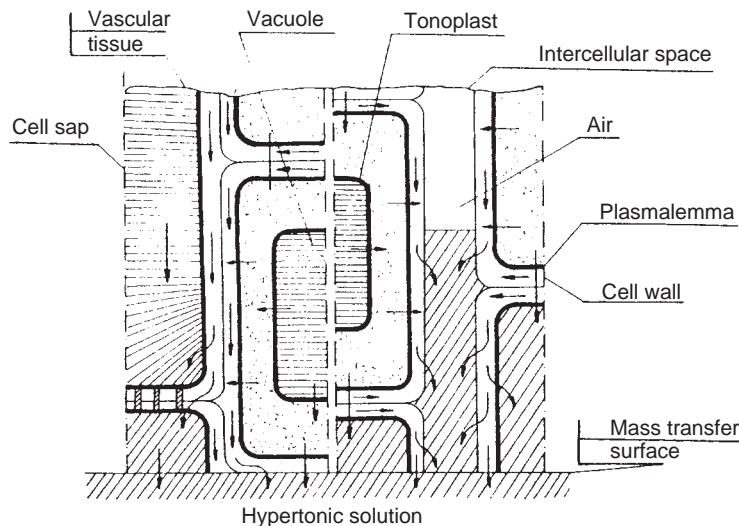


FIGURE 28.5 Possible ways of transport of water in a plant tissue during osmotic dehydration.

process of dewatering will be substituted by the impregnation of the tissue. The flux of osmoactive substance will be equal to or will surpass the flux of water.

The above picture of the mechanism of osmotic dehydration suggests that the plasmalemma resistance to mass transfer affects the process to only a small extent. The process will be rather dependent on the internal resistance to osmotic flow and apoplast dewatering and, to some extent, on external resistance to mass transfer.

28.2.4 MODELING THE OSMOTIC DEHYDRATION PROCESS

From the previous description of the structure of the plant material and processes that can be involved in the mass transfer between plant tissue and the osmotic solution, it is evident that the modeling of osmotic dehydration is not simple.

Models have been developed that describe the behavior of the plant tissue under normal growing conditions, i.e., when the osmotic pressure in the tissue is higher than that in the surroundings and the plant cell is in the turgor state [8–14]. They shed some light on the possible ways the water and solute molecules move in a plant tissue during osmotic dewatering, and that was presented in previous chapter. Moreover, the models were used, in some cases, to quantify the osmotic dewatering process.

In the last five years a lot of work has been done on modeling the osmotic dehydration process. Mostly, the theory of molecular diffusion in the solid has been used to predict the water loss during

the process. An unsteady unidirectional diffusion described by the second Fick's equation was used to quantify the process by the effective diffusivity. The resulting diffusivity is generally correlated with the concentration and the temperature of the hypertonic solution [15–28].

The models based on the second Fick's equation do not necessarily simulate the osmotic dehydration process. In this process countercurrent fluxes of water and the osmoactive substance occur. Moreover, a flux of soluble solids from the tissue accompanies a flux of water (Figure 28.6). Hence, there is a simultaneous mass transfer and probable interactions between flows cannot be taken into account. The estimated effective diffusivities are affected by the countercurrent flows and they cannot be used to predict the contribution of each flux to the process. Moreover, in these models the resistance at the surface of the

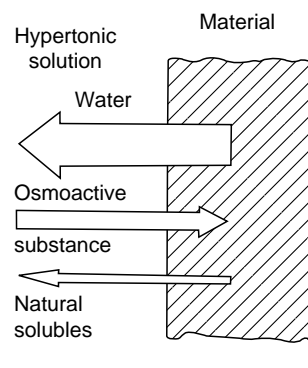


FIGURE 28.6 Mass transfer in osmotic dehydration.

solid is assumed to be negligible, thus the whole resistance to mass transfer is in the solid. Last, the models do not take into account the possible effect of the living cell on the mass transfer process.

A pseudo-diffusion approach [29] to model mass transfer in osmotic dewatering showed that effective diffusivity was not a unique function of Fourier number, as it would be expected in pure diffusion. The developed model had no predictive ability. On the other hand, experiments done on the frozen apple tissue showed that mass transfer was only diffusive [30].

Besides all the limits and contradictory results, the models based on the second Fick's equation proved to be quite successful. Hawkes and Flink plotted the normalized solids content of apple versus the square root of time and obtained a straight line, the slope of which was called the mass transfer coefficient [15]. This approach was used in numerous works [16,31–38].

A model was developed based on the irreversible process thermodynamics in which the cell membrane characteristics, the cell volume changes, tissue shrinkage, internal volumetric rearrangements, and diffusion of nonpermeating and permeating species are taken into account [39,40]. The set of equations solved numerically showed the model as satisfactorily representing the behavior of parenchymatous storage tissue undergoing osmotic dehydration. Moreover, the simulations have shown that the cell membrane represents the major resistance to mass transfer in such systems. The model needed simultaneous adjustment of four constants to obtain a good fit, hence its practical usefulness is rather questionable.

Marcotte et al. improved Toupin's model by giving a closer thermodynamic description of forces involved in the osmotic dehydration process [41]. The transmembrane transport is modeled on the basis of irreversible thermodynamics whereas transport in the intercellular space is modeled by relations derived from the second Fick's equation.

Further development of this model incorporating diffusion, bulk flow, transmembrane flux, and matrix shrinkage [42–44] showed that the cell membrane is the main barrier to mass transfer only for single cells or thin slices of tissue. When the thickness of the sample increases, the extracellular space may become the limiting factor [45].

The models based on the irreversible process thermodynamics show that the cell membrane (plasmalemma) represents the major resistance to mass transfer. This is contradicted by findings of Raoult-Wack et al. [46–48], who showed that membranes are not necessary for osmotic dehydration and merely diffusive properties of the material are responsible for high water flux with only marginal sugar penetration. These authors suggest the following mechanism.

At the beginning of the process the removal of water concentrates the superficial layer of solute in the surface of the material. This layer is detrimental to further solute incorporation but is favorable to water removal as it creates a pronounced concentration gradient [48–50]. The compartmental model was developed that provided good fit for the different situations tested. The solution of the set of differential equations was done by numerical methods.

The role of intercellular space and capillary flow in osmotic dehydration was well documented by Fito et al. [51–53]. On this basis the nondiffusional mass transfer model was developed incorporating hydrodynamic mechanism (HDM). Studies done by the same group [19,54] showed that long time of osmotic process is needed to obtain a fully developed water and sugar concentration profiles. A model based on the advancing disturbance front (ADF) was proposed that allows prediction of sample concentration during osmotic dehydration.

The above models of osmotic dewatering were developed taking the processes into account that take place in plant tissue when the tissue is contacted with hypertonic solution. Empirical models are also proposed in the literature.

Mass transport during osmotic dehydration was described by the first order kinetics in which the rate constant is a function of main process parameters [55] or by an empirical equation with two parameters, which correspond to the initial rate of mass transfer and to equilibrium conditions, respectively [56]. A kinetic model based on the theory of the decreasing nucleus was also proposed to describe the osmotic process [57].

Phenomenological or empirical models describing osmotic dehydration are of limited use. The phenomenological models try to explain processes taking place in plant tissue during osmotic dewatering. Because of that they are general and do not account for individual response of the tissue to osmotic stress. And a response of cells to osmotic stress depends on the origin and morphology of plant tissue [58]. On the other hand, empirical models are developed for the investigated product and treatment conditions, and they cannot be used to model the process in general. A lot of work has been done to understand osmotic dewatering of food products. The research done on micro- as well as macroscopic scale presents fairly well processes occurring in plant tissue during osmotic dewatering. However, the great variability of plant tissue structure and its response to osmotic stress makes it difficult to control the main variables of the process. Hence, current technologies are still somewhat empirical.

28.3 DESIGN OF OSMOTIC DEHYDRATION

Osmotic dehydration is a versatile process, which makes it possible to produce a variety of products based on the same raw material. By a proper choice of process parameters, a wide range of products can be obtained starting with highly dehydrated and low solute infiltrated products and ending with highly impregnated and little dehydrated products. If hypertonic solution contains other solutes the chemical composition of the final product can be formulated, according to the request [59]. Hence, the design of the osmotic process depends on the expected quality attributes of the final product.

28.3.1 PREDEHYDRATION TREATMENT

Fruits and vegetables undergoing processing come from different parts of a plant. They are roots (carrots, parsley, beetroots), stems (kohlrabi, potatoes), shoots (asparagus, onions), leaves (cabbages, spinach), flowers (cauliflower, broccoli), fruits (tomatoes, cucumbers, pumpkins, apples, pears, plums, green bean), and seeds (green peas, beans) and must be appropriately prepared for the osmotic process.

The epidermal tissue has very low permeability for water and solutes, hence the skin must be removed before osmotic treatment. In the case of small fruits such as berries and grapes the skin permeability must be increased. It can be done by treatment in NaOH solution containing ethyl oleate. It proved to be effective for tomato [60] and strawberries [61].

Most of fruits and vegetables are cut into pieces before they are contacted with hypertonic solution. Shape and size of the material pronouncedly affect the rate of the process. Osmosed fruits and vegetables have different forms that come from the technology and consumer requirements [62,63]. Plums were dehydrated in whole or in halves [64]; apples were cut into 12 segments [65,66] or sliced into 3-mm slices [67] or 3–4 mm thick [15]. Peaches were cut into 6 or 8 segments and pears into 8 segments [65]. Carrots were cut into cubes of 5 or 10 mm [68,69]. Potatoes were sliced 5 and 10 mm thick [70], or diced [20]. Papaya was cut into cubes.

Lenart and Lewicki have shown that the thickness of the material should not exceed 10 mm [71,72]. Taking into account further processing following osmotic dehydration and use of the product, they considered a cube with a side dimension close to 10 mm as an optimal size and shape for most materials. Lewicki et al. [73–77] and Lerici et al. [78] dehydrated apples, carrots, and potatoes as cubes of 8–10 mm on a side. Flink, as well as Simal et al. [20], likewise used this shape in most of his studies [79].

Blanching of fruits and vegetables before osmotic treatment strongly affects the course of the process. Blanching of carrots and potatoes reduces water loss and increases solids gain [73,77,80]. Hence, its effect is detrimental to osmotic dehydration of these materials. Steam or microwave blanching of strawberries affected volatile profile of the product inhibiting formation of esters of furanones [81]. Steam blanching gave better results than microwave treatment [82,83]. Blanching of apple pieces either by high temperature short time (HTST) or low temperature long time (LTLT) process resulted in softening of tissue [84] and adversely affected quality of the dried apricots [85].

It has been reported that immersion of some materials in CaCl_2 solution prior to osmotic dehydration affects the properties of the product. Texture of apple was improved [84]. Brining of cashew apple in NaCl solution before osmotic dehydration resulted in firmer texture of the candied product [86].

Immersion of cut material in ascorbic acid [87] or citric acid [88] is used to prevent browning of the tissue.

The effect of high pressure on osmotic dewatering was studied by Rastogi et al. [89,90]. It was shown that the treatment has a damaging effect on tissue and results in higher permeability to water and solutes. The process of osmotic dehydration is facilitated by high-pressure pretreatment.

High-intensity electric field pulses accelerated osmotic dehydration of carrot [18]. A Fickian diffusion coefficient for water and solute increased exponentially with electric field strength. This effect was attributed to increased cell wall permeability, which was also manifested by the softening of product.

Most of the above-described treatments are aimed at increasing permeability of plant tissue and facilitating the osmotic dehydration. Damage to the tissue structure results in increased permeability to water as well as to solute. Hence, the pretreatment has no selective effect and faster water removal is accompanied by greater infiltration of solute into osmosed material. In many cases, it is desirable to reduce infiltration of solute into tissue as much as possible to obtain dehydrated product with little or no change in its chemical composition.

The influx of osmoactive substance into the tissue can be hindered by special pretreatment or by artificial semipermeable membranes. It was shown that convective drying of apple cubes for time as short as 10 min forms a type of skin on the surface which does not affect the flux of water but reduces the flux of sugar during further osmotic dehydration [91]. Artificial semipermeable membranes are formed from pectins or starch [91–95], alginate or low-methoxyl pectin [96]. Coatings significantly reduce solute incorporation into tissue and result in increased weight loss during osmotic treatment.

28.3.2 OSMOACTIVE SUBSTANCE

Osmoactive substances used in food must comply with special requirements. They have to be edible with accepted taste and flavor, nontoxic, inert to food components, and if possible, highly osmoactive.

Quality of the final product was the main aim of most experiments testing the suitability of different osmoactive substances. Their technological applicability is estimated on water loss rate and final water content in the material. Usually, saturated solutions, or the solutions at the same concentrations, are compared. Flink [97] used two criteria to rate different osmoactive substances: water loss and the amount of the substance penetrating the material that is osmosed. Lowering of water activity in the material was also used as an indicator of suitability of an osmoactive substance [98].

Solutions of sugars are mostly used to dehydrate fruits; and glycerol, starch syrup, and sodium chloride are used for vegetables [62,73,91,99]. Sucrose is the most frequently used substance [17,65,100–104]. The control of pH of sucrose solution is recommended for banana slices osmotic dehydration [105]. It was also shown that control of pH of sucrose solution affects the course of osmotic dehydration of apple and carrot [106]. Addition of ascorbic acid to sugar solution is practiced to minimize browning of fruit pieces during osmotic process [72]. Sucrose can be substituted in part by lactose [15].

Glucose and fructose give a similar dehydration effect [107,108]. In other publications it is reported that fructose increases the dry matter content by 50% as compared with sucrose. Water activity of the final product was also lower with fructose as a hypertonic solution [109]. In apple, banana, and kiwifruit, glucose caused higher water loss and solids gain than sucrose [102]. Starch syrup makes it possible to have similar final water content in dehydrated material as that obtained with sucrose but at a much lower influx of osmoactive substance into tissue [87,101,110]. The dextrose equivalent of the syrup strongly affected the ratio between water loss and solids gain. Corn syrup solids [111], cane sugar syrup [112], palm sugar syrup [113], and hydrolyzed lactose syrup [114] were also used in osmotic dehydration of fruits and vegetables. The effect of the kind of osmoactive substance on the water content of osmosed material is presented in Figure 28.7 [115].

Mixtures of osmoactive substances are also used. Maltini et al. [116] used sucrose and starch syrup in a ratio of 1:1. Lerici et al. [117] dehydrated apples in a solution containing 42% fructose, 52% sucrose, 3% maltose, 3% polysaccharides, and 0.5% sodium chloride in dry matter. Mastrocola et al. [118] used solutions containing sucrose and fructose in varying

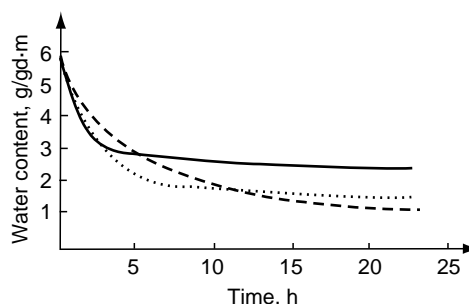


FIGURE 28.7 The effect of osmoactive substance on the course of osmotic dehydration of apples at 30°C (....., glucose; -----, saccharose; , starch syrup). (From Lenart, A. and Lewicki, P.P., of in *IDS '89* Mujumdar, A.S. and Roques, M., eds., Hemisphere Publ. Co., New York, 1990, p. 501.)

proportions. Water loss was similar for all solutions tested but the penetration of the osmoactive substance was different. Peaches dehydrated in solutions of glucose and fructose were especially suitable to pasteurization [119]. Solution of sucrose and glucose yielded high drying rate of apple slices [17]. Mixture of crystalline sucrose and glucose lowered water activity of guava to 0.77 whereas sucrose alone yielded a water activity of 0.80 [120]. Mixture of sucrose and sodium chloride is used to osmose fruits and vegetables [121–123]. The presence of NaCl facilitates the process. Mixture of sucrose and citrate was used in osmotic dehydration of peas [124–126] and papaya [127].

Sodium chloride was used to dehydrate vegetables. Speck et al. [68] used 10% solutions to dehydrate carrots. Lewicki et al. [80] used 15% NaCl to dehydrate carrots and potatoes. Adambounou et al. [128] dehydrated paprika, tomatoes, and eggplant in saturated salt solutions, getting water activity as low as 0.8. Vijayanand et al. [129] used 5–25% NaCl solution to dehydrate cauliflower. Use of sodium and potassium chlorides made it possible to regulate sodium and potassium content in dehydrated corn and green peas.

It has been found that the addition of low molecular weight substances such as sodium chloride, malic acid, lactic acid, and hydrochloric acid in concentrations of 1–5% to sugars or starch syrups improves the process of osmotic dehydration. In general, they promote removal of water from the material. Calcium chloride and malic acid were added to sucrose to improve the texture of osmosed apples [130].

28.3.3 PROCESSING PROCEDURES AND PARAMETERS

Osmotic dehydration can be done basically in two ways: by static or dynamic process. In a static process,

the material is mixed with an osmoactive substance, which can be used as crystals or solution, and the mixture is left motionless until the desired water loss is achieved. It has been shown that the mass transfer resistance in this method is higher than that observed in a dynamic process [131,132].

In a dynamic process, the mixture is mixed; different methods of mixing can be used. Movement of food particles in a stationary solution, mixing of the whole suspension, and the flow of the osmoactive substance through the stationary layer of food pieces are the commonly used designs of the dynamic process. If crystals of the osmoactive substance are used, the fluidized bed is the solution for the dynamic process. It has been shown that the rate of motion has little effect on the rate of osmotic dehydration [71,133]. It is just sufficient to induce motion of particles or solution in the system to have increased mass transfer rates. Moreover, it was shown that the motion of osmotic solution in a turbulent region affected water flux but no difference in solids gain occurred in comparison with laminar flow [134].

Azuara et al. [121] applied centrifugal force to suspension of potato and apple slices in hypertonic solution. The force affected solids gain much more than the water loss. In comparison with static method, the application of centrifugal force resulted in larger water flux and smaller solids gain.

Reduction of pressure during osmotic dehydration increases the rate of the process [115,135]. It has also been observed that low pressure facilitates penetration of the osmoactive substance into the tissue [136,137]. Osmotic dehydration under reduced pressure is done in two ways: reduced pressure is kept continuously or reduction of pressure is done in pulses [83,127,137–139]. In general, pulsed vacuum osmotic dehydration gives better results than vacuum osmotic dewatering.

Apple cubes subjected to osmotic process and treated by ultrasound, dewatered faster than the nontreated ones [20]. Water and solute transport rates were significantly higher in sonicated samples in comparison with those not sonicated during osmosis.

The rate of osmosis increases with increase of concentration of the osmoactive substance (Figure 28.8) [23,66]. The weight loss of mango and papaya is linearly dependent on sucrose concentration up to 60% [140]. Similar results were obtained for apple, carrot, and pumpkin. At higher concentrations, a lower rate is observed and the impregnation of fruits with saccharose is high. Ponting [65] used 65–70% sucrose solutions to dehydrate apples. Pinnavaia et al. [101] recommend 70% starch syrup for osmodehydration of apples. Lenart and Cerkowniak used glucose, saccharose, and starch syrup solutions to

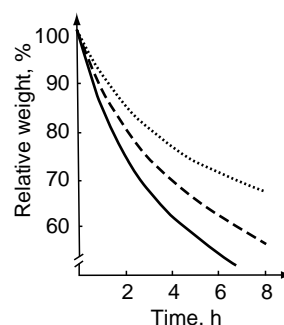


FIGURE 28.8 The effect of sugar syrup concentration on the course of osmotic dehydration of apples at 50°C (70°Bx; -----, 60°Bx; 50°Bx). (Adapted from Farkas, D.F. and Lazar, M.E., *Food Technol.*, 23, 688, 1969.)

dehydrate apples [141]. Rastogi et al. [23] recommend 40–70°Bx sugar solution for osmotic dehydration of bananas. A high rate of dehydration of carrot and potato was obtained with 15% sodium chloride solution [142,143].

Crystalline osmoactive substance is used at a weight ratio of 1:1 to fruits [45,105]. For solutions, investigations were done at weight ratios of 1:1 to 1:6 [87,144]. Osmotic dehydration of fruits and vegetables is recommended to be done at a weight ratio of 1:4 to 1:5 of food to osmoactive solution [77,145].

Temperature has a substantial effect on the course of osmotic dehydration. It not only affects the rate of the process but also influences the chemical composition and properties of the product. Increased temperature increases the rate of chemical reactions and mass transfer processes as well. Viscosity of hypertonic solution is lowered and the diffusion coefficient of water increases with the increase of temperature [21,77].

Andreotti et al. [31] recognize a temperature of 43°C as the optimal for osmotic dehydration of cherries and pears in glucose or glucose–fructose syrup. They recommended a temperature of 20°C for osmotic dehydration of apricots. Bananas were osmotically dehydrated at 60°C [146]; however, it was shown that optimal temperature was dependent on the concentration and pH of the osmotic solution [105]. Pineapple was dehydrated at 42–47°C [96] but application of vacuum and temperature higher than 40°C resulted in loss of volatiles [112]. Osmotic dehydration of plums is done at 50°C [147,148], kiwifruit at 37°C, and peas at 50–70°C [124].

Kowalska et al. [149] recognize a temperature of 50°C as the optimal for osmotic dehydration of strawberries and cherries in glucose, sucrose, and starch syrup solution. They recommended osmosing fruits at 30–50°C. Nsonzi and Ramaswamy [21] osmotically dehydrated blueberries at temperature 37–60°C and

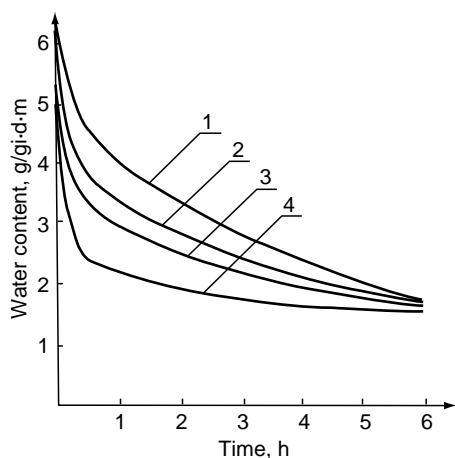


FIGURE 28.9 The effect of temperature on the course of osmotic dehydration of apples in saccharose solution (1, 30°C; 2, 50°C; 3, 70°C; 4, 90°C). (From Lenart, A. and Lewicki, P.P., in *IDS '89*, Mujumdar, A.S. and Roques, M., eds., Hemisphere Publ. Co., New York, 1990, p. 501.)

sucrose solution concentration 47–70°Bx. Apples were dehydrated at temperature 30–90°C [115] (Figure 28.9). It has been shown that the increase of temperature in the range of 30–80°C substantially shortens the time of dehydration [133]. However, increased temperature promotes penetration of osmoactive substance into the tissue [77].

A HTST process was proposed for osmotic dehydration by Mastrocola et al. [118], Lerici et al. [78], Levi et al. [150], and Dąbrowska and Lenart [151]. The process is conducted at 65–90°C at a time of 1–20 min. The degree of dehydration is equivalent to that at 20°C lasting for 2 h. The HTST process also gives the effect of blanching, which inactivates enzymes and removes part of the air from the intercellular space. To obtain a high ratio between water loss and solids gain (Figure 28.10), a temperature between 20°C and 40°C is recommended [117,152,153]. The degree of dehydration is regulated by the time of osmosis.

The course of mass loss versus time is curvilinear. The highest rates occur at the beginning of the process [154]. Porosity greatly increases during the first period of osmotic dehydration. This implies that a part of the air volume in the structure is replaced by the external solution [155]. The most significant changes of water content, water loss, and solids gain take place during the first 30 min of dewatering at 30°C. Rate of water loss is 5–10 times higher than the rate of solids gain and depends on the advancement of the dewatering process [156]. According to Lenart

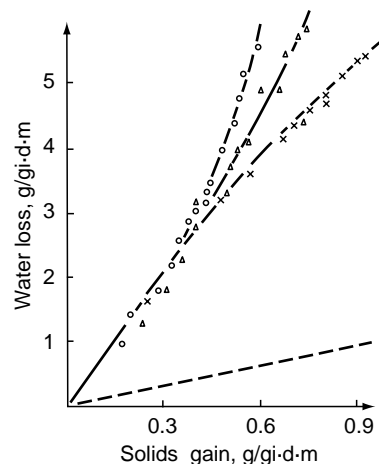


FIGURE 28.10 Relationship between water loss and solids gain in apples during osmotic dehydration in saccharose solution (○—○, 20°C; Δ—Δ, 30°C; +—+, 40°C; , diagonal). (From Lenart, A. and Lewicki, P.P., in *Drying '87*, Vol. 2 Mujumdar, A.S. ed., Hemisphere Publ. Co., New York, 1987, p. 239.)

[157], optimal time of apple dehydration is 5–6 h at 20°C, 3–4 h at 30°C, and 1–1.5 h at 40°C. A reduction of mass by 50% can be achieved after 2.5–3 h of osmosis at 50°C.

It is well recognized that the characteristics of the material undergoing dehydration by osmosis strongly affect the course of the process. Under identical conditions some materials lose water faster than others (Figure 28.11). The penetration of the osmoactive substance differs markedly (Figure 28.12); hence, properties of the product and its consumer acceptance are strongly affected by the initial properties of the material, supposedly by its tissue structure.

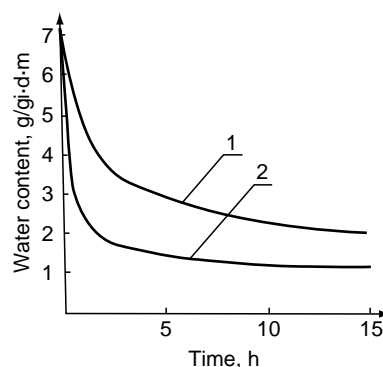


FIGURE 28.11 The effect of the kind of material on the course of osmotic dehydration in saccharose solution at 20°C. 1-apple, 2-carrot.

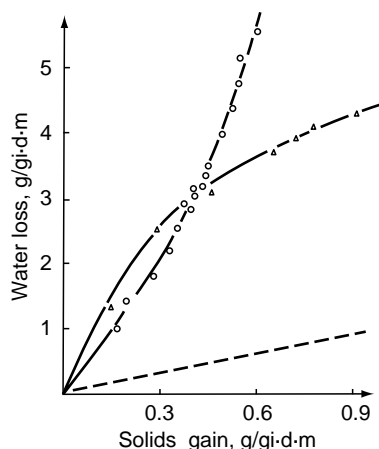


FIGURE 28.12 The effect of the kind of material on the relationship between water loss and solids gain in osmotic dehydration in saccharose solution at 20°C (○—○, apple; △—△, carrot; -----, diagonal).

Concluding the effect of procedures and processing parameters on the rate and efficiency of osmotic dehydration it can be stated that all of them are equally important. The kind and concentration of the osmoactive substance, the weight ratio of the solution to food, the kind of osmosed material, its size and shape, temperature and pressure, and the pretreatment of the material prior to osmosis affect strongly not only the course of the process but also, first of all, influence the quality and organoleptic attributes of the final product. To reach requested quality of the product the process of osmotic dehydration must be individually designed.

28.3.4 OSMOTIC SOLUTION MANAGEMENT

Water withdrawn from the material dilutes the hypertonic solution. Hence, it is important to keep its concentration constant, either by a continuous evaporation of excess water [133,144] (Figure 28.13) or by dissolution of osmoactive substance [144]. Both methods make it possible to use the same hypertonic solution several times.

Dilution of the hypertonic solution depends on the ratio between solids and the solution. At high ratio, i.e., 1:10, dilution is low and amounts to few percent, whereas at high ratio such as 1:2 the increase of volume of osmotic solution can be as large as several percent [158].

Evaporation of excess water, membrane separation, or dissolution of osmoactive substance can bring the concentration of hypertonic solution to the required value but the solution is not the same as that used in the beginning of the process. Leaching of solubles from fruits or vegetables [159,160] to the hypertonic solution changes its physical, chemical, and sensory properties. Moreover, debris of fruits and seeds are present in the solution after its use. And finally, may be this is the most important issue, the solution becomes a good medium for microbial growth.

Concentration of substances leaching from the processed material stabilizes after few uses depending on the kind of osmosed material, and reaches the level similar to that of the osmotically processed fruits or vegetables.

Designing the process of the reuse of osmotic solution the coarse filtration, pasteurization, and decolorization

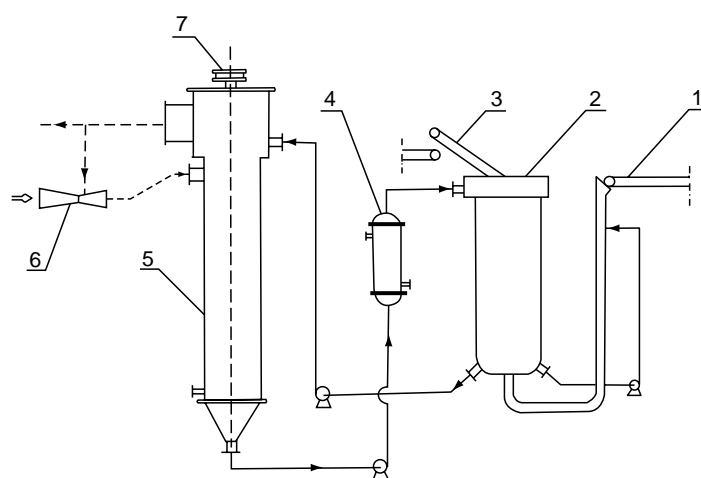


FIGURE 28.13 Osmotic dehydration with reconcentration of hypertonic solution (1) feeding conveyor; (2) osmotic dehydrator; (3) redler conveyor; (4) heat exchanger; (5) scraped surface evaporator; (6) thermocompressor; (7) driven wheel; flow of hypertonic solution; -----, vapor; ⇒, high pressure steam; -----, heating steam).

must be taken into account. The sanitation of the solution is a priority in the recycling process.

Microbial contamination of the solution comes from different sources but its water activity 0.90–0.95 limits the growth of nonosmotolerant bacteria and yeasts. Processing of fruits and vegetables results in contamination of osmotic solution with moulds, yeasts, and lactic bacteria. Reported microbial loads of osmotic solution range from 10^2 to 10^5 cfu/ml after long-time use [158]. Mild heat pasteurization is sufficient to lower the microbial load of the solution to value as low as 10^2 cfu/ml [161].

Heat treatment of sugar solutions containing acids and proteins results in nonenzymatic browning. The presence of 5-(hydroxymethyl)-2-furfuraldehyde (HMF) was shown to be a good indicator of Maillard reactions [158]. Decolorization of the solution can be done with activated carbon, charcoal as filtration coadjuvant and polyvinylpyrrolidone [132,160,161].

Filtered, pasteurized, and decolorized syrup can be used few to several times depending on the processed material and organization of the process. In continuous processing more recycling can be done in comparison to the process in which runs are not done consequently. The possibility to reuse the osmotic solution more than 20 times was reported [158]. However, number of cycles is dependent on:

- Kind of processed material
- Type of reconcentration technology
- Pasteurization parameters
- Organization of the process
- Individual adaptation to the given process

Osmotic solution, even after several uses must be disposed. In the case of fruit processing, some ways of further use of osmotic solution have been proposed [158]:

- Syrup for fruit canning
- Production of jams
- Mixing with fruit juices
- Production of fruity soft drinks
- Production of natural flavorings
- Bee feeding

Processing of vegetables, especially with sodium chloride yields solution which further management is not solved until today.

Spent solutions if not used in other processes must be discharged as wastewater. High carbohydrate content and the presence of other organic materials cause very high demand for oxygen. The biological oxygen demand (BOD_5) of the osmotic solution is high and efficient wastewater treatment is needed. The presence

of salt in osmotic solution creates additional problems and probably the use of reverse osmosis is the rational way of this spent-liquor treatment.

28.3.5 ENERGY ASPECTS OF OSMOTIC DEHYDRATION

Osmotic dehydration is distinctive in that water is removed from the product without undergoing the phase change. It offers a considerable potential for energy saving in comparison with convection drying.

Most publications consider energy consumption in a drying process that is preceded by osmotic dehydration [162] or analyze the process under laboratory conditions [142,163,164].

Energy consumption in the osmotic dehydration process arises from the following [165]:

- Heating of the material and osmoactive solution to the required temperature and making up a heat loss
- Solution mixing or pumping and recirculation, depending on the variant applied
- Dissolution of hypertonic substance in a diluted solution
- Evaporation of water in an appropriate evaporator

It is estimated that dissolution of osmoactive substance in a hypertonic solution needs some 1 kJ/kg of water removed from the material. Hence, this process affects energy consumption in osmotic dehydration negligibly.

The amount of water removed during osmotic dehydration is not large. Processing of 1 ton of fruits or vegetables per h will give, at the most, 450 kg of surplus solution (i.e., some 5.5 tons of water evaporated per day). Hence, a single-effect evaporator with vapor recompression will meet the needs.

Energy use in the evaporator consists of electric energy for syrup circulation and heat for water evaporation. Electric energy use is estimated to be equal to 10 kJ/kg of evaporated water, and heat consumption is some 1.8 MJ/kg of evaporated water.

The increase of temperature shortens the osmotic process. The use of energy for syrup mixing or circulation is estimated as 17.2, 10.0, and 4.3 kJ/kg of water removed at temperatures 20°C, 30°C, and 40°C, respectively. To keep the process running at a desired temperature, a supply of heat is necessary. Depending on the amount of water removed from the material, the heat supply amounts to 180–240 kJ/kg at 30°C and 380–500 kJ/kg of water removed at 40°C [166].

Energy consumption in osmotic dehydration of fruits and vegetables under industrial conditions is

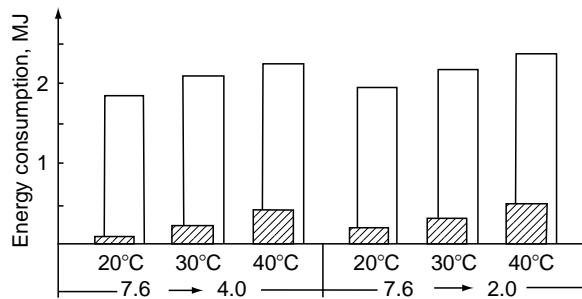


FIGURE 28.14 Energy use in osmotic dehydration expressed per kilogram of removed water. Temperature of the process and the degree of dewatering are the parameters (▨, diluted hypertonic solution completed with osmoactive substance; □, diluted hypertonic solution concentrated in evaporator).

estimated to be between 100 and 2400 kJ/kg of water removed (Figure 28.14), depending on the temperature of the process and the way the surplus solution is managed. It is worthwhile to notice that convection drying needs some 5 MJ/kg of evaporated water, which is at least twice as much as is needed in osmotic dehydration.

28.4 EQUIPMENT FOR OSMOTIC DEWATERING

Depending on the aim of osmotic process and desired product characteristics the process can be designed as far as processing parameters are concerned. To implement the designed process a special equipment is needed which must assure control of processing parameters and efficiency and economics as well.

Choice of the equipment is based on the following criteria:

- Type of processing; periodic or continuous
- Resistance of food to mechanical damage
- Shape of food; whole or cut into pieces
- Susceptibility of food to oxidation in contact with air
- Relative movement of phases, solid and liquid
- Possibility to control processing parameters
- Investment and running cost

According to Marouzé et al. [167] processes of osmotic dehydration can be categorized as follows:

- Those in which food is immersed in the osmotic solution
- Those in which solution is introduced onto the food

- Those in which osmotic substance in solid state is contacted with food
- Those in which reduced pressure is used to facilitate mass transfer

28.4.1 FOOD IMMERSED IN SOLUTION

The simplest way to contact food with osmotic solution is to immerse a basket with food into solution. The movement of solution is slight due to natural convection. Mass transfer is slow and most of processing parameters are not controlled. The method can be used to soft fruits.

Osmotic dewatering can be facilitated by decreasing mass transfer resistance. This can be done either by circulation of solution or by slow movement of food. Circulation of solution is done by installation of circulation pump to a vessel in which basket with food is immersed. Movement of food in the solution is done by vibration (Figure 28.15) or by a conveyor (Figure 28.16). The latter solution is used in Poland in the processing of apple slices [168,169].

Combination of solution circulation and movement of food particles is combined in such equipment as vibrating plate mixer (Figure 28.17) and percolated bed with slow displacement of food (Figure 28.18). In vibrated plate mixer [133], osmotic solution is circulating in two loops, one is a feed loop by which food is fed into a mixer, and the second loop maintains constant temperature of the solution. Food moves from bottom of the mixer to its top

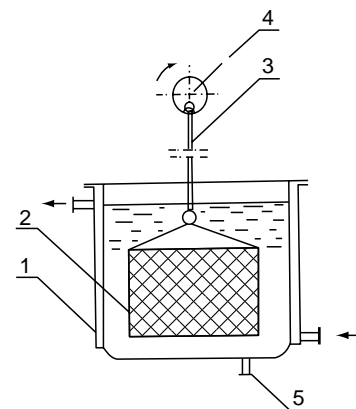


FIGURE 28.15 Osmotic dehydration with a vibrating basket (1, jacketed vessel; 2, basket; 3, shaft; 4, eccentric; 5, spout). (From Lenart, A. and Lewicki, P.P., in *IDS '89*, Mujumdar, A.S. and Roques, M., eds., Hemisphere Publ. Co., New York, 1990, p. 501. With permission.)

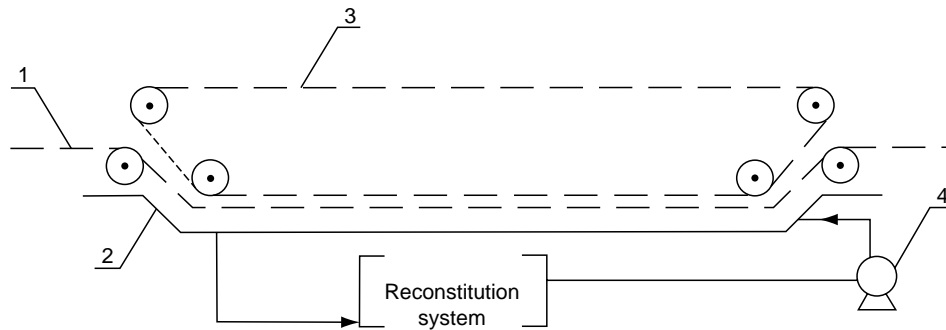


FIGURE 28.16 Conveyor osmotic dehydration (1, perforated conveyor with osmosed material; 2, vessel with hypertonic solution; 3, conveyor preventing apple slices from floating; 4, pump).

through a series of perforated vibrating horizontal plates mounted on a vertical axis. In a percolated bed, food is delivered at the bottom of the tank by a hydraulic feed and forms a porous bed. The bed moves slowly to the top of the tank and is extracted by a redler or bucket conveyor. Solution is fed at the top of the tank and is circulated through the feed leg. In this equipment, a countercurrent movement of food and solution occurs. A cocurrent movement of food and solution as a percolated bed was also designed for osmotic process [167].

Movement of solution and particles can be done by mechanical mixing. Mixing device can be installed vertically (Figure 28.19) or horizontally (Figure 28.20). In the first design, a worm screw is

placed coaxially inside a vertical cylindrical tank. The screw moves particles of food from top to the bottom of the tank. Then the pieces rise toward the surface under the buoyancy force. In the second technical solution or design, the screw is mounted horizontally. The food pieces together with the solution are moved along the cylinder axle. Pieces of food are carried in rotation toward the end of the cylinder where a deflector catches the pieces and directs them to the outlet.

Designs of equipment with mechanical motion of food pieces exert some force on processed material. Hence some disintegration and deformation of food can take place and increased pulp content in the solution can be observed.

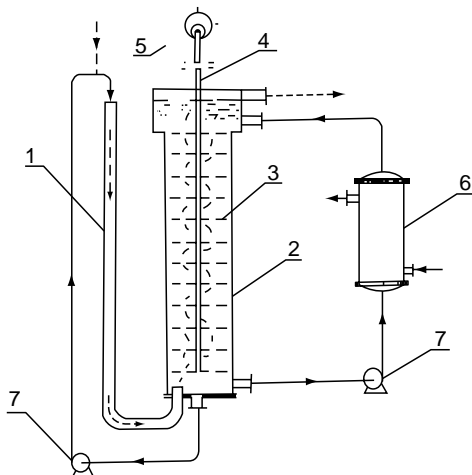


FIGURE 28.17 Osmotic dehydration with a vibrating plate mixer (1, feed leg; 2, vessel; 3, vibrating mixer; 4, shaft; 5, eccentric; 6, heat exchanger; 7, pump). (Adapted from Pavasovec, V., Stefanovic, M., and Stefanovic, P., *Drying '86* Vol 2, Mujumdar, A.S., ed., Hemisphere Pub Co., New York, 1986, p. 761)

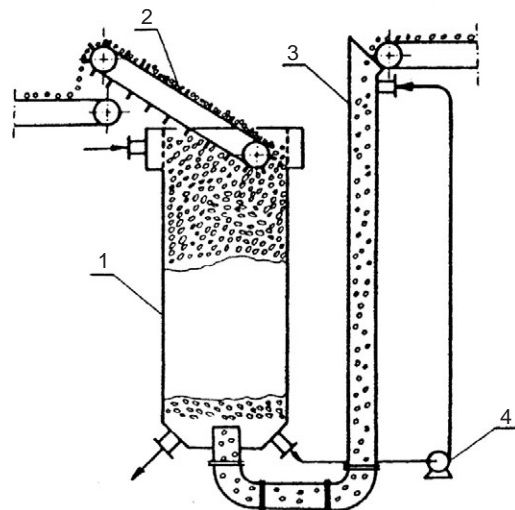


FIGURE 28.18 Osmotic dehydrator—a packed bed unit (1, vessel; 2, redler conveyor; 3, feed leg; 4, pump). (Adapted from Pavasovec, V., Stefanovic, M., and Stefanovic, P., *Drying '86* Vol 2, Mujumdar, A.S., ed., Hemisphere Pub Co., New York, 1986, p. 761)

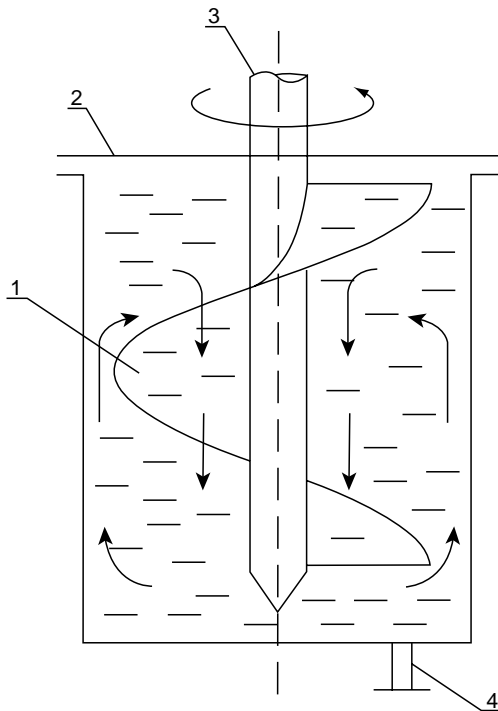


FIGURE 28.19 Osmotic dehydrator—a vertical mixer (1, worm screw; 2, vessel; 3, shaft; 4, spout). (Adapted from Marouze, A., Groux, F., Collignan, A., and River, M., *J. Food Eng.*, 49,207,2001. With permission)

28.4.2 SOLUTION SPRAYED ONTO THE FOOD

Reduction of solution and food ratio can be done by application of thin layer of hypertonic solution to food pieces. It is done by placing food pieces on perforated conveyor and spraying concentrated solution on processed material (Figure 28.21). The design is well suited to continuous processing but requires

food pieces to be spread on the conveyor in a single layer. Hence, a large area of the conveyor is needed to process any given quantity of food. This technical solution of osmotic dewatering was proposed by Le Maguer [170] and Dalla Rosa et al. [161].

28.4.3 SOLID OSMOTIC SUBSTANCE CONTACTED WITH FOOD

The lowest solution and food ratio is obtained when solid osmotic substance is contacted with food. Crystals of sugar or mixture of sugar and salt are mixed with food pieces in appropriate proportion and tumbled in slowly rotating cylindrical tank. The amount of osmotic substance used should be such that water removed from food pieces forms no solution in the tank. Wet but solid osmotic substance is separated from food on vibrating screen. However, some crystals stick to the food surface and can create problems in packaging or further processing of osmosed material.

28.4.4 EQUIPMENT WORKING UNDER REDUCED PRESSURE

Static or pulsed-vacuum processing of immersed fruit or vegetable products facilitates osmotic dewatering. Equipment used in this process can be of any type previously presented but requires hermetic design.

28.5 PRODUCT CHARACTERISTICS

Osmotic dehydration is a complex process of counter-current mass transfer between the plant tissue and hypertonic solution. This leads to dehydration of the material and changes in its chemical composition

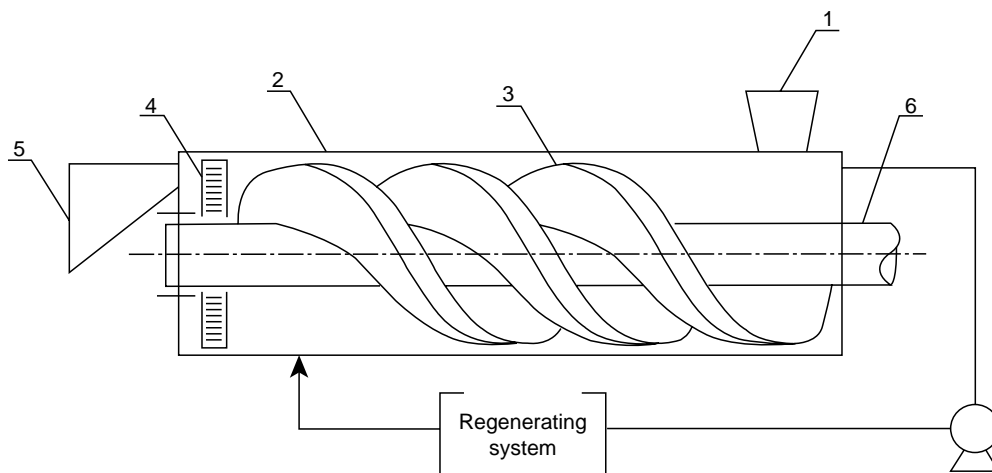


FIGURE 28.20 Osmotic dehydrator—a horizontal mixer (1, feed hopper; 2, vessel; 3, worm screw; 4, deflector; 5, discharge hopper; 6, shaft). (From Lenart, A. and Lewicki, P.P., *Zesz. Nauk. SGGW-AR, Technol. Roln.-Spoż.*, 14, 33, 1981. With permission.)

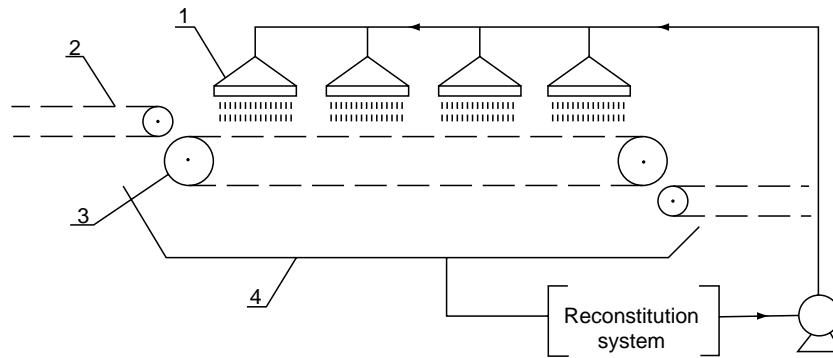


FIGURE 28.21 Spray osmotic dehydrator (1, spray nozzles; 2, feed conveyor; 3, perforated conveyor with material undergoing osmotic dehydration; 4, collector for used hypertonic solution).

as well. Hence, it must be expected that the properties of the material dehydrated by osmosis will differ substantially from those dried by convection [5,10,76,171].

The flux of osmoactive substance penetrating the osmosed tissue changes its chemical composition. It has been shown that the content of sucrose increases in cell sap during osmotic dehydration [15,67,172], and the sucrose flux is increased by the presence of sodium chloride [70]. On the other hand, use of starch syrup gives only a small influx of sugars to the material [144]. Glucose seems more effective than sucrose in the water loss and in the solids gain by fruits [102]. Sodium chloride penetrates tissue very effectively, hence contacting of the material with this substance leads to salting rather than to dewatering of the tissue [73,80]. There is also a flux of native substances leaving the tissue. Concentration of organic acids is lowered and native sugars are replaced by sucrose [67,117,173].

Penetration of an osmoactive substance, except sodium chloride, is a surface process. Sugars penetrate to the depth of 2–3 mm whereas changes in water content are observed up to the depth of 5 mm (Figure 28.22) [109,174,175]. When sodium chloride is used, it penetrates carrot tissue to a depth exceeding 12 mm.

Concentration of the cell sap and influx of osmoactive substance lower the water activity in the tissue to a value dependent on processing parameters. The water-binding capacity of the tissue is also affected by the osmotic process, although changes are observed only in surface layers [176]. Osmotic dehydration done for 0.5 h led to a sixfold decrease of water-binding forces at the surface of apple in comparison to the raw material [177]. Water removal from the tissue by osmosis shows a much stronger effect on water-binding forces than by the convection drying done to the same final water content [178].

As it has been stated previously, osmotic dehydration cannot be treated as a food preservation process *per se*. It is a pretreatment that removes a certain amount of water from the material; to achieve shelf stability, a further processing of the product is needed. Hence, the interaction of osmotic dehydration with further processing is important for quality assurance.

Use of osmotic dehydration practically eliminates the need to use preservatives such as sulfur dioxide in fruits. The process removes a substantial amount of air from the tissue, thus blanching prior to osmotic dehydration also can be omitted [166].

It has been shown that apples dried by osmosis and then frozen compared favorably with the conventional frozen fruits [144,179]. Osmotic dehydration preceding freeze drying shortens the time of the process and yields fruits superior to those not treated by osmosis [15,79,97]. Osmotic dehydration followed by

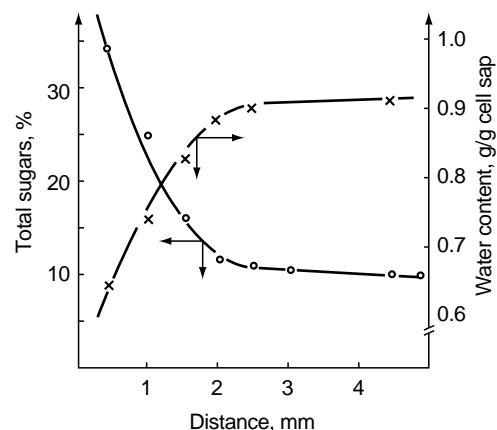


FIGURE 28.22 The depth of osmotic substance penetration and cell sap concentration in apple osmosed in 68.5% saccharose solution at 40°C for 4 h. (From Lenart, A. and Lewicki, P.P., *Zesz. Nauk. SGGW-AR, Technol. Roln.-Spoż.*, 14, 33, 1981. With permission.)

vacuum drying gives products that are very stable upon storage [65].

Most research has been directed toward combining osmotic dehydration with convection drying [75,88,180–183]. The approach is of special interest due to the growing consumer demand for commodities in the freshlike state. The IMF comply well with consumer expectations [184–186].

Osmotic pretreatment before microwave-assisted air drying increase the final overall quality of the product [187]. Fruits and vegetables treated by osmosis can be further dehydrated in a convection dryer to lower the water activity to the level of 0.65–0.90. At those water activities, water content in the material is still high and the product presents such organoleptic attributes as chewiness, softness, elasticity, and plasticity [76,188–190]. The product has a natural color, well-preserved flavor, and high retention of vitamins [132]. Its shrinkage is much smaller when compared with that observed in convection-dried products at the same water activity.

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29 Drying of Pharmaceutical Products

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29.1 INTRODUCTION

The pharmaceutical industry is one in which quality of the final product cannot be compromised. Any deterioration of the product (e.g., by microbial infection, oxidation, thermal decomposition, contamination by metallic particles or by unremoved organic solvent) must be avoided at any cost. In light of that

the Good Manufacturing Practices (GMP) for drug manufacture (see, e.g., Ref. [1]) put numerous demands on the drying stage of the drug manufacturing process. Noncontaminating dryer construction materials are used, like polished stainless steel or enameled iron. Closed-cycle dryers are often required as moisture removed is often an organic solvent or their mixture. Drying must be often

performed in inert gas to avoid oxidation or explosion if solvent is flammable. To avoid thermal decomposition in many instances vacuum and freeze drying must be employed.

All these requirements put dryers for pharmaceuticals among the most expensive and sophisticated drying equipment. At the same time the wide variety of drugs produced by any pharmaceutical company (tens of thousands of individual drugs are produced worldwide) in many different forms causes dryers in the pharmaceutical industry to encompass a wide range of types, both batch and continuous.

The manufacture of a drug that will be distributed in nonliquid form (tablets, capsules, dragees) is carried out in three subsequent stages:

1. Synthesis of intermediate products
2. Final synthesis of the drug
3. Manufacture of dosage forms

After each stage the products are dried. Selection of a proper dryer for these products depends on the properties of materials such as their form, thermal sensitivity, tonnage, drying kinetics, and so on. Since most pharmaceuticals are low-tonnage and high-value products, considerations of energy saving are usually of secondary importance in the selection of the drying process.

29.2 CLASSIFICATION OF PHARMACEUTICAL PRODUCTS WITH RESPECT TO DRYER SELECTION

From the point of view of drying technology, all substances dried in the pharmaceutical industry can be classified into three major groups:

1. Granular materials: Solids in the form of individual particles of the size approximately in the range 0.05 to 5 mm
2. Pastelike materials: Solids mixed with liquid to form a free-flowing paste; size of particles approximately in the range 0.1 to 50 μm
3. Solutions and suspensions: Solids dissolved or suspended in liquid in the form of fine (10–50 μm), ultrafine (0.1–10 μm), or colloidal (<0.1 μm) suspensions

The criteria of classification are not clear-cut here, and wet granular material of small particle size can as well be classified as pastelike and a thin paste can be classified as a suspension if it is pumpable. According to the type of the material, appropriate drying systems are chosen.

For the first group the following types of dryers are generally used: convective (tray, band, fluid bed, flash dryers, and their modifications) or contact (vacuum dryers such as double-cone dryer-blender, conical dryer with rotating helical mixer, paddle dryer). Pastelike materials are dried in tray dryers, band dryers equipped with extruding devices, and spin-flash dryers. Finally, thin pastes can be dried in spray dryers or on fluid beds or spouted beds of inert particles. Small amounts of solutions and suspensions are generally freeze-dried, especially if the product is thermolabile.

Further selection of dryers is based on the drying kinetics, which is closely associated with the structure as represented by their mean pore size and the type of the moisture–material bond. Sazhin has developed a classification based in principle on Rebinder's proposal, which classifies all materials into four groups. This classification [2], which can be used to help select a dryer, is presented in Table 29.1. Although the selection of the dryers is limited to dryers for granular materials, this classification scheme has general validity. The quantitative basis of the classification is the minimum pore diameter from which moisture is to be removed during drying.

TABLE 29.1
Classification of Granular Materials

Group	Pore Size (nm)	Drying Time in Suspended State	Types of Dryers Recommended
I	>100	0.5–3.0 s	Cyclone dryers Flash dryers Two-stage flash dryers
II	100–6	3–30 s	Two-stage flash dryers Fast spouted bed
III	6–4	0.5–2 min	Vortex dryers Batch dryers
	4–2	2–20 min	Fluid bed Vibrated fluid bed Batch dryers
IV	Ultramicropores, particle size 1–2 mm	10–60 min	Vibrated fluid bed
		40–90 min	Multistage fluid bed Batch dryers Batch dryers Suspended state dryers not recommended
	Particle size >2 mm	>90 min	

The four groups are categorized as follows:

Group I: Nonporous or capillary-porous solids with large pore sizes. Only free moisture is removed during drying. Sodium chloride and acetylosalicylic acid (ASA) belong to this group.

Group II: Uniformly and nonuniformly porous materials with pore sizes down to 6 nm. Moisture removed during drying is physically or physicochemically bonded. That is, it includes free moisture, moisture of macro- and micro-capillaries, and surface-adsorbed moisture. Phenobarbital, methenamine (Urotropin), and sodium perborate belong to this group.

Group III: Microporous or colloid-capillary-porous materials. During drying all physicochemically adsorbed moisture is removed. The following substances belong to this group: 6–4 nm subgroup, glucose and sulfadimethoxine; 4–2 nm subgroup, calcium gluconate and calcium glycerophosphate.

Group IV: Ultramicroporous materials. Pore size is comparable to the size of molecules of removed moisture. Moisture of ultramicropores is removed in intense drying, which corresponds to moisture contents as low as 0.2–0.1% or less. Chemically bonded moisture is also removed in this case.

Besides the pore size, which characterizes the drying time of individual particles, granular materials have certain flow properties that describe among other properties how easily they can be transferred into a suspended state. Materials that have large cohesion tend to form agglomerates and are difficult to dry in fluid beds or pneumatic dryers. Agglomerates have larger perimeters than individual particles, and the macropores formed between particles increase the diffusional heat and mass transfer resistance. Roughly, granular materials are classified as free-flowing and sticky materials. The degree of stickiness depends on particle size and viscosity of surface moisture. Generally the smaller the particles and the more viscous the fluid the stickier the product will be. Selection of a dryer for a free-flowing material is usually simple whereas drying of a sticky product may be a serious problem. Certain products not sticky at normal temperatures become sticky at elevated temperatures due to plasticizing or dissolving in moisture. This may lead to defluidization of a bed at a certain temperature, for instance.

Granular materials are often classified by size. According to the *British Pharmacopoeia* classification shown in Table 29.2, they are divided into six classes. This classification helps in dryer selection. Generally, very fine powders and ultrafine powders are not

TABLE 29.2
Classification of Granular Pharmaceuticals
According to Size (*British Pharmacopoeia*)

Class	Particle Size Range (μm)			
Coarse powder	100%	<1680	40%	<355
Moderately coarse powder	100%	<1680	40%	<250
Moderately fine powder	100%	<355	40%	<180
Fine powder	100%	<180		
Very fine powder	100%	<125		
Ultrafine powder	100%	<50	90%	<5

recommended for drying in fluid bed dryers owing to their very low entrainment velocities and often poor fluidization quality due to agglomeration.

29.3 PROPERTIES OF PHARMACEUTICAL PRODUCTS

29.3.1 PHYSICAL PROPERTIES

29.3.1.1 Particle Size

Pharmaceuticals are usually obtained by crystallization. Under most conditions their particle size seldom grows larger than 1 mm. Generally, owing to their low solubility, the crystals are small or very small. The particle size distribution of a typical pharmaceutical as measured by sieve analysis is shown in Figure 29.1. From the point of view of drying technology, for monodisperse materials, such as ASA and sodium *p*-aminosalicylate, the aerodynamic conditions in dryers are much easier to select. If particle size distribution is unimodal but widespread (e.g., Urotropin or phenyl salicylate) or bimodal (e.g., codeine, streptomycin, or sulfanilamide), the material is more difficult to dry in suspended state as particle segregation may occur.

Materials produced in spray dryers from solutions and suspensions are characterized by a very small particle size. Particle sizes in the range 1 to 500 μm are observed. All powders must be granulated before tableting. According to Remington [3], the following granulate size should be used to produce tablets of given diameter.

Tablet Diameter (in.)	Granulate Size (mesh)
0–3/16	20
7/32–5/16	16
11/32–13/32	14
>7/16	12

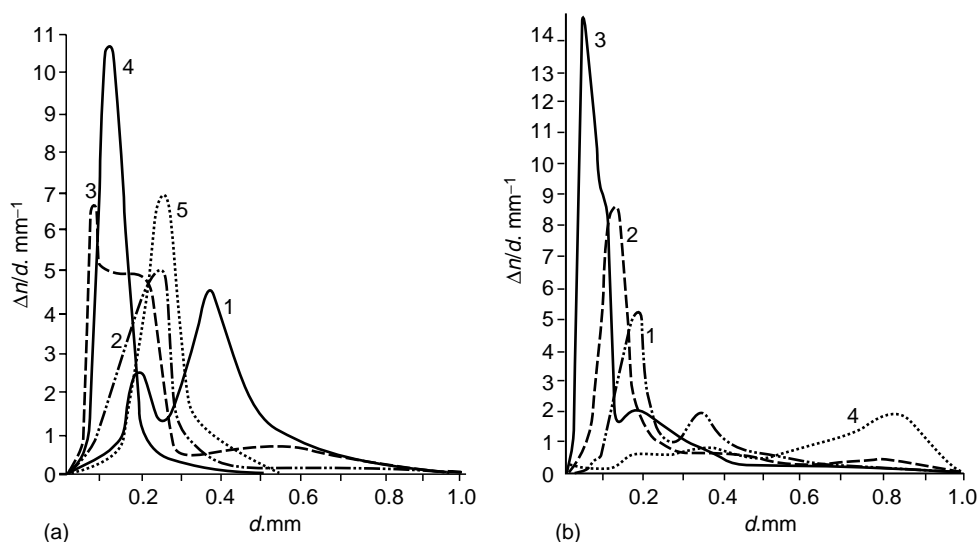


FIGURE 29.1 Size distribution of typical pharmaceutical products: (a) 1, streptomycin sulfate; 2, ascorbic acid; 3, hydrate of diaceto-2-keto-L-gulonic acid; 4, starch; 5, talcum; (b) 1, sulfanilamide; 2, acetylosalicylic acid; 3, sodium *p*-aminosalicylate; 4, codeine + terpin hydrate. (From Golubev, L.G., Sazhin, B.S., and Valashek, E.R., *Drying in Chemicopharmaceutical Industry*, Meditsina, Moscow, 1978 (in Russian). With permission.)

About 10–20% of fine particles in granulate are generally allowed as they improve the flow properties of the granulate.

29.3.2 THERMAL PROPERTIES

Among thermal properties, thermal decomposition rate is of primary interest. Usually thermal decomposition follows a first-order reaction kinetics, that is,

$$c = c_0 e^{-kt}$$

where c is the actual concentration, c_0 is the initial concentration, k is the rate constant (1/s), and t is the time (s).

The reaction rate constant depends on temperature according to Arrhenius theory:

$$k = Ae^{-E/RT}$$

where A is constant (1/s), E is the activation energy (J/mol), R is the universal gas constant (J/(mol K)), and T is temperature (K).

For example, thiamine hydrochloride in aqueous solution has $E = 92.18$ kJ/mol and $A = 9.58 \times 10^{10}$ 1/min. Processing such solution at 90°C for 10 min would result in $c/c_0 = 0.949$ (i.e., about 5% loss). One has to remember that thermal decomposition is usually influenced by water content of the processed material.

Other properties of interest include melting point temperature, specific heat, and heat conductivity of

solid. They may be found in specialized books [2,4] for most common pharmaceuticals.

29.3.3 SORPTIVE PROPERTIES

Sorptive properties of the pharmaceutical products as represented by their sorption isotherms reflect their ability to absorb water from the air and are the source of valuable information for the selection of dryers as well as the packaging conditions. The sorption isotherm is the relationship between the relative humidity of the air and the moisture content of solid in equilibrium with this air.

They are usually measured by gravimetric methods, where a known amount of solid of known initial moisture content is exposed to a gas of specified relative humidity. After sufficient time, the new weight of the sample is measured, which allows one to calculate the actual moisture content.

In most cases the sorption isotherm produced under conditions of increasing humidity (sorption) differs from that produced when humidity is decreased (desorption). This phenomenon, known as sorption hysteresis, can sometimes be very substantial. For drying purposes only desorption isotherms are of practical significance.

Sorption isotherms should if possible be measured under dynamic conditions, as in a condition of fluidization. The use of sorption isotherms measured under static conditions can lead to substantial errors when used for calculations of fluid bed, pneumatic, or

other dryers with particles suspended in the air. Sorption isotherms for a series of selected pharmaceuticals were tabulated by Reprintseva and Fedorovich [4]. Analysis of their sorption isotherms allows one to classify the products and give some indications about the selection of dryer.

Materials with intensive hysteresis for which sorption and desorption isotherms coincide only at two points, at zero and saturation humidity, belong to colloidal bodies according to Rebinder's classification. Most of the final pharmaceutical products belong to this group. All tableting formulations containing starch or gelatin are obtained by spray drying a mixture of the drug with all other tablet components, such as lubricants, binders, and diluents, which are colloid in nature.

Materials that display no sorption hysteresis, or if their hysteresis is limited (the range RH 0.2–0.9 or over a narrower range), are classified as capillary-porous bodies. As examples, ascorbic acid, *p*-aminobenzosulfamide, and hexamethylenetetramine may be mentioned here. Materials showing sorption isotherms intermediate between these two categories belong to the so-called capillary-colloid-porous bodies, and they include ASA and penicillin. The energy of the solid–moisture bond increases from capillary-porous to colloidal bodies. Adequate drying temperatures and residence times must be used for drying such difficult to dry solids. The temperature of drying is directly proportional to the energy of the solid–moisture bond. The drying time depends on particle size and the amount of moisture removed.

More information can be obtained from drying kinetics and experimental measurements. These experiments provide the ultimate basis for selection of the appropriate dryer and drying condition. Often one may arrive at several alternate drying equipment for drying a given product.

Some information on moisture–material bond type is presented in Table 29.3 for selected products. More specific information on sorption isotherms for most common materials used in the pharmaceutical industry together with measuring methods may be found in Stahl [5].

29.4 DRYER TYPES AND THEIR PERFORMANCE

High-quality standards and high price of the final product demand high reliability and justify more sophisticated construction of pharmaceutical dryers than dryers used in most other industries. Traditional dryers of the pharmaceutical industry, such as shelf or batch fluid bed dryers, are continuously supplemen-

ted by dryers first used in other fields of application. Increasing quantities of products may call for continuous dryers. Increased quality control and larger production rates continuously broaden the range of dryer types used in the pharmaceutical industry. This chapter presents only the most widely used types of dryers for pharmaceuticals. More information on performance of industrial dryers for pharmaceuticals is given by Simon [6]. It should be noted that most of these dryers are discussed in greater detail elsewhere in this handbook.

29.4.1 DIRECTLY HEATED DRYERS: BATCH DRYERS

29.4.1.1 Ovens

For small batches of pharmaceuticals, ovens are still a good choice. They allow for placing material upon several shelves, which can be carried by a truck. During drying, hot air is cross-circulated between the shelves to permit drying of different products at the same time. Internal circulating air filters are often provided (Figure 29.2) to protect against cross-contamination of the products. Similar ovens can be used for thermal sterilization of vials and bottles, for example.

29.4.1.2 Fluid Bed Dryers

Small batch fluid bed dryers (Figure 29.3) are very commonly used in the pharmaceutical industry. Owing to better air–solid contact, drying in fluid beds is faster than in tray ovens and because of good mixing product uniformity is much improved. Usually they are supplied with roll-on roll-off drying chambers often equipped with an agitator. Air after drying is filtered, usually in multicyclones or bag filters. The use of bag filters is, however, troublesome if the dryer is often used for different products as it requires very careful cleaning.

29.4.2 DIRECTLY HEATED DRYERS: CONTINUOUS DRYERS

29.4.2.1 Band Dryers

Relatively seldom used in the manufacture of final pharmaceutical products, band dryers find wide use in drying of raw materials, especially herbal and medicinal plants. Usually several bands in one-above-another configuration are used. Bands are made of stainless steel screens or perforated plates. Band speeds from several centimeters to about 0.5 m/min are used. Bandwidths vary from as low as 0.5 m up to 2 m. Drying air temperatures in the range 80–100°C, initial moisture contents of 45–100%, (d.6) and drying rates of 5–18 kg/m² h are usual in industrial practice.

TABLE 29.3
Data on Water Sorption Properties of Selected Pharmaceutical Materials

Substance	Moisture Content (% Dry Basis)							
	Physicochemical Bond			Desorption	Sorption	Desorption	Maximum Hygroscopic Moisture Content	Heat of Sorption ($\times 10^{-2}$, kJ/kmol)
	Physicomechanical Bond	Polymolecular Adsorption	Monomolecular Adsorption					
	Capillary	Osmotic	Sorption					
Penicillin	21.4	9.6	3.0	3.6	1.0	1.0	19.0	134.0
Streptomycin	20.0	10.0	9.0	12.0	6.8	6.0	25.0	158.0
Phenylsalicylate	—	—	—	—	0.18	—	1.0	58.12
Acetylosalicyclic acid	8.5	6.0	1.5	1.7	0.5	0.5	5.8	23.50
Sodium <i>p</i> -aminosalicylate	20.0	12.0	8.0	10.0	2.0	2.0	20.0	—
Ascorbic acid	13.0	13.0	0.5	0.6	0.25	0.25	2.0	8.95
Talcum	—	—	0.1	—	0.05	—	0.35	10.36
Starch	32.0	22.0	13.6	16.0	8.2	10.0	26.0	56.35

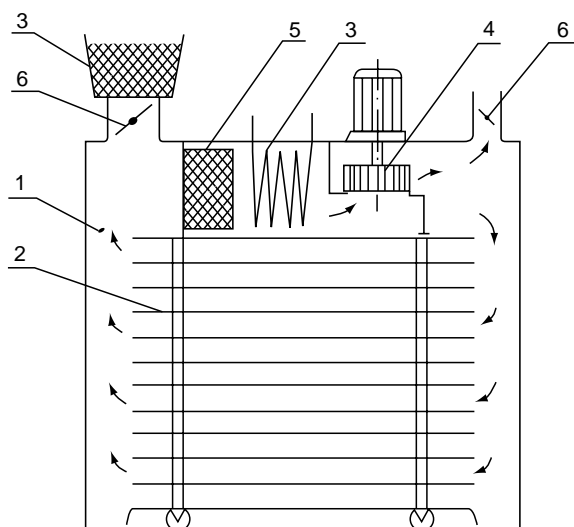


FIGURE 29.2 Shelf batch dryer (1, shell; 2, truck with shelves; 3, heater; 4, fan; 5, air filters; 6, valves).

29.4.2.2 Turbo-Tray Dryers

Turbo-tray dryers, suitable for granular feeds, operate with rotating shelves and forced convection of the air above the shelves (Figure 29.4). The product layer fed onto the first shelf is leveled by the set of stationary blades, which also scratch a series of grooves onto the layer surface. All blades are staggered so that a new surface layer is formed by each set. This way the layer is well mixed and dries uniformly. After traveling one rotation, the material is wiped off by the last blade and falls onto the lower shelf. The dryer can have up to 30 trays or more and provide large residence times. Hermetic sealing of the whole dryer is easy so that solvent recovery can be achieved.

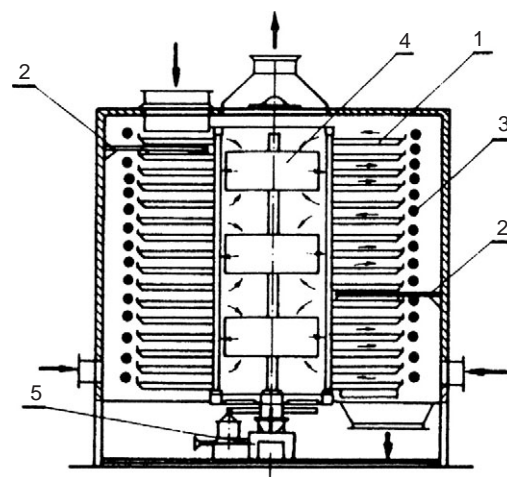


FIGURE 29.4 Turbo-tray dryer (1, shelves; 2, blades; 3, heater; 4, fan rotors; 5, shelf and fan drive).

29.4.2.3 Pneumatic Dryers

In pneumatic dryers the wet feed is introduced into a stream of hot gas flowing with velocities of several meters per second, usually in a vertical insulated duct. During their cocurrent contact moisture is evaporated, consuming the sensible heat of the gas. High evaporation rates (values as high as 240–1200-kg moisture per h and per m² of contact area are generally met) and cocurrent contact protect the particles from overheating. After a time of the order of a fraction of a second to a few seconds, the dried material and gas leave the duct and are separated in a cyclone. Very short times of contact allow drying only the unbound moisture from small particles.

To prolong the contact time with the same dryer height (or length), spiral inserts may be installed inside the dryer tube. Similar to other directly heated dryers, they can operate in a closed cycle. A closed-cycle pneumatic dryer is shown schematically in Figure 29.5.

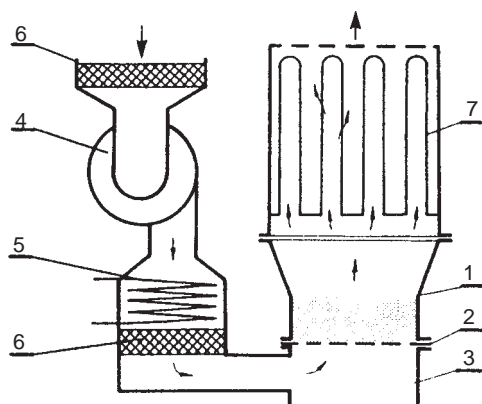


FIGURE 29.3 Batch fluid bed dryer (1, fluidizing chamber; 2, gas distributor; 3, plenum chamber; 4, blower; 5, heater; 6, filters; 7, bag dust collector).

29.4.2.4 Cyclone Dryers

Observations of pneumatic dryers proved that a considerable amount of total moisture removed evaporates in the cyclone separator. Several new dryer types were constructed that apply the principle of vortex flow. In addition, the cyclone dryers have particle separation features. Centrifugal force keeps large and wet particles rotating whereas small and dry particles can be carried away with gas. Adjustment of gas velocity can thus change the critical particle size as well as residence time in the dryer.

29.4.2.4.1 Convex-Type Dryers

The construction of convex-type dryers resembles a flat cyclone without the conical bottom (Figure 29.6).

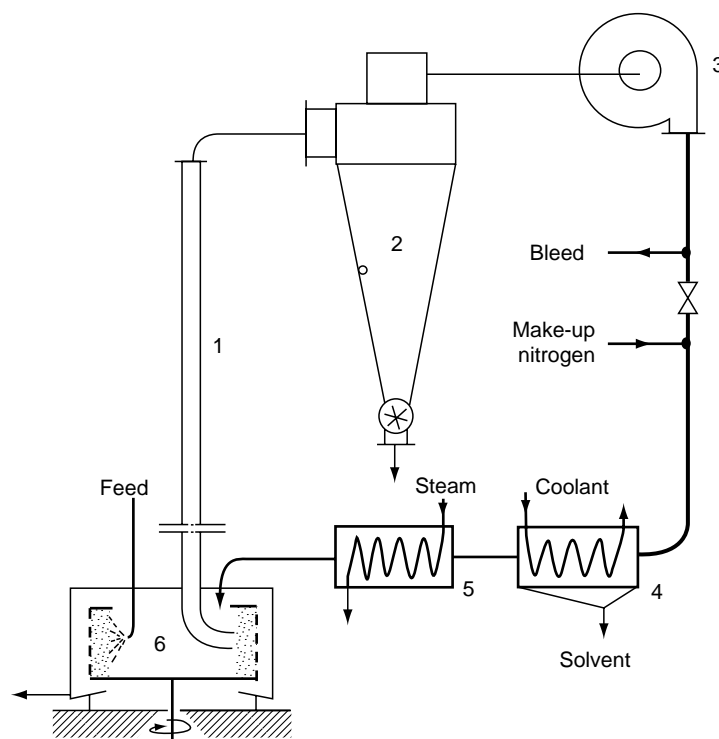


FIGURE 29.5 Closed-cycle pneumatic dryer in conjunction with a centrifuge (1, dryer tube; 2, cyclone; 3, blower; 4, condenser; 5, heater; 6, centrifuge).

A fast stream of hot gas carrying suspended wet particles is introduced tangentially into the chamber. After several spins (i.e., helical motion within the chamber), the gas and material escape through a central exhaust port and are separated in a cyclone. A characteristic feature of this type of dryer is that only small particles can be carried away as large wet particles are thrown by the centrifugal force toward the periphery. This type of dryer can be designed to

obtain a final product of required particle size. To help disintegrate the large particles, mechanical impact disintegrators are often built into the drying chamber.

29.4.2.4.2 Spin-Flash Dryers

The principle of separation of wet and dry particles is applied in the design of a spin-flash dryer for sticky and pastelike materials (Figure 29.7). In the bottom of a cylindrical chamber a high-speed impeller disintegrates the wet feed in a rapid stream of the tangentially introduced heated gas stream. The swirling gas carries away dry particles, which are then separated in a cyclone. Partially dried particles fall back into the impeller zone and are disintegrated. This type of dryer is especially suitable for thick pastes as it can handle them without dilution.

29.4.2.4.3 Countercurrent Spin Dryers

In a countercurrent spin dryer, a principle of two coaxial cocurrent swirls of gas is applied. An exemplary dryer scheme working according to this principle is shown in Figure 29.8. The cylindrical chamber is equipped with a series of side ports that introduce air tangentially into the chamber. The velocity of the side airstreams varies in the range of 5–100 m/s and is capable of producing a looser or tighter spiral, or finally a stationary circular motion of the entrained

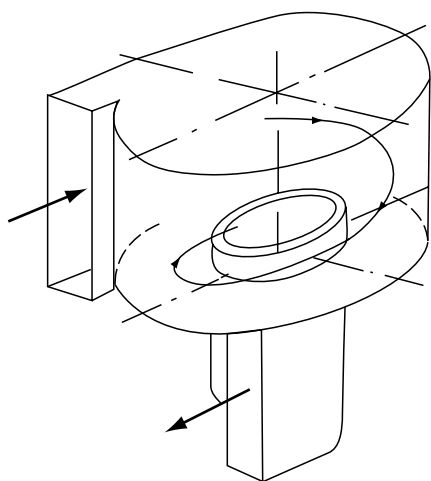


FIGURE 29.6 Schematic of a convex dryer.

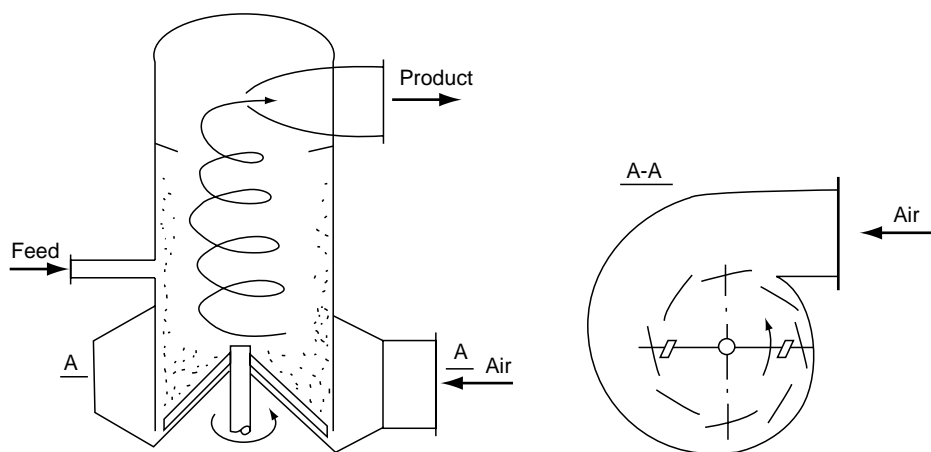


FIGURE 29.7 Schematic of a spin-flash dryer.

particles depending on the side nozzle inclination and gas velocity. In this way particle residence time can be controlled. Wet solids are introduced at the top of the dryer chamber. Drying occurs as the solids descend and finally separate from the gas, leaving the chamber through a bottom port equipped with an airlock.

29.4.2.5 Spouted Bed Dryers

A spouted bed is essentially a conically shaped chamber with a cylindrical section on top supplied with air

through a single nozzle at the bottom (Figure 29.9). The operational regime of the spouted bed dryer depends on the properties of feed and the assumed final product requirements. Polydisperse, slow-drying granular materials are continuously added and removed from the bed. Particles that undergo severe attrition or dramatically lose weight when dry can be carried away from the chamber with the heat carrier. Product size and moisture content are thus controlled by superficial air velocity in the cylindrical part of the drying chamber. Pastelike materials can be sprayed onto a spouted bed of inert material. Owing to severe attrition the dry product is removed from the particles and carried away with gas for collection in a separator.

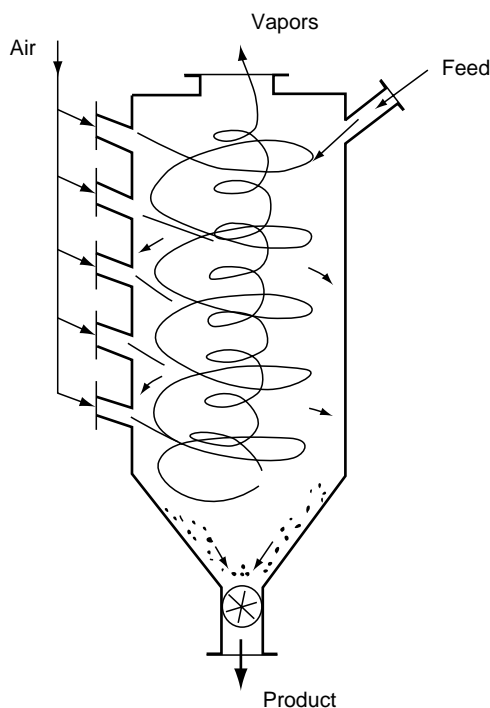


FIGURE 29.8 Schematic of a countercurrent spin dryer.

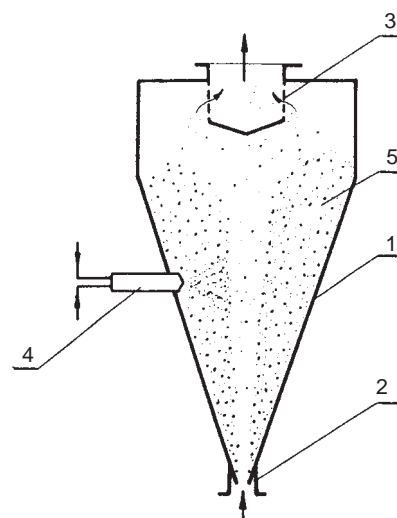


FIGURE 29.9 Schematic of a spouted bed dryer with an inert bed (1, shell; 2, nozzle; 3, particle separator; 4, spray nozzle; 5, inert bed).

29.4.2.6 Vibrated Bed Dryers

For highly polydisperse or sticky products with long drying times, the vibrated bed type of dryer offers the possibility of continuous operation accompanied by gentle material handling and uniformity of the final product moisture content. A trough vibrating in a direction slightly inclined to the vertical at frequencies up to 60 Hz and amplitudes up to several millimeters offers the possibility of simultaneous material transportation and vigorous mixing. In such a dryer, granular material can be heated directly from gas blown through a perforated trough bottom or indirectly from heating surfaces immersed in the bed. Vibratory action reduces material cohesion, which makes this dryer suitable for very wet or sticky materials. Spiral construction of the trough can, if necessary, extend the residence time of the material up to 30 min. These dryers can operate with virtually any gas velocity up to the entrainment limit. The gas velocity can be adjusted to the particle size. In direct drying conditions, evaporation rates 30–100 kg/h·m² of grid area are observed. If the amount of heat carried with gas is not adequate for drying, additional contact heaters can be employed.

Radiant heating by infrared radiation (IR) radiators can also be used. However, in this case only very thin beds can be uniformly heated. For glutamic acid with material load 15 kg/m² and radiant heat flux 3.5 kW/m², evaporation rates up to 4.5–9 kg/m²·h have been observed.

29.4.2.7 Fluid Bed Dryers

Continuous fluid bed dryers are generally built in two versions: vessel-like ideally mixed dryers and trough-like plug-flow dryers. Perfectly mixed fluid bed dryers are characterized by a rapid decrease of the feed moisture content as it is diluted in the large volume of the relatively dry bed. Therefore, they can handle relatively high initial moisture content feedstock. Trough dryers are much more sensitive to initial feed moisture content since the solid's moisture decreases gradually with dryer length. As expected, they provide a much better residence time distribution (RTD) of solids in the dryer. Fluid bed dryers can be built virtually in any required scale. Besides drying they can be used for cooling and granulation. Their modifications, such as vibrated, stirred, and pulsed fluid beds, can be used to dry or process polydisperse and relatively sticky materials.

29.4.2.8 Spray Dryers

Solutions and thin slurries of drugs can be spray dried by contact with hot air. Particles of dry product

obtained in such a way are very fine, so after mixing with other components and granulation they can be used directly for tableting. When the spray drying operation is cocurrent, that is, hot air is introduced into the dryer close to the atomizing device, there is no danger of overheating as evaporation rates are high (34–160 kg/h·m² of particle area). Thus, highly heat-sensitive materials can be spray dried.

A typical arrangement of an open-cycle spray dryer for pharmaceuticals is shown in [Figure 29.10](#). Gas–liquid nozzles are commonly used to spray feeds that are solutions. However, in the production of certain pharmaceuticals, such as antibiotics, the powders obtained by spray drying low-concentration aqueous solutions have low bulk density. Higher bulk density can be obtained if the feed is partly precipitated. High bulk density antibiotics are produced from suspensions of precipitated substrates in organic solvents; this requires the use of a closed-cycle dryer to recuperate the solvent and also an inert gas as drying medium to avoid the risk of ignition. Several pharmaceutical products after filtration form pastes that can be spray dried if they are pumpable. Pastes that are too thick must be thinned with solvents or by disintegration of their structure, as in the case of thixotropic pastes. From a practical point of view, disk atomizers are better capable of handling thicker pastes than are two-fluid nozzles.

In many cases, when inert gases are used for drying or valuable solvents must be recovered, a closed-cycle dryer can be designed according to the rules explained in [Section 29.6](#).

29.4.3 INDIRECTLY HEATED DRYERS

29.4.3.1 Drum Dryers

Drum dryers are frequently used for drying slurries and thin pastes, especially those that easily adhere to the metal surface and therefore are difficult to dry in other dryers. The slurry or paste is fed onto the drums by means of various types of feeders. Some of the standard feeding arrangements and recommendations for their use are indicated in [Figure 29.11](#). The dry product is removed by doctor blades. Depending on the material properties, the product is removed in the form of powders, flakes, or webs. The drum can be entirely enclosed in a hood and supplied with the necessary amount of drying gas. Thus, organic solvents can be recovered in a closed-cycle operation.

29.4.3.2 Vacuum Dryers

Materials that are thermolabile or easily oxidizable, especially when wet, can be dried in vacuum dryers.

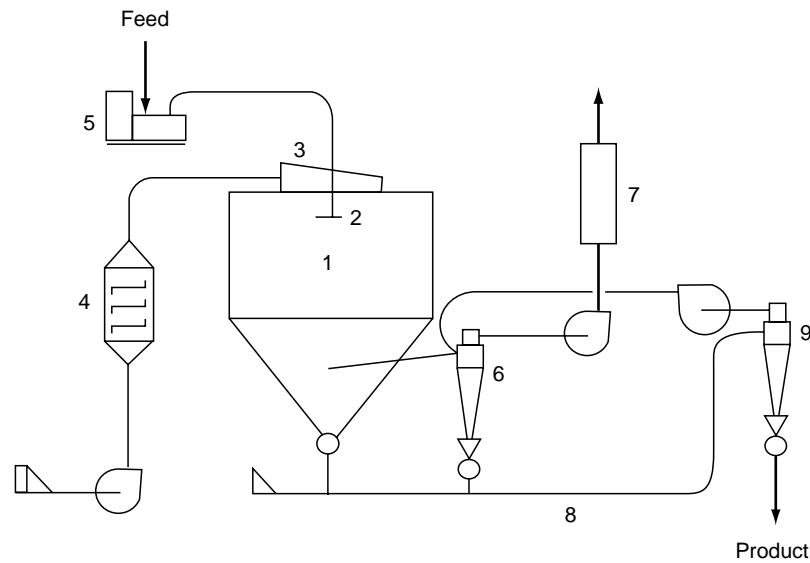


FIGURE 29.10 Spray dryer installation (1, drying chamber; 2, atomizer; 3, air dispenser; 4, air heater; 5, feed pump; 6, main cyclone collector; 7, wet scrubber; 8, pneumatic conveying system; 9, conveying cyclone). (Courtesy of A/S Niro Atomizer, Soeborg, Denmark.)

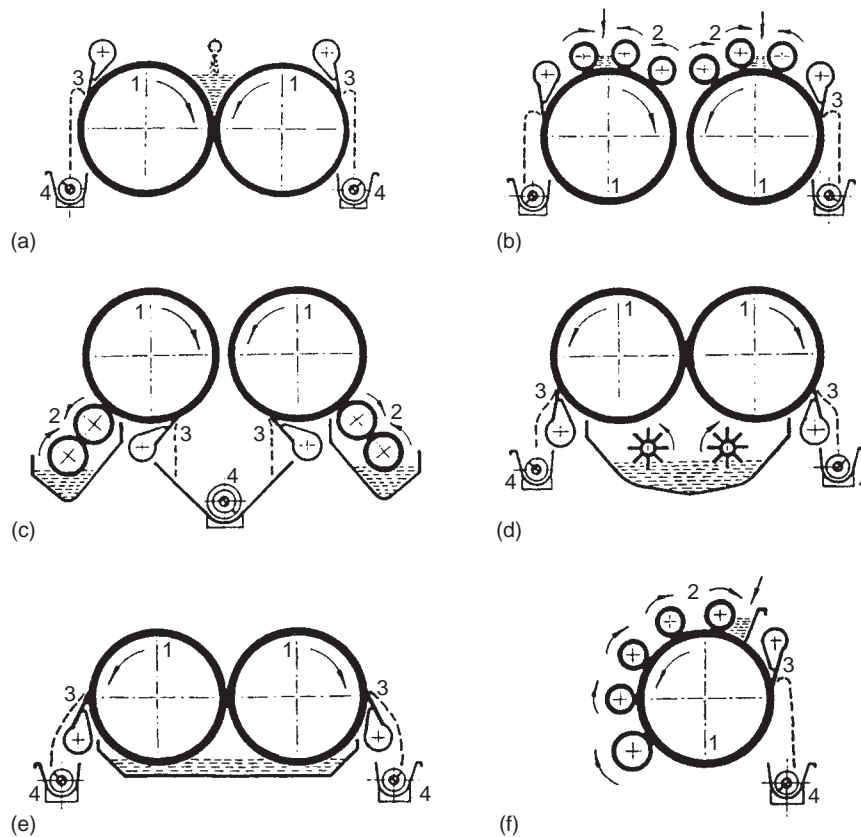


FIGURE 29.11 Some standard feeding arrangements for drum dryers: (a) nip feed, suitable for thin solutions; (b) feed roll, suitable for glutinous materials, such as starch; (c) double applicator roll, for heat-sensitive materials; (d) splash feed, used for slurries; (e) dip feed, for suspensions; (f) multiple applicator roll, used for increasing film thickness (1, drums; 2, applicator rolls; 3, doctor blades; 4, conveyors). (Courtesy of R. Simon and Sons, Basford, England.)

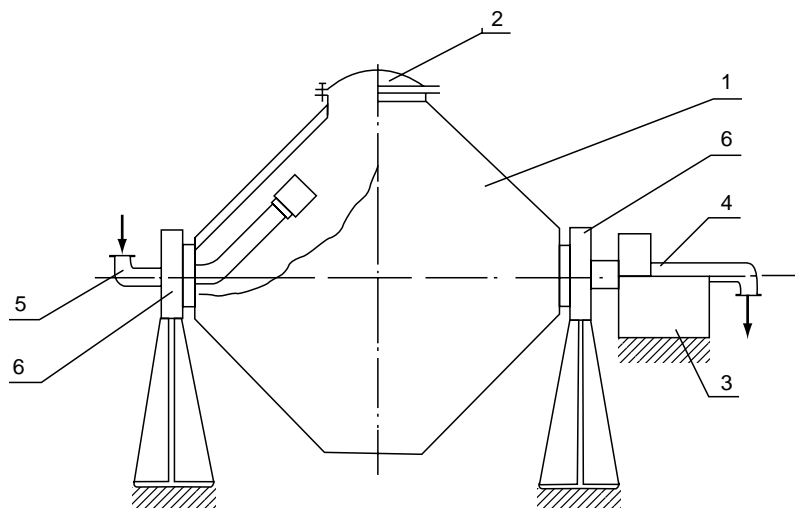


FIGURE 29.12 Double-cone batch vacuum dryer (1, steam jacket; 2, hatch; 3, drive; 4, vacuum ducts; 5, steam supply; 6, bearings).

Various types of vacuum dryers are in use for drying pharmaceutical products.

29.4.3.2.1 Double-Cone Dryer-Blender

Operated in batch mode, the double-cone dryer-blender dryer (Figure 29.12) is suitable for drying wet granular materials. Owing to the intense blending that occurs during the drying cycle, they can be used for preparation of the tableting formulations. Commercial double-cone dryers have capacities up to 10 m^3 . They may be glass lined to provide high purity of the product. They are especially suitable to perform consecutive chemical reaction-drying cycles. Reagents can be supplied through a feed tube or through a hollow shaft. The heat necessary to evaporate the moisture is supplied from steam-heated dryer walls. Drying rates observed here are in the range $2\text{--}7 \text{ kg/h}\cdot\text{m}^2$ of the heated area.

29.4.3.2.2 Conical Dryer with Screw Mixer

The conical dryer with screw mixer is similar in principle to the double-cone dryer-blender although the vessel is stationary and solids are mixed by epicyclic screw mixer (Figure 29.13). For pressures from 25 to 150 mbar and wall temperatures from 40 to 60°C , Simon [6] reports evaporation rates of water up to $10 \text{ kg/h}\cdot\text{m}^2$. Units up to 5 m^3 are commercially available. The dryer has no dead spaces and is easy to clean.

29.4.3.2.3 Paddle Dryers

Continuously operated dryers of the paddle type (Figure 29.14) provide drying times up to several hours. Suitable for pastelike and granular materials, they are steam heated. A horizontal rotating shaft fitted with paddles scrapes the product from the

walls and mixes and transports it along the dryer length. Solvents can be fully recovered if a suitable condenser for vapors is provided. Evaporation rates of water up to $10 \text{ kg/h}\cdot\text{m}^2$ are observed.

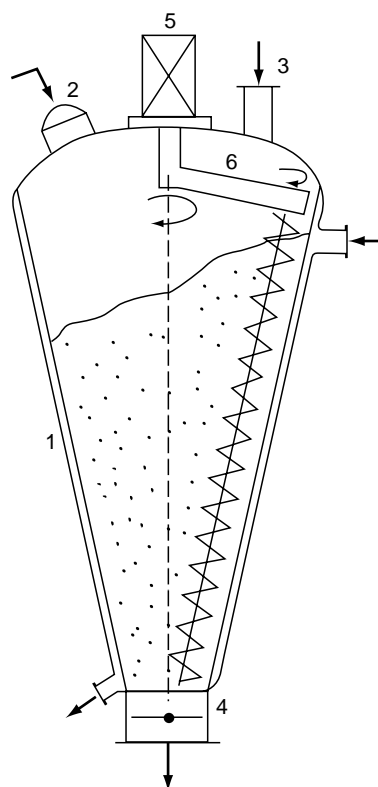


FIGURE 29.13 Conical vacuum dryer with epicyclic screw mixer (1, steam jacket; 2, hatch; 3, feed tube; 4, exit port and valve; 5, motor; 6, drive arm; 7, mixing auger).

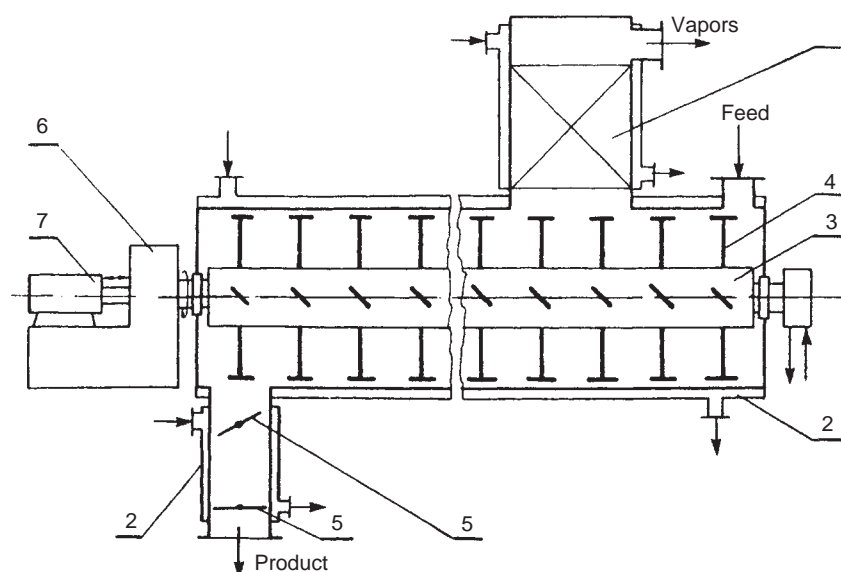


FIGURE 29.14 Continuous vacuum paddle dryer (1, vapor filter; 2, steam jacket; 3, shaft with paddles; 4, paddles; 5, valves; 6, shaft drive; 7, shaft oscillator).

29.4.3.2.4 Vacuum Band Dryers

Vacuum band dryers are suitable for all types of pastelike materials. Wet material is dried as it is transported on a moving band. Heat is supplied by infrared low-temperature radiators or by contact with heated bands. A multiband vacuum dryer is shown in Figure 29.15. To allow continuous production, such dryers are equipped with an automatic discharge system that prevents dehermetization of the dryer

chamber. Drying chambers that provide drying surface areas up to 120 m² are available commercially.

29.4.3.2.5 Filter Dryer

The filter dryer is a combination of two operations in one vessel. Usually it has a cylindrical form with a bottom that serves as a filter. It is equipped with a paddle agitator that can be heated internally or heat may be transmitted through jacketed walls. In the

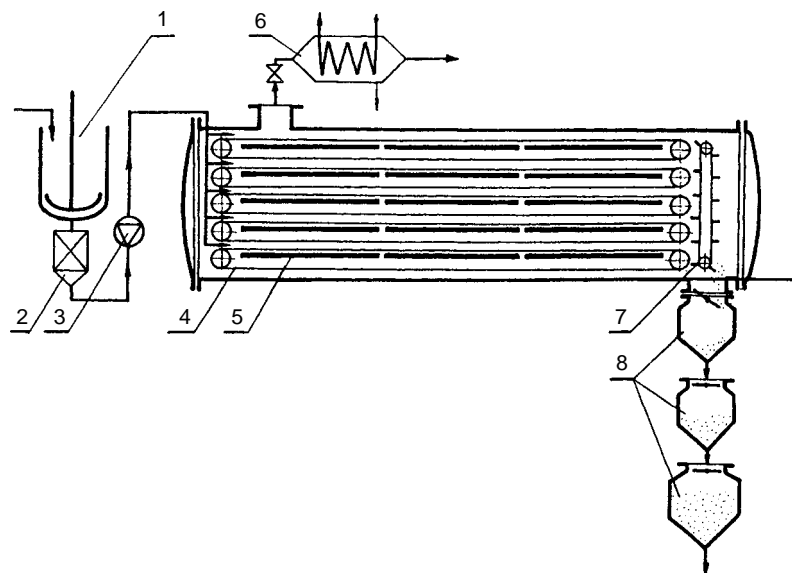


FIGURE 29.15 Band vacuum dryer installation (1, feed mixer; 2, filter; 3, feed pump; 4, band; 5, heating panels; 6, vapor condenser; 7, scraper; 8, product collector system).

latter case, the vessel is turned upside down after filtering. Units of filter area up to 15 m² and volumes up to 22.5 m³ are built.

29.4.3.3 Freeze Dryers

Freeze drying is the best method for drying highly heat-sensitive or easy oxidizable materials. However, its selection must be governed by economic considerations. Early applications of this method were the drying of blood plasma and serum during World War II.

The high cost of freeze-drying discourages its wider application. In freeze-drying, solutions of pharmaceuticals dosaged into vials are frozen and placed in a vacuum chamber. Ice sublimates, consuming the latent heat of sublimation, supplied from heated shelves. Recently, microwave heating has also been used in freeze dryers. The vapor is continuously removed by a vacuum pump. After all moisture sublimates, a very fine porous structure remains that can be easily rehydrated. It was noticed that solutions diluted below 1% (weight) concentration of solids do not produce such structure. The structure is labile and can collapse if its temperature is raised above a certain temperature of collapse. This temperature can be as low as -40°C (e.g., for glucose and fructose) or -10°C (e.g., for starches and proteins) and increases with the increase of molecular weight of the lattice and its dryness. To protect against even slight rehydration the freeze-dried vials are immediately stoppered, preferably by internal devices without decompression of the chamber.

Among other requirements that freeze dryers for pharmaceuticals have to fulfill, the actual trend is that they should provide the following range of performance parameters (Table 29.4) [7]. Under the conditions referred to as normal, most of the pharmaceuticals can be dried. Under normal conditions, with shelves fully covered with vials at an operating temperature of 60°C, a drying rate of about 1 kg/m²·h may be expected. Peak evaporation rates can reach 1.5 kg/m²·h for an operating temperature of about 70°C. Because of the critical drying conditions met during drying of pharmaceuticals, the whole drying system (Figure 29.16) and its components must meet the following requirements:

Chamber: Should if possible be built of easy-to-clean and noncorrosive materials and provided with a vial-stoppering facility.

Shelf heating and cooling: In order to protect against product melt-back, dryer shelves must be frozen during loading and chamber evacuation. Further, during drying they must be

TABLE 29.4
Performance Parameter Ranges for Pharmaceutical Freeze Dryers

Parameter	Normal Conditions	Special Conditions
Chamber pressure (μmHg)		
Peak	100–200	50
Final	20–50	5–10
Shelf temperature (°C)		
Lowest	-40	-50
Highest	70	80
Shelf heat-up rate (min), from -40°C to 70°C	90	60
Shelf cool-down rate (min), from 22°C to -44°C	90	60
Evacuation rate (min) to reach 100 μmHg	15–30	5–10
Vapor handling capacity (kg H ₂ O/h·m ²)	0.98	1.47
Total shelf area (m ²)	11–25	0.75–3.5

heated, preferably with the possibility of temperature control.

Vacuum pump: In a typical dryer with the shelf-freezing feature, the vacuum pump should have a capacity of about 50 L/s·m² of the shelf area. As leakage of air is inherent in any practical vacuum system, this capacity is used to maintain the desired operating pressure. If, however, the shelves do not have the freezing capacity, the vacuum pump should have a much higher capacity to provide nearly instantaneous dryer evacuation.

Condenser: All freeze dryers are equipped with condensers to remove water vapor from the gases coming out of the drying chamber.

Instrumentation and control: Provide monitoring and control of pressure and temperature at strategic points in the dryer, which includes the control of shelf temperature and often the vacuum if product sublimation is possible.

Only the most sensitive and expensive products are usually freeze-dried. Among these are antibiotics, antitoxins, antisera, certain vitamins, cancer chemotherapy drugs, diagnostic reagents, and other unstable and easily oxidizable materials. Because of the high value of the dryer load, they are often equipped with standby vacuum pumps and refrigeration units, as well as a standby electrical power supply and control instrumentation to protect against any failure.

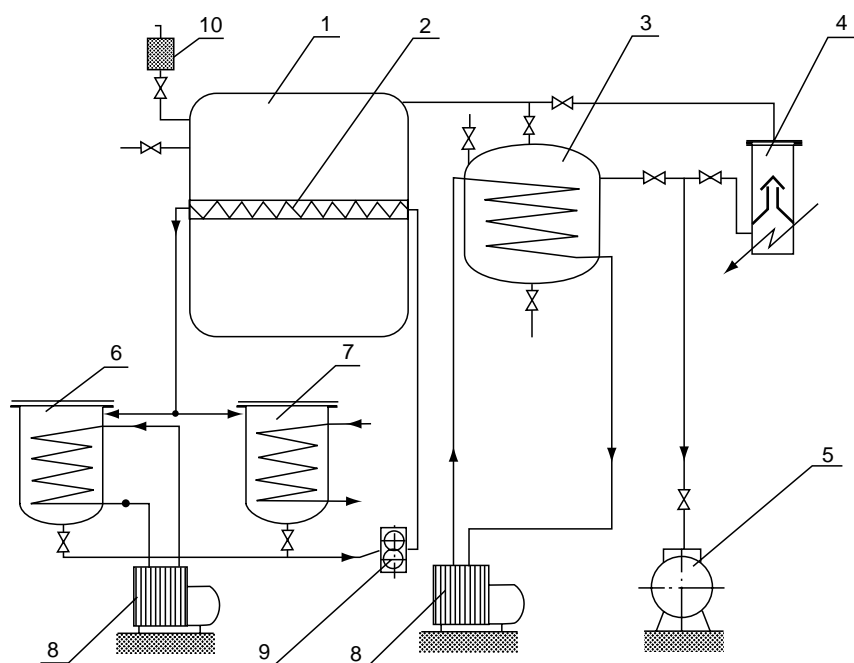


FIGURE 29.16 Freeze-drying system (1, drying chamber; 2, heated or cooled shelf; 3, ice condenser; 4, diffusion vacuum pump; 5, rotary vacuum pump; 6, tank for cooling shelf circulating liquid; 7, tank for heating shelf circulating liquid; 8, freezing aggregate; 9, circulation pump; 10, air filter).

29.4.4 GRANULATION AND DRYING

To have proper solubility the tablets are formed from very finely dispersed drug mixtures. However, the process of tableting is not smooth if these powders are fed directly to the tableting machines. Granulation of the powders before tableting is usually required. Also, several pharmaceutical products, such as glucose preparations, are sold in granulated form.

Granulation may be carried out as a dry or a wet process. Agglomeration of particles of very fine (i.e., few micrometers in size) dry powders can occur owing to van der Waals forces of attraction. Aspirin, acetophenetidin, thiamine hydrochloride, ascorbic acid, and others can be granulated using a dry process.

For wet granulation, a suitable liquid is sprayed onto the vigorously mixed powder. Adhesion of particles takes place owing to the development of liquid bridges and capillary forces between particles. In the drying phase following the granulation process, solid bridges develop owing to crystallization of dissolved substances.

Spray granulation may be carried out in fluid beds or vibrated fluid beds as a separate process, or this may be performed in the dryer itself as the final stage of the drying process. Figure 29.17 illustrates application of a vibrated fluid bed dryer as a second stage of a combined drying and granulation process. In the first stage the material is spray dried. Residual

moisture is removed from the powder in a vibrated fluid bed dryer. In the same dryer the powder is spray granulated and the granules are dried to the required final moisture content.

In an alternate method of agglomeration, saturated steam is blown for a specific period of time through the fluid bed of cold particles. Condensation of moisture on the particles forms agglomeration centers and allows formation of granules of the product during fluidization.

29.5 DRYING OF DOSAGE FORMS

Final drug formulation is a mixture of active drugs with a suitable excipient and necessary additives (binder, color, aroma, etc.). The mixture may be obtained directly through spray drying of solution or suspension; otherwise, crystalline substances must be carefully ground and blended. Powdered preparation is usually converted into granules, tablets, or dragees. Granules obtained by wet granulation require postgranulation drying, which may be performed in the same unit (see Section 29.4.4). Tablets usually do not require drying although they contain some water that was present in powder to improve tableting. Dragees, which are essentially lacquered tablets, need drying as a final stage. Their coating is sprayed usually in many layers, including

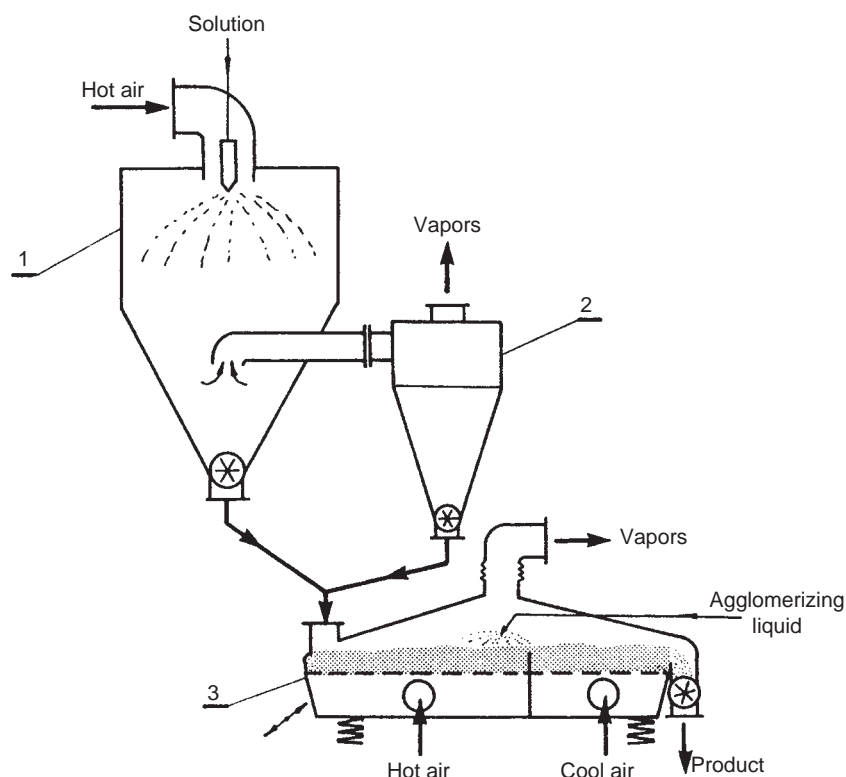


FIGURE 29.17 Combined spray drying-agglomerating system (1, spray dryer; 2, cyclone; 3, vibrated fluid bed dryer-granulator).

an antioxidant layer, dissolving rate-limiting layer, and taste and color layer. Coating layers may be water or solvent based. Dragees may be dried in a batch dryer with flow of air through the bed or in a continuous fluid bed. These dryers are integral parts of dragee-coating machines.

29.6 SOME TECHNOLOGICAL DATA ON DRYING OF PHARMACEUTICAL PRODUCTS

The following is a short summary of some technical information on drying of some of the more commonly encountered pharmaceuticals and intermediate products.

Table 29.5 presents a summary of the various types of dryers, together with examples of pharmaceutical products for which they are used commercially. It should be noted that in most cases alternate dryers are used in practice to dry the same product. Typical operating data for drying of selected pharmaceuticals are given in Table 29.6. The information contained in Table 29.6 is derived from operating dryer performance data as well as from laboratory-scale experiments. Laboratory and often pilot-scale tests are necessary before a commercial-scale dryer

may be designed with confidence. For pharmaceutical products, laboratory tests are performed to provide data on the thermal sensitivity, oxidizability, stability, and final product moisture content. This forms the basis for the selection of a dryer and process parameters. If the product is produced in small quantities, a batch dryer may be selected. In large-scale production, energy losses, losses due to deterioration of the product quality, and other losses can be quite substantial if the dryer type and operating parameters are not optimally selected. To show how the dryer type influences the drying kinetics, several drying rate curves for selected pharmaceuticals are presented in Figure 29.18.

29.7 ASEPTIC CONDITIONS IN PHARMACEUTICAL DRYERS

The problem of maintaining aseptic conditions exists in all stages of pharmaceutical production. The final stages of the pharmaceutical production process, drying, tableting, and packaging, are especially susceptible to microbial infection. High-sterility standards are required for pharmaceuticals of biologic origin, standards, laboratory reagents, and others. The two possible sources of bacterial infection during drying

TABLE 29.5
Applications of Different Pharmaceutical Dryers

Type	General Application	Example Product
<i>Directly heated</i>		
<i>Fixed bed</i>		
Ovens	Small batches of all types of pharmaceuticals	Various products
Band dryers	Organic raw materials, preformed pastes	Various products
Turbo-shelf dryers	All kinds of medium and coarse granular materials with long drying times	Penicillin, ascorbic acid, mannitol, monosodium glutamate, nicotinic acid, riboflavin, sorbitol, starch
<i>Suspended state</i>		
Pneumatic dryers	Removal of unbound moisture, narrow particle size	Ascorbic acid, hydrate of 2-keto-L-gluonic acid, <i>p</i> -aminobenzosulfamide, tetracycline sulfathiazole, aminopyrine, ASA
Cyclone dryers	Removal of unbound and bound moisture, possible self-adjustment of particle size, longer residence times than in pneumatic dryer	Thiamine bromide, phthalimidopropionitrile, ascorbic acid, folic acid, nicotinic acid
Spouted bed dryers	Long drying times, increased thermal sensitivity, suitable for drying pastes on inerts	ASA, hexamethylenetetramine, 2-amino-5-ethyl-1,3,4-thiodiazole
Vibrating bed dryers	Granular polydisperse materials, long residence time	Nicotinic compounds, sorbose, ascorbic acid, lactose
Fluid bed dryers	Medium and coarse granular materials, possible granulation	Vitamins, lactose, glucose, sodium glutamate, Urotropin, antibiotics (oxytetracycline), paracetamol, pancreatin powder, ASA
Spray dryers	Pastes, solutions, and suspensions	Aminosalicylic acid, bacitracin, blood plasma, blood serum, methicillin salts, culture media, dextran, enzymes, gamma-globulin, hormones, streptomycin, iron dextran, lysine, casein hydrolysate, penicillin, serum hydrolysate, penicillin, serum hydrolysate, tetracycline vitamins, oleandomycin, chloramphenicol succinate salts
<i>Indirectly heated</i>		
Drum dryers	Thin pastes and slurries	Calcium panthothenate, streptomycin sulfate
Vacuum dryers	All types of heat-sensitive materials	Various products
Freeze dryers	Extremely heat-sensitive materials in solutions and suspensions	Antisera, antibiotics, antitoxins, vitamins, cancer chemotherapy drugs, various diagnostic reagents and standards

are the feed and air used for direct drying. Proper sanitary handling of the feed must be maintained throughout the whole production cycle. If the drug is not thermodegradable, sterilization by superheating can be applied. If the final product is thermolabile, thermal sterilization of the raw materials sometimes can be applied or chemical, ultraviolet, or radioactive means of sterilization should be used.

For all high-sterility manufacturing operations, class 100 air cleanliness according to U.S. Federal Standard 209A (1966) must be maintained. High-efficiency air filtration and laminar flow units must be installed in packaging areas. Removal of the airborne microbes from the air that feeds the dryers and packaging areas is usually performed by

high-efficiency particulate air (HEPA) filters. They are claimed to remove up to 99.97% of all particles larger than 0.3 μm , that is, most of the known bacteria (but not viruses).

To protect HEPA filters against premature clogging, air filtration is usually multistage. At least one prefilter is added to remove dust from the air before it enters the HEPA filter. [Figure 29.19](#) shows a schematic diagram of a typical aseptic spray dryer system.

If drying is performed at low temperatures, air washing with sterilizing liquid also may be used. This is also applicable to air supplied to the packaging areas. Prewashing of the air with antibacterial liquids can remove up to 95% of all viable microorganisms present.

TABLE 29.6
Process Parameters for Drying of Selected Pharmaceuticals

Material ^a	Dryer Type, Bed Height	Moisture ^b	Initial Moisture Content ($\times 10^2$ kg/kg)	Final Moisture Content ($\times 10^2$ kg/kg)	Initial Air Temperature (°C)	Final Air Temperature (°C)	Dryer Throughput (kg/h)	Evaporation Rate (kg/m ² ·h)	Residence or Drying Time	Unit Air Consumption (kg air/kg material)
AET	Batch fluid bed		17–18.6	0.5	150–155					
	Batch fluid bed, 100 mm		20–23.5	0.5–1	90–95					
	Batch fluid bed, 100 mm				110–115					
Aminopyrine	Combined ^d		15	0.68	95	45		86		83 ^c
Antibiotics	Batch fluid bed		19.3	0.8	70				80 min	
ASA	Batch fluid bed		15	1	40				20 min	
Ascorbic acid	Combined		6.34	0.06	170	60		35.7		
	Vortex	Ethanol	12	0.35	90	55			45 s	10.3
Azorybitylamine	Combined		50	0.5	160	65		252		
Barbituric acid	Combined		30	0.2	250	65		109.7		
Biomycin	Pulsed fluid bed		25–48	0.05					1–1.5 h	
DKGA	Combined fast spouted bed		15	0.5	45	38		65.3		117 ^c
			11	0.2	55				10 s	
Folic acid	Vortex		20	0.4	80	50			52 s	10.6
Nicotinic acid	Combined		20.9	0.005	100	60				
	Vortex		20	0.55	120	80			36 s	16.6
	Vortex		27	0.34	130	78			38 s	15.0

	Vortex		28	0.54	130	80		42 s	14.0
	Vortex		27.6	0.36	150	98		57 s	13.6
	Vortex		27.8	0.43	160	105		62 s	12.0
Oxytetracycline	Pulsed fluid bed		18–24	3–5				1.5–2 h	
Phthalimidopropionitrile	Vortex		18	0.3	80	52		45 s	10.2
Piperazine	Combined		30–5	5–0.5	90–160	50–80	40–79		30–50 ^c
Riboflavine	Combined		40	0.75	130	80	150		
Sorbose	Combined		6	0.2	124.7	32.8	62.7	29.2	
Sulfaguanidine	Pneumatic		14.8	8.93					0.3–0.5 h
Terpin hydrate	Pulsed fluid bed		20–28	0.1–0.3	130	50		80.1	58.8 ^c
Tetracycline	Combined		30	18	130			20	
	Pneumatic		30–35	15–12					0.3–0.35 h
Granulated	Pulsed fluid bed		24–26	12–14					0.4–0.5 h
Hydrochloride	Pulsed fluid bed		17–21	1–2					
Thiamine bromide	Vortex	Ethanol	15	2	65	46			
			16	1.8	80	50		38 s	13
			18	1.5	90	55		36 s	12
			20	0.6	100	70		23 s	11
			20	0.4	100	75		42 s	10
Vitamins	Turbo-tray		20	5			90	30 s	8

^aAET, 2-amino-5-ethyl-1,3,4-thiodiaz ASA, acetylosaloicyclic acid; DKGA, hydrate of 2-keto-L-gluonic acid.

^bMoisture removed is water if not otherwise indicated.

^cIn kg air/kg water evaporated.

^dFree fall of extruded particles in countercurrent with air and fluid bed.

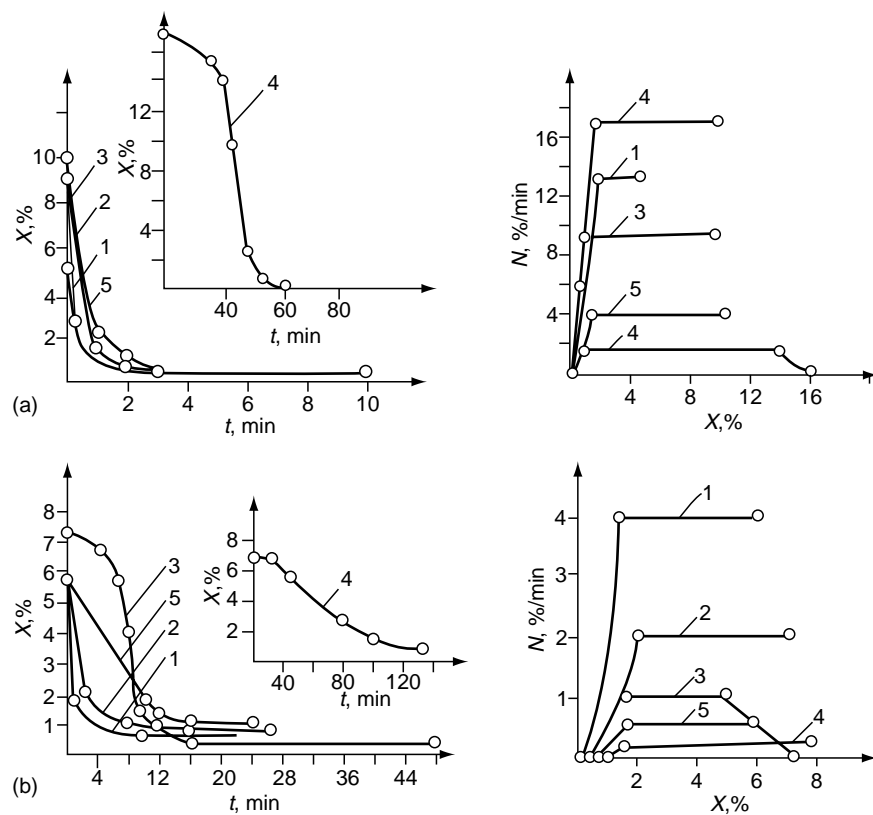


FIGURE 29.18 Drying kinetic data for (a) penicillin; (b) ascorbic acid (1, countercurrent spin dryer; 2, flash dryer with impacting streams of gas; 3, spouted bed dryer; 4, gas filtration in stationary bed; 5, fluid bed; X , moisture content; t , time; N , drying rate).

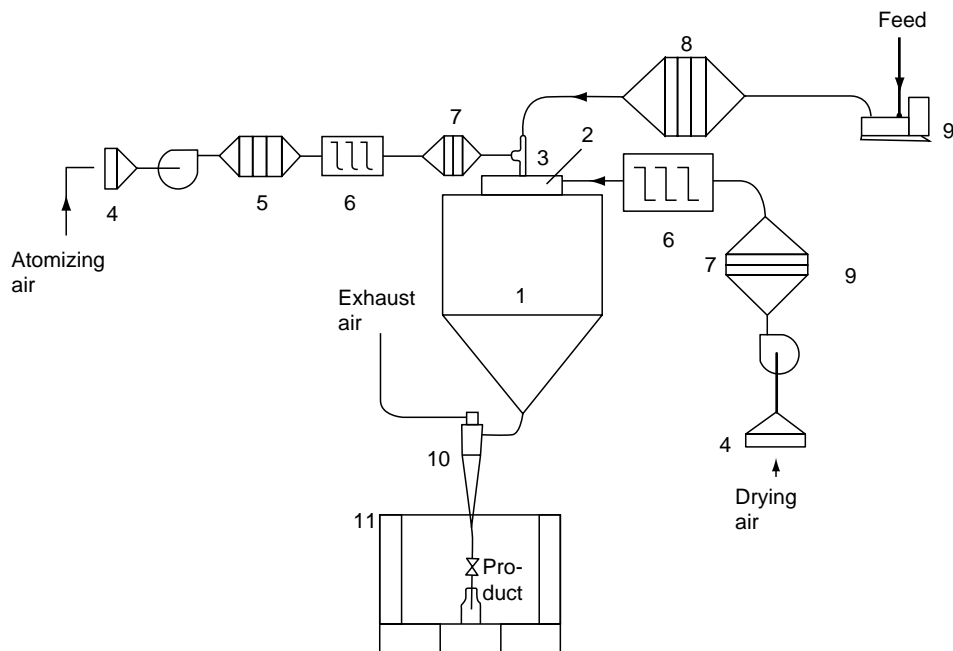


FIGURE 29.19 Aseptic open spray drying system (1, drying chamber; 2, air dispenser; 3, atomizer; 4, prefilter; 5, filter; 6, heater; 7, HEPA filter; 8, sterile feed filter; 9, feed pump; 10, cyclone collector; 11, packing room). (Courtesy of A/S Niro Atomizer, Soeborg, Denmark.)

The packaging areas should be provided with laminar flow units supplied with filtered or washed air. All personnel and ancillary equipment must be placed downstream from the sterile product. A schematic of the packaging area of a typical spray dryer for aseptic operation is shown in Figure 29.20. In general, for aseptic noncontaminating drying operation, rotating or sliding parts should be avoided in dryer installations. Dryers themselves and their ductwork can be easily sterilized by maintaining the installation at elevated temperatures for specified periods. The *U.S. Pharmacopoeia* specifies heating at 160–170°C for 2–4 h. However, most aseptic dryers are sterilized at temperatures in the range 200 to 250°C for several hours in practice.

Effectiveness of thermal sterilization may be tested by placing samples of test bacteria (*Bacillus subtilis* for hot air and spores of *Bacillus stearothermophilus* for steam sterilization) in strategic locations in the installation. After sterilization, samples are sown onto culture media in petri dishes and observed for growth. Absence of any growth indicates that sterilization was successful.

To maintain high-purity standards of the product, several precautions must also be taken in the design of the drying chamber and installation [8].

The installation must be leakproof and must operate under slight overpressure.

All inside surfaces must be as smooth as possible; all corners must be rounded to guard against material deposition.

All ductwork should be as short as possible to prevent product buildup.

All locations where mechanical friction of parts take place should be eliminated as they can produce small metallic contaminations in the product. Rotating or sliding parts should be avoided.

Aseptic dryers with an open drying cycle provide satisfactory protection against microbial infection. Nevertheless, closed-cycle dryers offer a much better possibility of maintaining proper sanitary conditions. Unfortunately, their higher costs make them justified only in some special cases.

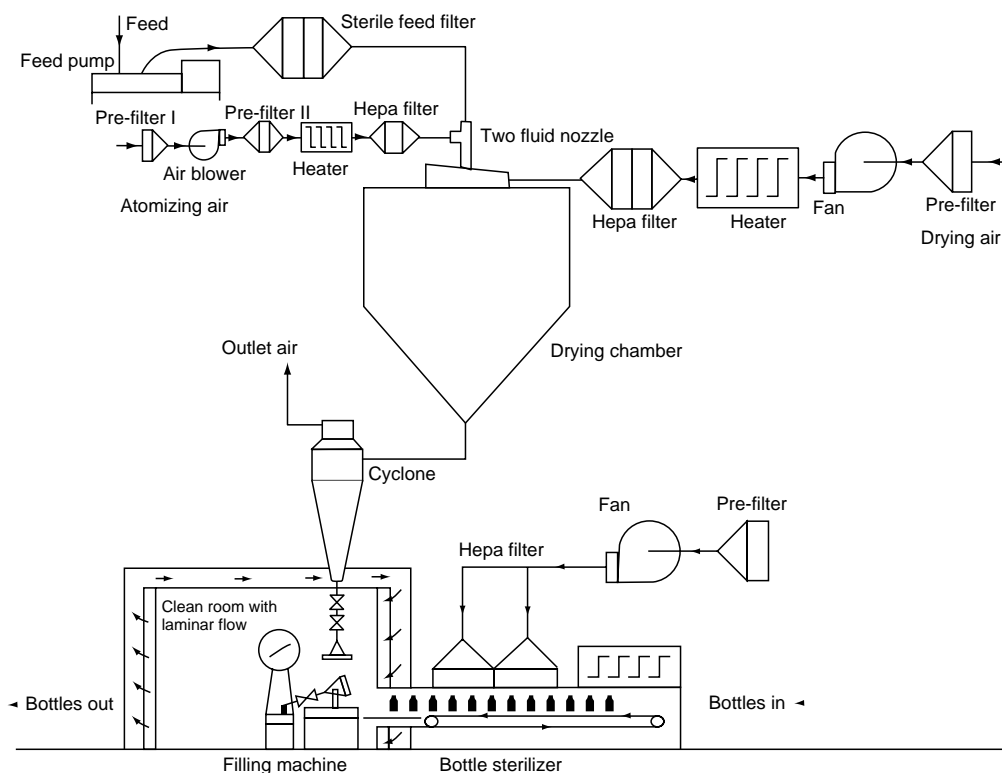


FIGURE 29.20 Aseptic spray dryer and packaging area. (Courtesy of A/S Niro Atomizer, Soeborg, Denmark.)

29.8 SOLVENT RECOVERY AND CLOSED-CYCLE DRYING

A large portion of all pharmaceuticals is obtained from nonwater solutions. The solvent is either a single component solvent (i.e., ethanol, methanol, acetone, etc.); however it is often a mixture of solvents. Table 29.7, containing some information selected from Pakowski [9], presents exemplary drugs with moisture that is a solvent mixture.

These solvents, due to their high cost and detrimental influence on the environment, may not be disposed off into atmosphere even at very low concentrations. In vacuum dryers solvent vapors after filtering in dust collectors undergo condensation in single or multistage condensers. In convective drying the presence of solvents and powders can produce an explosive mixture with air, which calls for inert gas as the heat carrier. It is then necessary to close the drying cycle.

In closed-cycle drying the spent gas, after separated from the dried material, is dehumidified and after reheating returned to the dryer.

A closed-cycle dryer is justified in the following cases:

Flammable, toxic, or valuable organic solvent is used. In this case the solvent will be fully recuperated.

Inert gas is used as drying medium, which is recommended if flammable solvent or solid is dealt with. Easy oxidization of the product also requires inert gas drying.

Solids or its solvent or vapors produced during drying are toxic, have an unpleasant odor, or can by other means pollute the atmosphere.

Inert gas, usually nitrogen, circulating in the cycle is continuously supplied with a fresh makeup gas. For a spray dryer with evaporation rate of acetone, 110 kg/h, approximately 3 m³/h makeup nitrogen is necessary during normal operation. Purging of the installation after washing and sterilization requires approximately 75 m³ nitrogen.

A closed-cycle spray dryer is shown in [Figure 29.21](#). Essentially the same elements of installation are used as for normal open-cycle dryers; however, disk or liquid-nozzle atomizers are preferred as they do not require gas for spraying. Solvent evaporates into the stream of inert gas, usually nitrogen, leaving the material dry. Most of the particles then formed are separated in the dryer; the rest are recovered in high-efficiency cyclones. Hot and humid effluent gas is contacted with cooled solvent in a scrubber where the solvent vapors partially condense. Dehumidified gas is reheated and returned to the dryer. Another solution for water-based solids is a self-inertizing

TABLE 29.7
Some Pharmaceuticals Containing Multicomponent Moisture

Solid	Moisture	Composition, wt%	Total Moisture Content, wt%	
			Initial	Final
Raw vitamin C	1,2-Dichloroethane	68	20	0.1
	Trichloroethylene	20		
	Acetone	10		
	Ethanol	2		
Commercial vitamin C	Methanol		10.6	0.5
	Water			
Vitamin B6	Ethanol	80	20	0.3
	Water	20		
Dibromobenzene	Acetic acid	11	60	0.3
	Water	89		
Pancreatin	Acetone		30	0.2
	Water			
	Trichloroethylene			
Steroids	Methanol		13	1.5
	Ethanol			
	Water			
Multivitamin	Ethanol	85	40	0.3
	Water	15		
Antibiotics	Isopropanol			
	Acetone			

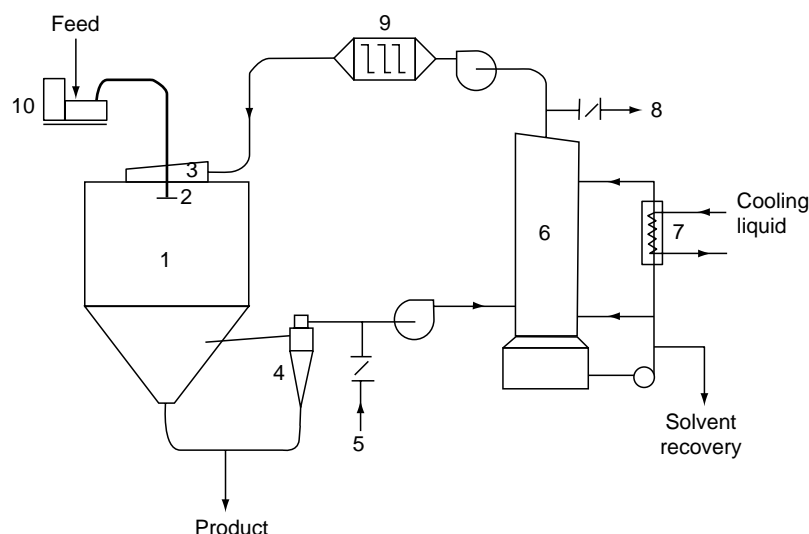


FIGURE 29.21 Closed-cycle spray dryer system (1, drying chamber; 2, atomizer; 3, gas dispenser; 4, cyclone collector; 5, inert gas supply; 6, scrubber-condenser; 7, heat exchanger; 8, vent for purging dryer; 9, gas heater (indirect); 10, feed pump). (Courtesy of A/S Niro Atomizer, Soeborg, Denmark.)

dryer (Figure 29.22) in which oxygen concentration in drying air is lowered to 4–5% in a directly fired gas heater.

The throughput of closed-cycle dryers is generally limited by the volume of the scrubber used for cooling and condensation. The dimensions of the scrubber increase with the amount of gas used for drying; however, closing the cycle of even such extensively gas-consuming dryers as pneumatic or fluid bed dryers was found justified.

Among the pharmaceuticals dried in closed-cycle spray dryers, antibiotics and antibiotic by-products

can be mentioned. Coating of powders with polymeric coating material dissolved in the feed or so-called microencapsulation is a new application of closed-cycle spray dryers. For dryers using very high gas flow rates, such as fluid bed dryers, the following principle may be used. The outgoing gas from the drying chamber is split into two parts. One part is heated and recirculated whereas the second part is sent to a condenser; after solvent removal it is mixed with the recycle stream and returned to the drying chamber. This allows operation at high gas flow rates in the drying chamber

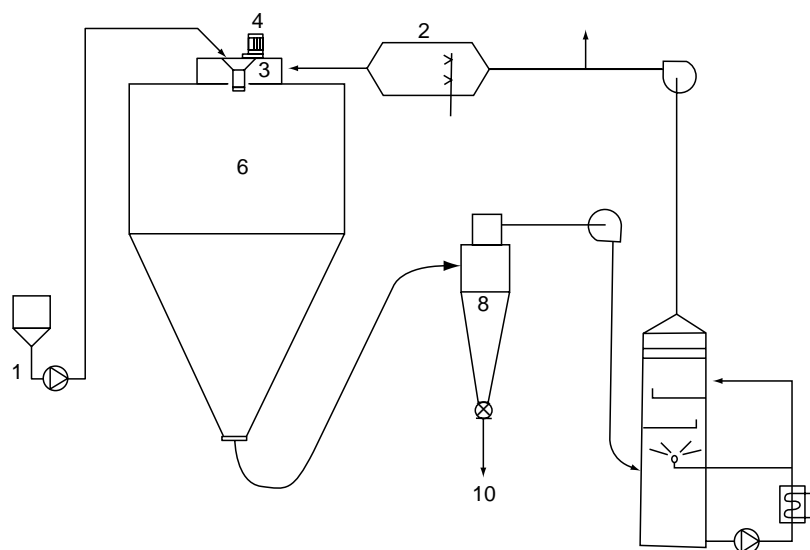


FIGURE 29.22 Self-inertizing closed-cycle dryer (1, feed tank; 2, direct heater; 3, atomizer; 4, atomizer drive; 5, scrubber; 6, drying chamber; 7, pumps; 8, cyclone; 9, fans; 10, cooler). (Courtesy of APV Pasilac Anhydro, Soeborg, Denmark.)

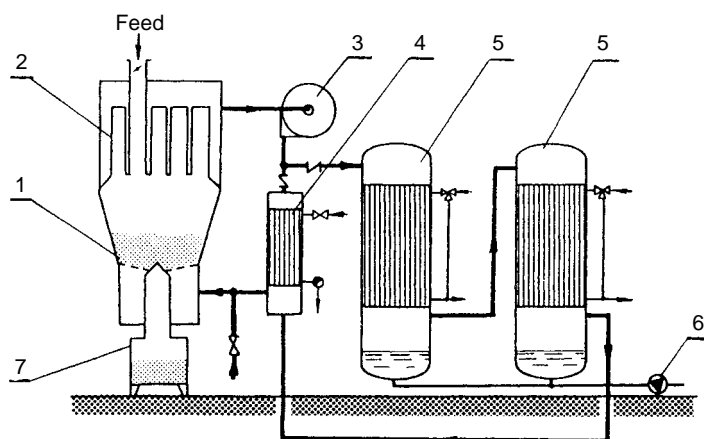


FIGURE 29.23 Closed-cycle batch fluid bed dryer (1, drying chamber; 2, bag filter; 3, blower; 4, heater; 5, condenser; 6, pump; 7, product collector).

and also relatively high solvent concentrations in the exit streams, which makes condensation more efficient. In the falling rate period, when the concentration of the solvent in the exit gas decreases, removal of moisture usually requires lowering the temperature of the cooling medium in order to obtain condensation at the required concentration of solvent. Figure 29.23 shows schematically a batch fluid bed dryer working in a closed-cycle operation [10]. Two condensers operating at different temperatures are used; this arrangement is especially appropriate when condensing a mixture of two different solvents.

Traces of solvents in gas vented to atmosphere from closed-cycle dryers are usually removed by incineration in gas or catalytic burners. This method may also be used for deodorization of exhaust gas. Chlorine-containing solvents cannot be burned and must be adsorbed on a suitable sorbent.

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30 Drying of Nanosize Products

Baohe Wang, Li Xin Huang, and Arun S. Mujumdar

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30.1 INTRODUCTION

Nanomaterials represent today's cutting edge in the development of novel advanced materials, which promise tailor-made functionality for unique applications in all important industrial sectors. Nanomaterials can be clusters of atoms, grains 100 nm in size, fibers that are less than 100 nm in diameter, films that are less than 100 nm in thickness, nanoholes, and composites that are a combination of these. In other words,

it implies that the microstructures (crystallites, crystal boundaries) are nanoscale [1]. Nanomaterials include atom clusters, nanoparticles, nanotubes, nanorods, nanowires, nanobelts, nanofilms, compact nanostructured bulk materials, and nanoporous materials [2]. Materials in nanosize range exhibit fundamentally new properties and functionalities such as surface effects, dimensionality effects, quanta effects, and quanta tunnel effects, etc.

Due to their unique optical, electronic, acoustic, magnetic, thermal characteristics and advantages as catalysts, etc., the applications of nanomaterials can be found in many fields and some are listed as below [3–6]:

- *Pharmaceuticals, healthcare, and life sciences*: New nanostructured drugs, gene and drug delivery systems; biocompatible replacements for body parts and fluids, self-diagnostics for use in the home, sensors for labs-on-a-chip, material for bone and tissue regeneration; blood replacement
- *Manufacturing*: Improvements for precision engineering; new processes, and tools to manipulate matter at the atomic level; nanopowders that are sintered into bulk materials with special properties that may include sensors to detect incipient failures and actuators to repair problems; chemical–mechanical polishing with nanoparticles, self-assembling of structures from molecules; bioinspired materials and biostructures
- *Automotive and aeronautics industries*: Nanoparticle-reinforced tires; external painting; non-flammable plastics; self-repairing coatings and textiles
- *Electronics and communications*: Media-recording devices using nanolayers and dots, wireless technology; dramatically more capable electronic circuits
- *Chemicals and materials*: Catalysts; superhard and tough-drill bits and cutting tools; “smart” magnetic fluids for vacuum seals and lubricants
- *Energy technologies*: New types of batteries, artificial photosynthesis for clean energy, quantum well solar cells, safe storage of hydrogen for use as a clean fuel, energy savings from using lighter materials and smaller circuits
- *Space exploration*: Lightweight space vehicles, economic energy generation and management, ultrasmall and capable robotic systems
- *Environment*: New membranes that can selectively filter contaminants; nanostructured traps for removing pollutants from industrial effluents, characterization of the effects of nanostructures in the environment
- *National security*: Detectors and detoxifiers of chemical and biological agents, hard nanostructured coatings and materials, camouflage materials, light and self-repairing textiles, miniaturized surveillance systems

Research done on nanotechnology is found throughout the literature. For example, when a drug is administered as nanoparticles, the absorption and

oral bioavailability of them, e.g., heparin [7], enalaprilat [8], tobramycin [9], and antitubercular drugs [10], were significantly enhanced in comparison to oral-free drugs. Chastellain et al. [11] synthesize the magnetic nanoparticles with specific shape and precise size with tailored surface chemistry and topography for biochemical purposes, i.e., the precise delivery of drugs using magnetic nanoparticles to the exact tissue by applying the external magnetic fields. Pandey et al. [12] investigated the process that the drug was encapsulated in nanoparticles of a synthetic polymer. The results emphasize the power of nanotechnology to make the concept of enhancement in oral bioavailability of azole antifungal drugs come to reality.

Since 1976, over 80,000 nanotechnology-related patents have been issued by the United States Patents and Trademarks Office (USPTO) [13]. However, among them, the United States leads the biomaterials such as tissue engineering and advanced controlled release as well as semiconductor technology. Europe is in a strong position in molecular sensors and diagnostics. In Asia, Japan leads in ceramics and magnetic materials. China and Korea also are amongst the most rapidly developing nations with strong R&D activities in nanotechnology in recent years [1].

The preparation of nanomaterials can be classified into three main approaches according to the states of the reactants used: liquid-phase, solid-phase, and gas-phase method [14,15]. One of the most popular methods in both laboratory and industry at present is the liquid-phase method. In the preparation of nanoparticles by the liquid-phase method, drying is an indispensable unit operation. Nanoparticles tend to agglomerate and properties of nanoparticles are adversely affected if we do not choose appropriate drying methods. So to assure that nanoparticles are well dispersed during drying is a vital requirement in the preparation of such products.

Rabani et al. [16] have presented an important finding that the drying process may mediate the self-assembly of nanoparticles since the systems may exhibit complex transitory structures [17] when the equilibrium fluctuations are mundane. They used a numerical model to show how the choices of solvent, nanoparticle size, and thermodynamic state affect the final morphologies. Since drying plays a role in nanomaterial processing, Pakowski [18] investigated the phenomenon of nanomaterials drying, such as drying stress and deformation of structure during drying, as well as the diffusion among the nanopores, etc. Wang et al. [6] presented and discussed more about the typical drying methods used in nanomaterials processing.

The preparation of gels includes mainly two steps: synthesizing ramiform- and continuous-structured gels in a liquid medium by the sol–gel process, and

then removing solvents from the gel pores [19]. Because gels are prone to deform or crack during drying, one of the difficulties in preparing aerogels is to preserve the microporous-structured gels. So rigorous-controlling conditions are demanded during drying and a minor mistake may result in failure of the whole preparation process.

30.2 RESEARCH IN DRYING OF NANOMATERIALS

Compared with drying of conventional materials, drying of nanoporous materials is really a complex task, because we must remove the solvents carefully without damaging the porous microstructure of gels. In the case of nanoparticles, agglomeration among particles must be prevented. The theoretical research on the drying of nanomaterials focuses mainly on the mechanism for cracking of porous-structured gels, agglomeration of nanoparticles, and solvent evaporation during drying.

30.2.1 SOLVENT EVAPORATION MECHANISM IN NANOMATERIAL DRYING

It is generally believed that solvent evaporation involves three periods, namely a constant rate drying, a first falling rate, and a second falling rate period [20].

In the constant rate-drying period, drying rate is independent of time and thickness of the gel. Additionally, the contraction (or shrinkage) rate of gel volume is equal to the solvent evaporation rate. Therefore, pores of gels are always filled with solvents in this period. In the initial period when contraction of gel volume takes place, the skeleton of the gel is very soft and capillary tension acting on the gel body is also very low. As the gel skeleton contracts and tends to become harder, its resistance ability against shrinkage is also strengthened while capillary tension increases simultaneously. In the constant rate-drying period, liquid flux from the pores interior to surface is equal to the liquid-evaporating rate. Liquid transfer occurs mainly by flow; diffusion is responsible for only a small fraction of the total flow.

Along with liquid solvent evaporation, the gel body shrinks and its skeleton strength also increases until it is strong enough to withstand compressive stress. When the contraction rate of gel is no longer equal to solvent evaporation rate, the solid surface is exposed and the first falling rate-drying period begins. At the critical point where the constant rate-drying period ends and the first falling rate-drying period begins, the radius of curved liquid is equal to the size of the gel pores, if contact angle is 0° , capillary tension reaches maximal value, and the gel skeleton

does not contract anymore. Below the critical point, the evaporation rate starts decreasing; liquid evaporation front recedes into the interior of the gel body, but primarily takes place over the solid surface. Endophragms of unsaturated pores are covered by a continuous thin liquid layer, which can provide a channel for transfer of liquid from pores. Therefore, in the first falling rate-drying period, liquid transfer occurs primarily by flow accompanied by vapor diffusion. In the first falling rate-drying period, a vapor-liquid interface appears in the interior of gel body, so cracking of the gels appears mainly during this period. Moreover, it is dependent on the thickness of the gel.

In the first falling rate-drying period, a thin liquid layer covers the unsaturated region of gel pores. Therefore, the outside surface of the gel does not change immediately so long as liquid flux is comparable to evaporation rate, and this state may be continuously preserved. However, as evaporation goes on, distance from the solid outside surface to drying front increases, the pressure gradient decreases and so does the flow rate. Hence, distribution of liquid on the outside surface becomes discontinuous, and drying enters the second falling rate period. In this period, evaporation completely occurs inside the gel body. The solvent evaporation rate is no longer sensitive to external conditions, the liquid near the external surface of the gel pores appears discontinuous, and transfer of liquid from the gel pores to the outside is taken by flow with diffusion as predominant. The force exerted on the gel is significantly mitigated during the second falling rate-drying period. Therefore, the gel body may dilate slightly. Because compressive stress on the nondrying side of skeleton is larger than that on the drying side, this may deform and collapse the gel skeleton. In this period, the gel volume no longer changes but its quality gradually decreases. Along with this excess part of solvent evaporation, various degrees of collapse may appear, micropores vanish, and the original skeletal structure shrinks into large agglomerations.

30.2.2 MECHANISM OF CRACKING OF GELS AND AGGLOMERATION OF NANOPARTICLES

30.2.2.1 Capillary Pressure Theory

Capillary pressure is the major cause leading to cracking of the gel skeletons or agglomeration of nanoparticles during drying [21]:

$$P = \frac{2\sigma \cos \theta}{r_p} \quad (30.1)$$

where P is the capillary pressure, σ is the liquid–vapor interfacial energy (or surface tension), θ is the contact angle, and r_p is the radius of curvature. As we note from Equation 30.1, in order to prevent nanoparticles from agglomerating or gels from collapsing, capillary pressure must be decreased to a minimum. The following means can decrease capillary pressure:

- Reducing surface tension (using supercritical drying, freeze-drying, and solvent-replacement drying)
- Changing wetting angle and making it to approach 90° (surface modification drying for nanomaterials)
- Enlarging pore size of gels properly (for capillary pressure is inversely proportional to the radius of pores) and making it uniformly (uneven pore size of gels leads to different capillary pressure and damage of the nanostructure)

Also we can use the microwave drying technique since it can shorten the drying period and heat materials evenly.

30.2.2.2 Models for Cracking of Gels

There are many divergent opinions about the causes of gel cracking and many models have been suggested based on experiments and observed results. The macroscopic model and the microscopic model are among the more representative models reported.

30.2.2.2.1 Macroscopic Model

The macroscopic model [6] assumes that drying stress is responsible for cracking of gels. Stress produced during drying is macroscopic and is exerted on the whole body of gels and not only on local drying regions. Its experimental basis is that gels either are integrated or crack into pieces, but never turn into powders. This model can illustrate quota relationship among cracking and evaporation rate, thickness of gels, and penetration coefficient. It can also explain some drying phenomena adequately. But it also has limitations, for example, the macroscopic model fails to explain why cracking mainly appears at the critical point, because there is not a sudden stress change that can cause gel cracking according to this model.

30.2.2.2.2 Microscopic Model

The microscopic model [6] assumes that the main cause of gel cracking is asymmetry of pores. According to this model, behind the critical point evaporation of liquid occurs in the larger pores initially, thus the stress on the larger pores can be relaxed. However, stress is still present in the smaller pores, which

can cause shrinkage of pores before cracking. Unlike the macroscopic model, the microscopic model assumes that the stress produced by local asymmetry of gel microstructure during drying is microscopic, and thus only affects this local region. So the distribution of pores plays a key role in this process; cracking is likely to occur with wider distribution of pores. This model can explain why the cracking takes place at the critical point. However, it fails to explain why cracking is reduced when the drying rate and the dimension of gel decrease. Clearly, this is also a deficiency of the microscopic model.

30.2.2.3 Mechanism of Agglomerations of Nanoparticles

At present, there are no generally accepted views on the formation of hard agglomerates of nanoparticles. There are several representative theories proposed, e.g., including crystal bridge theory, capillary pressure theory, hydrogen bond theory, chemical bond theory, etc. [21]. We shall only mention the essential concept of each theory.

Crystal bridge theory: During drying, capillary pressure enables particles to approach each other; crystal bridges form and tighten due to deposit of dissolved surface hydroxyls. With time passage these crystal bridges combine and large agglomerates can be formed.

Capillary pressure theory: When gels are heated, a solid surface is partially exposed due to evaporation of the absorbed water. Thus, surface tension resulted from the existence of capillary pressure in water may cause shrinkage of capillary walls. This is thought to be the main cause of agglomerates formation.

Hydrogen bond theory: Nanoparticles become attached as a result of hydrogen bond effects, and thus form agglomerates.

Chemical bond theory: Jonts and Norman believed that nonbridging hydroxyls existing on a nanoparticle surface are the basic sources of hard agglomerates. The reaction of nonbridging hydroxyls on the surface of neighboring particles taking place in this case is



Here Me–O–Me group accounts for the formation of agglomerates.

In fact, it is very difficult to illustrate mechanism of agglomerations by a single theory. Most authors attribute agglomeration to capillary pressure, but capillary pressure theory fails to explain why gels

exhibit large differences in agglomeration states when using different organic solvents with similar surface tension to replace water. Agglomerates can be easily dispersed in water if the particles are combined only by hydrogen bond effect, but actually it is very difficult, and therefore this type of agglomerate cannot be attributed solely to hydrogen bond theory.

30.3 DRYING METHODS FOR NANOMATERIALS

There are many advanced drying techniques for nanomaterials that have not been classified systematically so far. In this chapter, we try to classify these methods into direct drying, solvent-replacement drying, and modification drying for nanomaterials.

30.3.1 DIRECT DRYING

Direct drying method involves drying or calcinations of precipitates prepared by liquid-phase method after simple washing or filtering, but without solvent-replacement and surface modification. There is direct contact between the wet solid and the convective drying medium. The following are some of the more common methods in use.

30.3.1.1 Oven Drying

It is difficult to preserve slender porous gels when we are removing liquid solvents from gel pores. During oven drying, tremendous capillary pressure is produced in micropores when the solvents are removed, which may lead to serious agglomerations. Rapid drying can cause cracking of gels because gels and liquid solvents have different thermal expansion coefficients, so drying rate must be controlled as low as possible. But according to literature, low rate of safe drying will take as long as 1 year [14,21]. Oven drying is mainly used in the laboratory. Nanoparticles obtained by this method result in severe agglomerations, so quality of products is not very good.

30.3.1.2 Spray Drying

Spray drying is an established method that is initiated by atomizing and spraying suspensions into droplets followed by a drying process, resulting in solid nanoparticles. This method is widely used due to its many advantages, including the fact that it is a simple system, low cost, and easy to be industrialized. Several authors [22–27] have reported the fabrication of nanoparticles via spray drying.

Chow et al. [22] used spray drying to prepare nanosize hydroxyapatite (HA) particles. They used a nozzle to spray the acidic calcium phosphate solution and an electrostatic precipitator to collect nanosize powder. Their high-resolution transmission electron microscopy (TEM) showed that the particles, some of which were only 5 nm in size, exhibited well-ordered HA lattice fringes. The thermodynamic solubility of the nanosize HA is also presented better than that prepared by the conventional methods.

Li et al. [25] synthesized nanoscale LiCoO_2 powders via spray drying in which equivalent amounts of lithium acetate and cobalt were dissolved in deionized water and some polyethylene glycol (PEG) was added. The suspension was spray-dried to produce mixed precursor in a spray dryer with a two fluid nozzle. Atomizing pressure was controlled at 0.1 MPa, inlet air temperature was 300°C , and the outlet air temperature was 100°C . LiCoO_2 powder was synthesized by calcining the mixed precursor at 800°C for 4 h. The resulting particle had homogeneous particles size in the order of hundreds of nanometers.

There are also other methods such as spray pyrolysis and electrospray pyrolysis besides the above method [26]. To prepare nanoparticles by spray pyrolysis, a starting solution is prepared by dissolving, usually, the metal salt of the product in the solvent. The droplets atomized from a starting solution are introduced to furnace. Drying, evaporation of solvent, diffusion of solute, precipitation, reaction of precursor, and surrounding gas, pyrolysis may occur inside the furnace before the formation of product. It is similar to spray drying except the type of precursor. For this, colloidal particles are typically used as precursors. Some products prepared by spray pyrolysis are listed in Table 30.1.

TABLE 30.1
Nanoparticles Prepared by Spray Pyrolysis

Particles	Chemical	Mean Size (nm)
$\text{CuO}\cdot\text{Cr}_2\text{O}_3$	Nitrate	70
$\text{CoO}\cdot\text{Fe}_2\text{O}_3$	Chloride	70
$\text{MgO}\cdot\text{Fe}_2\text{O}_3$	Chloride	70
$\text{Cu}_2\text{Cr}_2\text{O}_4$	Nitrate	70
MgFe_2O_4	Chloride	50
MnFe_2O_4	Chloride	50
$(\text{Ni,Zn})\text{Fe}_2\text{O}_3$	Chloride	50
$\text{BaO}\cdot 6\text{Fe}_2\text{O}_3$	Chloride	75

Source: From Wang, B., Zhang, W., Zhang, W., Mujumdar, A.S., and Huang, L., *Drying Technol.*, 23(1–2), 7, 2005.

In spray pyrolysis, mean size of final nanoparticles can be determined from droplet size of solution sprayed [23,24]. But a typical atomizer such as twin fluid or ultrasonic nebulizer that is used to generate droplet in spray pyrolysis is only capable of generating droplet with mean size in the range of several microns. For example, a droplet with a diameter of 5 μm can produce a particle with a diameter of 100 nm, the initial concentration of solute must be 0.0008%. But in practice, low concentration may lead to low rate of particle formation and affect the purity of product. So, ultrafine nanoparticles can only be produced from ultrafine droplets. The electrospray technique has been examined as a method to generate ultrafine droplets; during this process, a meniscus of a spray solution at the end of a capillary tube becomes conical when charged to a high voltage (several kilovolt) with respect to a counterelectrode. The droplets are stably formed by the continuous breakup of a jet extending from this liquid cone, generally referred to as a “Taylor cone.” According to previous studies, mean size of droplets sprayed by this method can be controlled in a range of 1 nm to several microns.

Faezeh [28] used the technique of electrospray drying to prepare carbon molecular sieve (CMS) nanoparticles. The electrospray drying experiments were operated in the cone-jet mode by making use of a strong electric field (metallic nozzle connected to a high-voltage source), in which a pendular droplet deforms into a conical shape and then passes through a weaker electric field (shielding electrode) at the same polarity as the first one. The charged particles were neutralized with the aid of corona discharge. During the investigation of the effects of applied voltage, liquid flow rate, and polymer concentration, he found that the liquid flow rate has the most important effect in determining the particle size. The narrower particle size distribution with an average size of 200 nm was obtained under the experiment with flow rate of 0.05 mL/h, at the voltages of 13.6 and 7 kV (on the capillary and ring, respectively), with a 0.05 wt% polyetherimide solution. The better morphology was also found at this operation condition.

30.3.1.3 Freeze-Drying

This method involves atomizing of the solution into a freezing agent (for example, liquid N_2) where tiny droplets are turned into solid particles and after necessary filtering, these solid particles are moved to a vacuum freeze-drying chamber in which they are heated and ice sublimates. The dried material is sintered to produce oxidized nanoparticles. Obviously freeze-drying consists of two steps namely spray freezing

TABLE 30.2
Particles Prepared by Freeze-Drying

Particles	Reactants	Size (nm)
W	Ammoniacal brine	3.8–6
W-25%Re	Ammoniacal solution	30
Al_2O_3	Brine	70–220
MgO	Sulfate	100

Source: From Wang, B., Zhang, W., Zhang, W., Mujumdar, A.S., and Huang, L., *Drying Technol.*, 23(1–2), 7, 2005.

and vacuum drying, respectively. Table 30.2 lists some powder products prepared by freeze-drying.

As regards drying of gels, the situation is more complex [29,30]. Water undergoes some volume expansion when freezing, which can separate adjacent particles when water turns into ice. Nanoparticles are prevented from agglomerating due to the formation of the solid phase. On the other hand, high-energy vapor–liquid interfaces are replaced by low-energy vapor–solid interface, so agglomerations that are induced by surface tension during drying can be mitigated in theory.

30.3.1.4 Microwave Drying

Zhang et al. [31] prepared CeO_2 nanoparticles by sol-gel process using citric acid and cerium nitrate. The solution was heated during reaction and dried by microwave energy. It was found that the reaction period is shortened from original 1–3 d to 30–60 min and drying period also shortened to only several minutes from 2–3 h. Titanium dioxide, aluminum oxide, zirconium oxide, and silicon dioxide have been successfully synthesized by this method [32]. Compared with conventional direct drying, microwave drying has the following merits:

- Rapid heating rate that only takes 1/10–1/100 the time needed by conventional ways.
- There is no temperature gradient in the materials during microwave heating, so the materials can be heated up more evenly due to volumetric heat absorption.

30.3.1.5 Supercritical Drying

Supercritical drying [33,34] was initially developed by Kistler to obtain materials having large pore volume and specific surface area. A fluid is qualified as supercritical when its pressure and temperature exceed values, e.g., carbon dioxide has the low critical parameters ($T_c = 31^\circ\text{C}$, $P_c = 7.29\text{ MPa}$), but the critical pressure for water is as high as 21.77 MPa. Supercritical

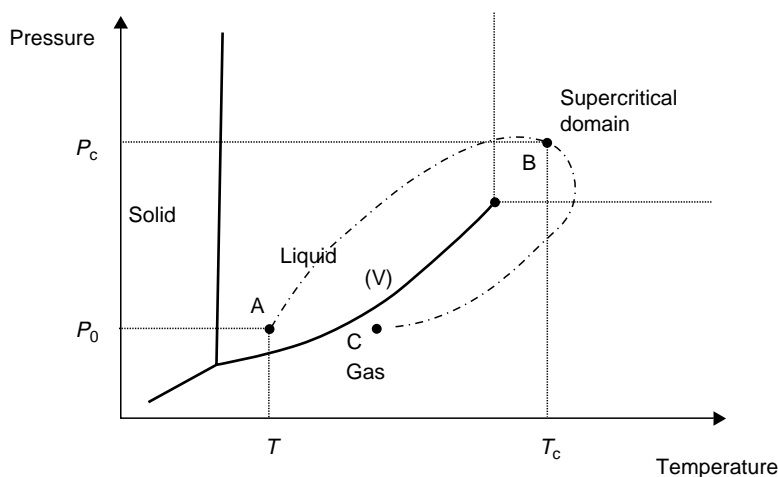


FIGURE 30.1 The supercritical drying procedure.

drying prevents capillary pressure between the vapor–liquid interface and the solid part of gel, which occurs during supercritical drying. During this process, the solvent within the gel is removed leaving only the linked gel network. This process can eliminate the solvent from the sol–gel without generating a two-phase system and the related capillary forces. The process is simply described in Figure 30.1. The sol–gel mixture at point A is initially heated and pressurized to the supercritical state (Point B). Then it is depressurized and cooled to room condition (Point C). However, the solvent vaporization curve (V) is not been crossed, i.e., no two-phase system appears. Only a low-pressure solvent vapor is present in the porous gel, which will be changed by air diffusion since the gel is highly porous with open pores. So gel skeleton is free of capillary tension during supercritical drying. Cracking of microporous structure and agglomeration of particles can be avoided theoretically.

The supercritical drying may be separated into four stages [18]:

- *Solvent-replacement stage:* Liquid CO_2 replaces the solvent to cover the gel.
- *Diffusional replacement:* Liquid CO_2 replaces the solvent to fill in the pores of gel.
- *Supercritical transition:* Liquid CO_2 and gel mixture are pressurized and heated to the supercritical conditions.
- *Isothermal expansion:* The mixture of liquid CO_2 and gel under supercritical condition is depressurized to the atmospheric condition.

30.3.1.6 Subcritical Drying

In contrast to supercritical drying, operating parameters for subcritical drying including temperature and

pressure are below the critical point. Wei et al. [35] used E-40 (multi-polysiloxane) as silicon source and isobutyl alcohol as solvent to prepare SiO_2 gel. The prepared gel was placed in autoclave, adding an appropriate amount of isobutyl and surfactant, preserved sometime under pressure of 2.3–2.6 MPa, temperature of 240–260°C, and then deflated slowly and cooled naturally; hydrophobic SiO_2 aerogel was finally obtained.

30.3.1.7 Direct Calcining

Direct calcining involves placing the precipitates or wet gels in a muffle furnace in which dehydrating and calcining simultaneously take place at high temperature. As shown in Table 30.3, Zhang et al. [36] also prepared TiO_2 nanoparticles, using a muffle furnace at 450°C for 3 h. Particle size of both TiO_2 and MgO increases with calcination temperature.

Direct calcining is economical and convenient, but seldom gives satisfactory nanoscale products because of serious incidence of agglomeration.

30.3.2 SOLVENT-REPLACEMENT DRYING

Solvent-replacement drying involves replacement of water in gels or precipitates with special organic solvents and removal of the organic solvents.

When we prepare nanoparticles by the liquid-phase method, wash with deionized water and filtration are necessary steps before drying in order to get rid of the residual ions. This usually causes severe agglomerations. Replacing water with selected solvents having low surface tension can yield products with minor agglomeration. Regarding gels, solvent replacement is required for subsequent treatment. Solvent replacement can be carried out by one of the following ways:

TABLE 30.3
Relation between Particle Size and Calcining Temperature

Temperature (°C)	400	430	450	500	600	700	800	900
MgO particles size (nm)	22	25	32	—	50	—	—	—
TiO ₂ particles size (nm)	—	—	22.1	23.9	34.4	41.1	46.9	49.6

Source: From Zhang, M., Yang, J., and Yang, X.J., Missiles Space Vehicles, 4, 51, 2001.

- *Washing with organic solvents*: Soaking, washing, and filtering with organic solvents several times are necessary steps before drying precipitates or gels. The functional groups of organic molecules replace nonbridging hydroxyls partially as well as produce sterically hindered effects to prevent agglomeration of particles. The commonly used organic solvents include methyl alcohol, ethyl alcohol, butyl alcohol, *tert*-butyl alcohol, acetone, cetane, silicone oil, etc. The need for washing with organic solvents is one of the reasons for the high preparation cost.
- *Azeotropic distillation*: Precipitates or wet gels are dissolved in solvents, which possess higher boiling points than water, but are immiscible in water. Under heating and vigorous stirring, water and organic solvents vaporize as an azeotropic mixture. Solvents that can be used for azeotropic distillation include butyl alcohol, isoamyl alcohol, isopropyl alcohol, propyl alcohol, glycol, ethyl alcohol, benzene, and toluene. Butyl alcohol is the most popular one. Researchers have concluded that surface hydroxyls on particles are replaced by butyl alcohol molecules, which produce steric hindrance effects to prevent agglomeration. But butyl alcohol can cause serious pollution problem and its recycling is quite problematic, therefore commercialization is very difficult.
- *Liquid carbon dioxide replacement*: Due to poor miscibility of liquid carbon dioxide with water,

water must be replaced with suitable organic solvents before supercritical CO₂ drying.

- *Supercritical carbon dioxide extraction*: Because gel pores are very fine, it takes a long time for liquid carbon dioxide to percolate into the interior of the micropores. The whole drying process can take much longer time. This problem is solved if supercritical carbon dioxide is used to extract organic solvents, which are used to replace water.

30.3.2.1 Solvent-Replacement Oven Drying

Dong et al. [37] aged TiO₂ hydrosol for 2 h in mother liquor, washed, and filtered with deionized water many times to obtain a hydrogel, and then replaced water with ethanol to get an alcogel. Alcogel was dried in oven at 90°C for 3 h, again kept in oven for 24 h at 110°C, finally calcined in a muffle furnace for 3 h at 550°C. Particle size of product was reduced and other properties also improved compared with direct oven drying (as shown in Table 30.4).

30.3.2.2 Solvent-Replacement Freeze-Drying

The medium of direct freeze-drying is water, but melting temperature of water differs from its boiling temperature by 100°C. The sublimation heat of water is high as a result of hydrogen bonding. Moreover, the cold trap temperature in freeze-dryer must be controlled under −50°C in order to prevent the saltwater

TABLE 30.4
Properties of TiO₂ Nanoparticle Obtained by Different Drying Methods

Samples	Drying Methods	Particle Size (nm)	Specific Surface Area (m ² /g)	Porosity (m ³ /g)
CP1	Direct oven drying + calcining	50–100	4.88	0.027
CP2	Ethanol-replacement supercritical drying + calcining	10–20	113.8	0.41
CP3	Ethanol-replacement oven drying + calcining	20–40	60.7	0.34

Source: From Dong, G.L., Gao, Y.B., and Chen, Sh.Y., *Acta Phys. Chim. Sin.*, 14(2), 142, 1998.

system from melting owing to its low plait point and extremely slow sublimation rate. Sublimation heat of *tert*-butyl alcohol is much lower than that of ice and has small difference between its melting temperature and boiling temperature. So cold trap temperature only needs to be controlled at -15°C to guarantee no melting to take place. *tert*-Butyl alcohol experiences smaller volume change than water when frozen, which is in favor of maintaining integrity of porous structures. Saturated vapor pressure of *tert*-butyl alcohol is higher than that of water (see Table 30.5), so drying time can be greatly reduced. Products dried by this method have excellent mesoporous structure.

Tamon et al. [30] synthesized resorcinol–formaldehyde (RF) hydrogel by sol–gel polycondensation of resorcinol with formaldehyde, and the gel had been immersed in ten times volume of *tert*-butyl alcohol for over 3 d to replace water in its pores. Then the gel was frozen at -30°C for 1 d, at -10°C for 1 d, and at 0°C for 1 d to obtain RF cryogel. Carbon cryogel was prepared by pyrolyzing RF cryogel at high temperature. Although pores undergo shrinkage during high-temperature carbonization, they maintain within mesoscale. Property of this type of carbon cryogel lied between carbon aerogel and carbon xerogel. Its specific surface area and mesovolume are smaller than carbon aerogel, but larger than carbon xerogel dried by direct evaporation.

30.3.2.3 Solvent-Replacement Supercritical Drying

30.3.2.3.1 Organic Solvent-Replacement Supercritical Drying

Because hydrogels are not suited for supercritical drying, hydrogels prepared from inorganic salt must be converted to alcogels by replacing water with alcoholic solvent before supercritical drying. Methyl alcohol was used firstly, alcogels are put in dryer, and some methyl alcohol is added; and it is essential to boost temperature and pressure at a proper rate

until methyl alcohol reaches its supercritical state and preserve for sometime in order to assure liquid methyl alcohol can turn into supercritical fluid completely. Release methyl alcohol gradually under constant temperature and reduced pressure. It was Kistler who prepared silicon aerogel by this type of supercritical drying for the first time in 1933. Al_2O_3 , TiO_2 , ZrO_2 , organic, and carbon aerogels were also prepared later by organic solvent-replacement supercritical drying [38]. Besides, nanoparticles of nickel hydroxide, nickel oxide, zinc oxide, zirconia, and $\alpha\text{-Fe}_2\text{O}_3$ were prepared by ethanol-replacement supercritical drying.

30.3.2.3.2 Liquid Carbon Dioxide-Replacement Supercritical Drying [21]

In general, alcohol (methyl alcohol) can cause esterification on gel surface. The surface of this type of gel is hydrophobic, therefore it is unlikely to absorb water in air. But alcohol has high critical temperature and pressure; in addition alcohol is flammable and toxic, especially methyl alcohol. Moreover, the gel skeleton may be compact during supercritical drying, in particular under the condition of the existence of base catalysis and water. Mechanism of this kind of compact is similar to aging. In 1985, Tewari et al. took CO_2 as a drying medium in supercritical drying which greatly reduced the supercritical temperature and enhanced reliability of drying equipment. Under operating conditions, CO_2 is chemically inert to gel skeleton. Due to the poor solubility of CO_2 in water, water must be replaced by organic solvents before supercritical drying. Aerogels prepared by this method have strong water affinity and tend to absorb water in air, so aerogels may gradually turn to cream color after sometime in air and even crack. Absorbed water can be removed by heating up to $100\text{--}250^{\circ}\text{C}$ without damage to aerogels' skeleton. Liquid carbon dioxide-replacement supercritical drying is currently the most common method to prepare aerogels and nanoparticles.

TABLE 30.5
Properties of *tert*-Butyl Alcohol and Water

Substances	Melting Point ($^{\circ}\text{C}$)	Boiling Point ($^{\circ}\text{C}$)	Density Change (g/cm^3)	SVP ^a /Pa
<i>tert</i> -Butyl alcohol	25.5	83	3.4×10^{-4} (26°C)	821
Water	0	100	7.5×10^{-2} ($^{\circ}\text{C}$)	61

^aSVP is the saturated vapor pressure.

Source: From Luan, W.L., Gao, L., and Guo, J.K., *NanoStruct. Mater.* 10(7), 1119, 1998; Tamon, H., Ishizaka, H., Yamamoto, T., and Suzuki, T., *Carbon* 37, 2049, 1999.

30.3.2.3.3 Supercritical Carbon Dioxide-Replacement Extraction Drying

If we substitute supercritical CO₂ for liquid CO₂ which is used in carbon dioxide-replacement supercritical drying, the drying operation is called supercritical carbon dioxide-replacement extraction drying. Novak et al. [33] prepared BaTiO₃ particle aerogel by this method. Compared with liquid carbon dioxide-replacement supercritical drying, drying period of supercritical carbon dioxide-replacement extraction drying is greatly shortened, so preparation cost is reduced accordingly.

30.3.2.4 Azeotropic Distillation Drying

Azeotropic distillation drying is used to remove water in precipitates or wet gels as completely as possible and then the residual solvent still needs further drying and calcining after removal of water. Organic solvents can be reused after condensation and lamination. Nanoparticles of zirconium oxide, aluminum oxide, nickel oxide, silica dioxide, indium oxide, aluminum and magnesium hydroxide (AMH) had already been prepared by this method. Hu et al. [39] removed water in TiO₂ gel using butyl alcohol azeotropic distillation followed by supercritical drying and obtained TiO₂ nanoparticle.

30.3.2.5 Solvent-Replacement Microwave Drying

Yamamoto et al. [40] replaced water in the already prepared organic hydrogel with *tert*-butyl alcohol and followed it by 10 min of microwave drying, after carbonizing at high temperature and obtaining a carbon xerogel. The result indicates that efficiency of solvent-replacement microwave drying is much higher than that of solvent-replacement oven drying and solvent-replacement freeze-drying.

30.3.3 MODIFIED DRYING FOR NANOMATERIALS

Although aerogels or nanoparticles obtained by supercritical drying have excellent quality, rigorous operating conditions, long preparation time, and expensive equipment cost are needed. To realize subcritical or ambient drying for gels and nanoparticles, a series of modification measures must be implemented, including strengthening hardness of gel skeleton as well as changing size and uniformity of pores, surface modifying, etc. Some drying control chemical additives (DCCA) can also be added to reduce cracking and shorten drying period [41,42].

The modification of nanomaterials is a pretreatment process that wet gels or precipitates are subjected to some modification measures in order to dry them at lower pressure or in ambient conditions.

30.3.3.1 Improving Uniformity of Gel Pores

It is impractical to expect skeletal structure of gel to be highly uniform, as the skeletal structure of gel is obtained by hydrolyzation and condensation reaction of pentamethide. According to Equation 30.1, the capillary pressure is inversely proportional to the radius of pores, and hence there are different stresses on gel pores which lead to cracks or breaks during drying. Moderate increase of pore size can decrease capillary pressure as well as increase penetration coefficient of the gels, which in turn reduces drying stress. On the other hand, adding formamide can restrain hydrolysis rate and speed up condensate rate of silicon alkoxide, hence larger gel skeleton can be obtained. This method also has a disadvantage, since large pores need higher sintering temperature [43].

30.3.3.2 Altering Volatilization Order of Solvent Mixture

If the solvent in gel pores is a mixture of water and methanol (or ethanol), appreciable amount of water may be left when drying is finished because methyl alcohol is easier to volatilize than water. But if some surfactants of low surface tension and volatility (for example, *N,N'*-dimethylformamide (DMF)) are added to the wet gel before drying, the possibility of gel cracking is reduced.

30.3.3.3 Surface Modification

Deshpande et al. [42] replaced water in hydrogel with ethanol (or acetone, cetane). The hydrogel was placed in trimethylchlorosilane (TMCS) (solvent is benzene, toluene, or cetane) for modification treatment, and then washed with ethanol. Surface wettability of gels was improved significantly and the contact angle nearly reached 90° (see Table 30.6). The capillary pressure was also reduced. Then it was dried at ambient pressure and room temperature for 24 h, and then at 50 and 100°C for 24 h each to produce a porous SiO₂ xerogel having properties of aerogels which are obtained via supercritical drying.

Shen et al. [43] prepared silica gel from polysiloxane E-40 precursor. After aging and solvent replacement, the silica gel was placed in TMCS solution (10%) for surface modification for 3 d, followed by washing with silicone oil repeatedly, and dried under ambient pressure to obtain SiO₂ xerogel. Its pore size was 10.4 nm, the specific surface area 969 m²/g, and shrinkage ratio 9%.

TABLE 30.6
Contact Angles of Modified and Unmodified Gels

Solvents	Surface-Modified (Atmospheric Drying)	Nonsurface-Modified (Supercritical Drying)
Ethanol	76.7°, 78.4°	30.3°, 35.1°
Acetone	79.3°, 77.2°	29.1°, 37.2°
Hexane	89.6°, 82.7°	41.3°, 48.4°
1:4 Dioxane	81.1°	66.4°

Source: From Deshpande, R., Smith, D.M., and Brinker, C.J., Preparation of High Porosity Xerogel by Chemical Surface Modification, U.S. Patent 5,565,142, 1996.

30.3.3.4 Reinforcement Strength of Gel Skeleton

In order to reinforce the mechanical strength of the gel skeleton, the following measures are usually taken [41]:

- *Adjusting hydrolysis condition:* Hydrothermal and chemical treatment are applied to adjust hydrolysis condition in the interest of accelerating condensation reacting to obtain gels that have high polymerization degree, and reinforcing gel skeleton strength accordingly.
- *Aging:* Aging is a process of dissolving and re-depositing of gel granules, by this means connective status of the gel skeleton can be improved. Einarsrud et al. had succeeded in reinforcing gel skeleton using water–ethanol, ethylate–ethanol to age gels.
- *Adding DCCA:* Formamide, dimethyl formamide, dimethyl acetamide, glycerol, and oxalic acid are generally used as DCCA reagents. DCCA can constrain hydrolysis of alkoxide and speed up condensation rate and all these are in favor of high strength of gel skeleton. On the other hand, it can make size distribution of gel pores more narrow and reduce drying stress. However, DCCA can also bring certain side effects, for example, some organic impurities are difficult to get rid off and produce bladders during sintering. Some of them even cause nigrescence of gels at high temperature due to its carbonization, etc.

30.4 COMPARISON AND SELECTION OF DRYING METHODS FOR NANOMATERIALS

30.4.1 COMPARISON OF DRYING METHODS FOR NANOMATERIALS

Different drying methods were selected by different researchers to produce various nanoparticles or nanostructure materials. Different drying methods have

strong effects on the properties of nanomaterials, including particle size, particle morphology, porous structure, specific surface area, etc. Here we compare the results obtained by some researchers who employed different drying methods to prepare nanomaterials.

Dong et al. [37] aged their TiO₂ hydrogel in the mother liquor for 2 h, filtered, and washed the hydrogel with deionized water several times. The hydrogel was separated into two parts: A and B. Part A was dried in oven at 90°C for 3 h, then at 110°C for 24 h and obtained as A1. B was converted to B1 by replacing water with ethanol. B1 was separated into C and D. Sample C was subjected to ethanol-replacement supercritical drying (260°C, 8 MPa, constant temperature and pressure for 0.5 h) to obtain C1. Sample D was dried in oven at 90°C for 3 h, and then at 110°C for 24 h to obtain D1. A, C1, and D1 were calcined in a muffle furnace for 3 h to produce products CP1, CP2, and CP3. Figure 30.2a shows the morphology of CP1, which displays serious agglomeration, with irregular particle shapes. The reason for this phenomenon is that during drying strong capillary forces develop that pull the particles into closer contact. Figure 30.2b shows relatively minor agglomeration; it can be attributed to replacement of water with ethanol, which has lower surface tension. Figure 30.2c shows the morphology of the ethanol-replacement supercritically dried CP3; it is observed that the cross-linked structure is well preserved during supercritical drying. Furthermore, supercritical drying can increase the specific surface area and porosity, as well as effectively prevents the formation of hard agglomerates. The particle and pore size distributions are more even. Table 30.4 shows the properties of three products.

Luan et al. [14] dried BaTiO₃ precipitate in three different ways: direct oven drying, azeotropic distillation drying, and *tert*-butyl alcohol-replacement freeze-drying. Their results showed that the solvent-replacement freeze-dried nanoparticle's size is approximately 30 nm, and

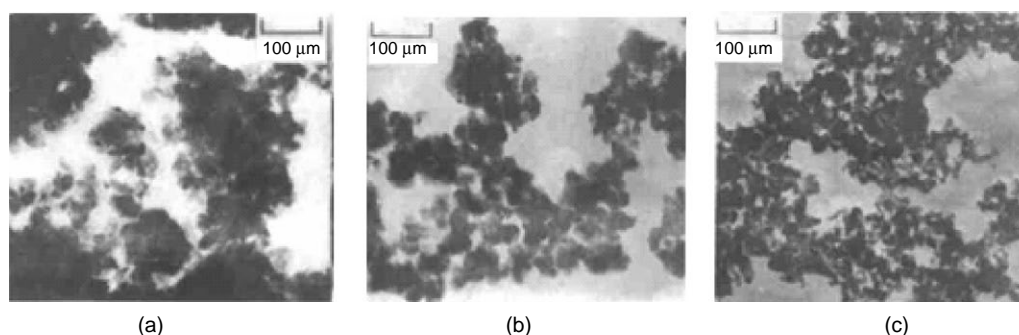


FIGURE 30.2 Nanoparticle morphologies prepared by different drying methods: (a) direct oven drying and calcining, (b) ethanol-replacement supercritical drying and calcining, and (c) ethanol-replacement oven drying and calcining. (From Dong, G.L., Gao, Y.B., and Chen, Sh.Y., *Acta Phys. Chim. Sin.*, 14(2), 142, 1998.)

its crystal form is perfect. Particle size of powder obtained by direct oven drying was about 80 nm and agglomeration of particles is quite obvious. Finally, particle size of powder obtained by solvent-replacement azeotropic distillation drying is between that of the above two; however, partial agglomeration and hollow spherical structure formation are found. Besides, sintering results indicated that there is a big difference in sintering activity of the three dried particles. Particles obtained by solvent-replacement freeze-drying exhibit higher density (larger than 85% TD) when sintered at 1250°C, which is several hundred centigrade lower than that of the conventional method. Compared with oven drying, sintering activity of particles obtained by azeotropic distillation drying is improved.

Tamon et al. [30,44–46] and Yamamoto et al. [47,48] prepared RF aerogels, RF cryogels, RF xerogels, and RF MW gels from RF hydrogels, respectively, by supercritical drying freeze-drying, oven drying, and microwave drying. Their porosity properties were determined by nitrogen adsorption. It was found that RF aerogels have the best mesoporous structure among the three gels, next in line are RF

cryogels, RF xerogels, and RF MW gels. Their scanning electron microscopic (SEM) images showed that the RF aerogels were composed of interconnected spherical particles; RF cryogels had cross-linked structure composed of interconnected primary nanoparticles; and RF xerogels and RF MW gels had cross-linked structures similar to the structures of RF cryogels. The primary nanoparticles of RF xerogels and RF MW gels are confirmed to link densely and the mesopores seem to be similar to those of RF cryogels, which facts support the results of nitrogen adsorption. From the above results, we can conclude that supercritical drying can preserve the skeleton of gel and minimize its shrinkage effect during drying; freeze-drying is also useful to prepare porous structure from hydrogels, and that it is difficult to obtain mesoporous RF xerogels by oven drying. Using microwave energy, one can dry hydrogels at higher rates than freeze-drying and oven drying while retaining the mesoporous structure of hydrogels.

Zhang et al. [49] prepared magnesium hydroxide precipitate and dried them in four different ways, as shown in Table 30.7. Then the dried precursors were calcined in a muffle furnace at 550°C for 4 h to

TABLE 30.7
Drying Conditions of the Experiments [49]

Sample	Parameters		
	Solvent Replacement	Drying Methods	Temperature (°C)
1	No solvent replacement	Oven	90
2	Alcohol	Oven	90
3	<i>N,N'</i> -Dimethylformamide	Oven	90
4	Butyl alcohol	Azeotropic distillation	97–120
5	Butyl alcohol + alcohol	Azeotropic distillation	97–120
6	Alcohol	Microwave drying	180
7	Alcohol	Supercritical drying	60

produce magnesium oxide nanoparticles. It is observed that MgO nanoparticles made from the precursor using direct oven drying get the largest particle size and they are also most irregular among the tested samples. It is due to severe agglomeration of the precursor during oven drying. On the other hand, it also indicated that sample 7 has the smallest particle size (18 nm) and perfect crystallinity; mild temperature (60°C) and free of capillary pressure during supercritical drying are responsible for this.

The morphologies for different samples using TEM are shown in Figure 30.3. The agglomeration and size distribution status can be clearly observed under the TEM images. Figure 30.3a is the image of sample 1, which displays serious agglomerations and irregular particle shapes. The reason for this phenomenon is that strong capillary pressure pulls the particles into closer contact during oven drying without solvent exchange. Comparing the images of sample 1 with samples 2 and 3 (Figure 30.3b and Figure 30.3c), it is seen that more uniform particle size distributions are obtained due to fewer agglomerations. The TEM images for samples 4 and 5 are shown in Figure 30.3d and Figure 30.3e. The finding is consistent with the fact that they have large particle sizes, which are calculated from Scherrer equation. It was also found that sample 4 presents cross-linked structure. Figure 30.3f is the photograph of sample 7 from which it is clearly observed that the particle size is the smallest compared with the others. It was also found that minor agglomeration occurs and spherical morphology of the particle is obtained.

Finally, microwave drying is used to dry the precursor. From the TEM image of sample 6, shown in Figure 30.3g, it is found that significant growth of the magnesium oxide particle size occurs due to the growth of the precursor during the high-temperature drying. However, its agglomeration status is bit similar to that in sample 7.

30.4.2 CHOICES OF DRYING METHODS FOR NANOMATERIALS

It is a difficult task to select the most suitable drying method for nanomaterials. There is no unified selection basis or set of criteria at present. Generally speaking, drying methods for nanomaterials should be determined according to laboratory test results and product quality parameters specified. The following observations are useful:

- Although direct drying has many advantages, such as simple operation, low preparation cost, and equipment investment, it is difficult to obtain

high-grade nanomaterials by this method due to excessive agglomeration.

- Solvent replacement is a necessary process in drying of nanomaterials, especially for nanoparticles prepared by the precipitation method. Organic solvent replacement (or washing) can improve quality of product significantly. Azeotropic distillation is one of the solvent-replacement methods with potential for development.
- Quality parameters of dried products (nanoparticles and aerogels) obtained by solvent-replacement supercritical drying (including organic solvent-replacement supercritical drying, liquid carbon dioxide-replacement supercritical drying, and supercritical carbon dioxide-replacement extraction drying) are the best. Next in line are solvent-replacement freeze-drying and solvent-replacement microwave drying.
- Among all solvent-replacement supercritical drying methods, organic solvent-replacement supercritical drying has limitations of poor safety due to its high temperature and pressure operation; in addition its equipment and operation costs are high. As for liquid carbon dioxide-replacement supercritical drying, operation period is very long. Supercritical carbon dioxide-replacement extraction drying has advantages of short operation time, low-operating pressure, and milder operating temperature.
- Microwave drying has advantages of even heating and a short-drying period and therefore it has great potential for development.

When the morphologies are considered, Rabani et al. [16] suggested four basic regimes of drying-mediated nanoparticle assembly. Disk-like or ribbon-like nanoparticle would domain reminiscent of spinodal decomposition form at the early time when the solvent evaporates homogeneously from the surface. However, when solvent disappears inhomogeneously because of the infrequent nucleation events, network structures are found at early time as vapor nuclei meet. These patterns are unstable if the domain boundaries are not frozen following the evaporation.

30.5 CONCLUSIONS

With ever increasing research and development of preparation technology for nanomaterials, drying technology for nanomaterials also needs to be enhanced accordingly. At present, research in this field is at a low level. Experimental study is mainly on the level of small batch operation and the whole drying cycle is quite long with operation cost remaining very high. Therefore, there is massive work to be done

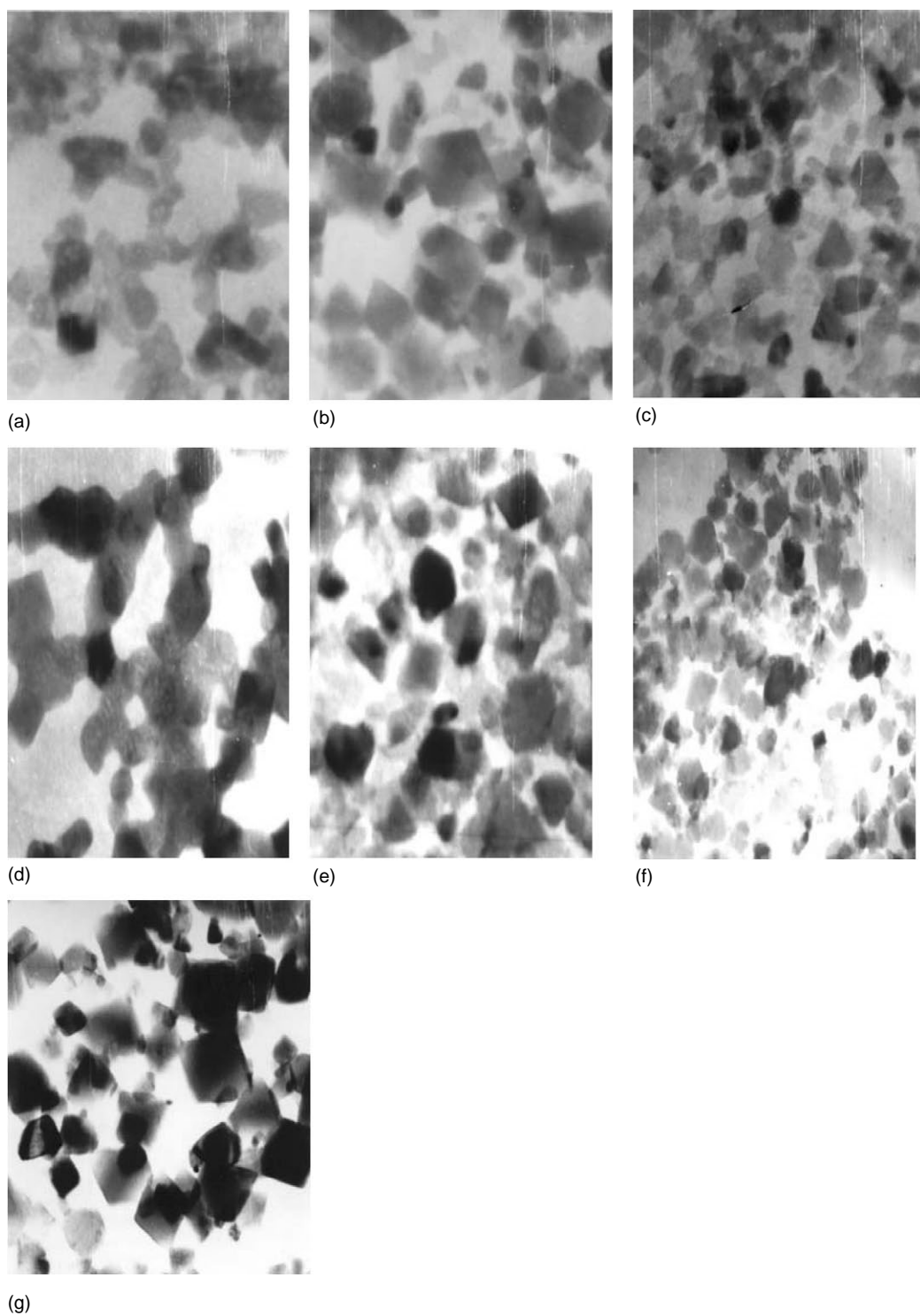


FIGURE 30.3 TEM photographs of the MgO samples: (a) sample 1, (b) sample 2, (c) sample 3, (d) sample 4, (e) sample 5, (f) sample 6, and (g) sample 7.

in order to industrialize drying technology for nanomaterials:

- Strengthening theoretical research on mechanism of drying for nanomaterials
- Exploiting subcritical drying or atmospheric pressure drying based on modification technology for nanomaterials due to their low operation cost and operation pressure, and mild operation temperature

- Making full use of the superiority of microwave drying and probing into further studies on mechanism of its drying
- Considering synthesizing, washing, filtering, modifying, and drying for nanomaterials as a whole in order to optimize the process
- Increasing transfer rate of drying process for nanomaterials
- Realizing continuous operation of drying for nanomaterials

Finally, the model of self-assembly of nanoparticles by Rabani et al. [16] is the first step in model of nanomaterials drying. We would like to emphasize that numerical models or simulations using molecular dynamics have still not matured. But this can help us to understand more about the phenomena during drying of nanosize materials. At present, the focus appears to be on empirical methods to develop nanoproducts by the wet process of desired properties. Agglomeration remains a major challenge. As new applications develop for new nanomaterials, new drying techniques may evolve. Also currently, all applications appear to require smaller quantities, so even expensive drying techniques may be utilized. In future, safety and environmental issues will become very critical since nanoparticles are difficult to collect if emitted in air. The fact that they can be easily inhaled or even ingested by the skin is especially worrisome and must be taken into account in the design and operation of any drying process.

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31 Drying of Ceramics

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31.1 INTRODUCTION

The demand for high-quality ceramics is increasing in many fields. The applications are spreading to highly functional materials known as the “fine ceramics” as well as materials for houseware and buildings. Although material science for ceramics has advanced significantly during the last few decades, the production processes still rely on old methods. Particularly, R&D has not addressed the drying process, and drying is usually carried out slowly in order to avoid undesirable cracks and deformations. The reason is often attributed to the fact that ceramic drying involves maintaining a molded configuration. Although ceramics are sensitive to the surrounding atmosphere during drying, it is very difficult to predict the heat- and mass-transfer rates and the air flow pattern in dryers with sufficient accuracy and generalize the results to the wide variety of ceramics. Many ceramic manufacturers are small companies, and they may not wish to invest much effort in ceramic drying R&D as drying occurs only at the pre-treatment stage in ceramic production. However,

drying is important for precisely designing the molding, determining the high-efficient heating rate in a dryer, and completing the drying so as to reduce the ratio of failure during sintering and yielding the qualified productions.

The objective of this chapter is to outline the R&D problems of the ceramic-drying process in order to promote its R&D. Industrial drying technology is almost kept as a confidential issue by each manufacturer and the available publications are quite limited. The research on the problem of drying shrinkage under these circumstances is reviewed here briefly as well. Future directions are also considered for improving the drying process for shortening drying time, energy saving, the precise design of molding, and high quality ceramic production.

31.2 ROLES OF DRYING IN PRODUCTION PROCESS

Ceramics are produced industrially by various processes. The typical processes for the production of

refractory are diagrammed in Figure 31.1. Drying is always involved in any process but sometimes takes place during the preparation processes such as grinding, classification, purification, mixing, and granulation. For such drying, the conventional technology for powders or granules may be available although there might be problems that are yet to be solved. The other drying phases after molding or pressing must be performed to remove water in order to maintain the molded configuration. Some of the major difficulties that arise during ceramic drying are as follows: (1) the rise in the drying rate must be naturally limited because the faster rate in the beginning of drying results in rapid attainment of the critical moisture content and then the remarkable reduction of the rate in the falling rate period; (2) failures and irregular deformation may be generated due to drying shrinkage and internal stress it induces; (3) molding must be designed with predicting the deformation; (4) uniform drying is difficult to perform because temperature and humidity, as well as local heat- and mass-transfer coefficients, are often highly dependent upon the location in the industrial dryers. The quality of the ceramic produced is seriously affected by these problems. The ceramic drying described in this

chapter is mainly confined to the drying of this step after molding.

Typical kinds of molding or pressing and their characteristics are summarized in Table 31.1. The uniaxial pressing is a common pressing method and is available for relatively flat ceramics like tiles. This pressing can be classified into two categories: dry and wet pressings. Slip casting is mold from slurry. In the injection molding, the cake of clay is injected into the mold through a nozzle from the cylinder. It is easy to supply a constant volume of cake into the mold that is appropriate for the mass production of complicated molding. The doctor-blade process is used for very thin-ceramic production, and a continuous operation is possible. The extrusion method involves extruding the cake through the dice screw. Pipes and honeycombs are produced by this method. We can see that some methods do not involve any drying process. However, the size and shape of ceramics are limited because molding with a uniform density distribution in the ceramics is difficult. Therefore, the wet methods are still required for the production of complicated shapes or large sizes. The drying process would thus become more important for the precise design of molds and the accomplishment of highly quality-controlled efficient drying.

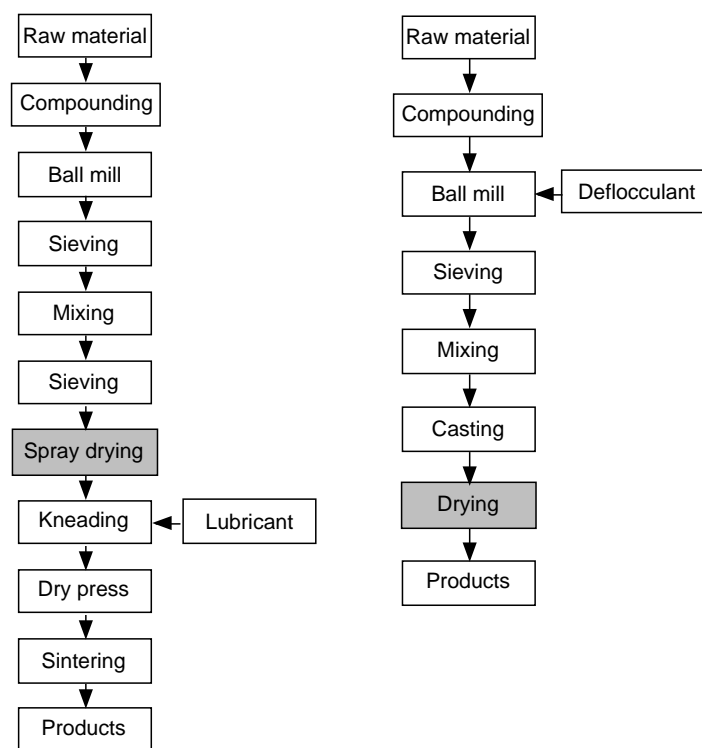


FIGURE 31.1 Typical flow of ceramic production.

TABLE 31.1
Classification of Molding Methods in Ceramic Production Process

Method			Product Shape	Merit	Demerit	Drying
Pressing	Uniaxial pressing	Dry press	Flat (container of electronic parts, brick for mill's wall)	<ul style="list-style-type: none">• The cheapest cost• Easy to automate factories• Direct sintering after pressing	Limitation of size and shape	No
		Wet press	Flat (container of electronic parts, brick for mill's wall)	<ul style="list-style-type: none">• The cheapest cost• Easy to automate factories	Limitation of size and shape	Yes
	Isostatic press (hydrostatic press or rubber press)		Cylinder, pillar, sphere (mill's ball, plug, vacuum bulb tube)	<ul style="list-style-type: none">• Uniform density distribution• Less distortion• Mass production	Frequent reformation of mold	No
	Hot press		Ceramics with high density and strength	<ul style="list-style-type: none">• Low porosity and high strength• Sintering at low temperature	<ul style="list-style-type: none">• Limitation of shape	No
Casting	Slip casting		Thin and complicated (melting pot, yarn guide of spinning or weaving, blade)	<ul style="list-style-type: none">• Simple equipments• Available for very complicated-shape products	<ul style="list-style-type: none">• Difficult in mass production• Easy to cause distortion• Taking long term for the production of casts• Wide space for storage of casts	Yes
Molding	Injection molding		Complicated (yarn guide of spinning or weaving, bulb, turbine blade)	<ul style="list-style-type: none">• Complicated shape with large size• High accuracy of size	<ul style="list-style-type: none">• Expensive metal molds• Not appropriate for small ceramics	Yes
	Doctor-blade process		Thin plate less than 1.5 mm thick (laminare package, multifold slab, thick slab)	<ul style="list-style-type: none">• Smooth surface• Good productivity• High accuracy of size	<ul style="list-style-type: none">• Difficult to remove waxes• Expensive equipment• Wide space for installation	Yes
	Extraction		Long tube and pillar (insulation tube, protection tube)	<ul style="list-style-type: none">• Uniform shrinkage• Unlimited length of products• Continuous production	Large equipment for the molding of large products	Yes

31.3 DRYING SHRINKAGE

Shrinkage during drying is one of the most important factors for the quality control of molded products. It is not only related to the deformation and the generation of crack, warp, and sagging but also influences the drying process itself. Hence the drying of ceramics cannot be discussed without taking into account the shrinkage problem.

Water in materials exists with a potential for occupying voids, i.e., a suction potential for granular beds like sand, powder, clay, and brick, an absorption potential for activated alumina and activated carbon, and a chemical bond for organic materials. The gradient of these potentials, corresponding to that of moisture content, becomes the driving force for the moisture movement in wet materials. The suction potential can be classified into: (1) capillary suction potential and (2) osmotic suction potential, and it is known that the shrinkage due to drying generally arises particularly for materials including water by the osmotic suction potential within their internal voids as clay. while the capillary suction potential is affected by the capillary force in the materials consisting of particles larger than 50- μm order, the osmotic suction potential is brought about owing to an electrochemical effect between the water and the surface of particles smaller than micrometers or submicron order. The surface of the particles at the interface between clay and water is generally negatively charged because positive ions are dissociated from the surface into the water and negative ions are exposed on the surface unless the other positive ions are also displaced so as to keep it electrically neutral. Then the hydrogen ions in the water are gathered in the vicinity of the surface as illustrated in Figure 31.2. Hence it is well

understood that water molecules would be subjected to osmotic force around the particles and the particles exist apart from each other forming a water film in wet clay. Norton and Johnson [1] predicted the water-film thickness of Florida kaolin, fully water based, on the data of pressure vs. linear shrinkage using the pressing chamber. They reported that the thickness varied between 3 and 30 nm for particles 0.32 μm in diameter depending upon the pressure (<4 MPa).

The osmotic suction potential for clay was measured against the moisture content by Newitt and Coleman [2] and Ohtani et al. [3] in comparison with the capillary suction potential for granular materials. Figure 31.3 [2] shows the relation between the osmotic suction potential and the moisture content for china clay. The potential was observed to be proportional to the exponential of the moisture content. The correlation was influenced significantly by the additives due to electrochemical interactions. The capillary suction potential for the bed of glass spheres of 40 μm in diameter was much lower than the osmotic suction potential for clay, and it was between 0 and 2500 mm H_2O depending upon the moisture content [3].

The shrinkage of clay is affected by the reduced thickness of the water films formed on the particles due to the osmotic suction potential and particles approaching closer to one another as water is removed during drying. If all the clay particles are separated from one another by water films and the void among particles of clay is sufficiently filled up with osmotic water, the volume of the shrinkage may be equal to the volume of the removed water until particles contact with the neighbor particles and the removed water begins to be replaced in the void by air [2]. When the drying proceeds beyond this stage, the

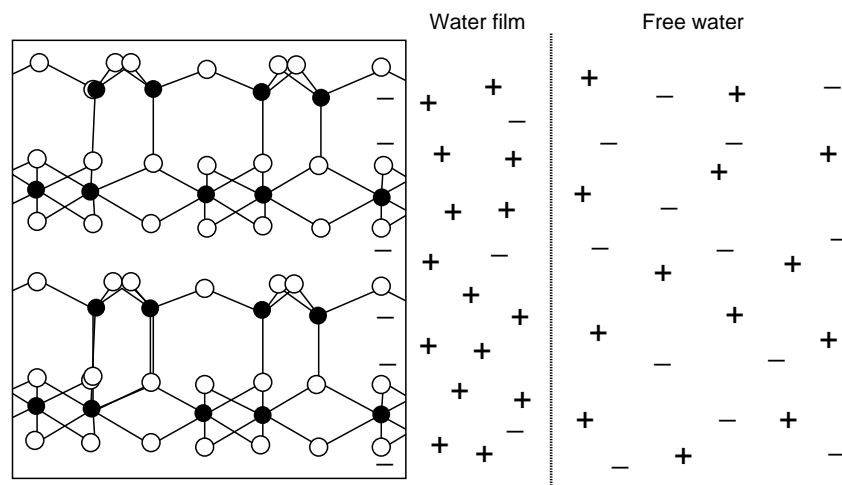


FIGURE 31.2 The electrochemical effect on a fine particle in clay–water system.

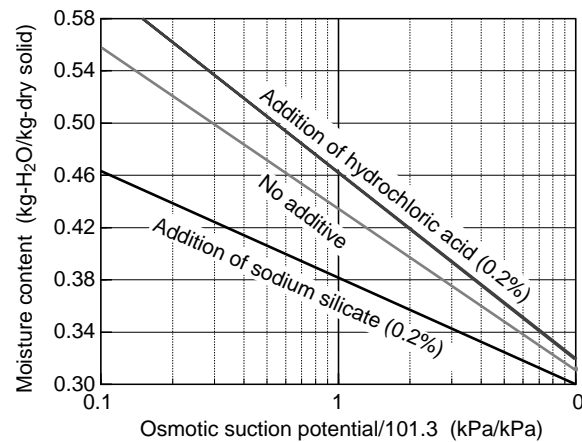


FIGURE 31.3 Relation between osmotic suction potential and moisture content for china clay.

shrinkage would be considered completed. The pattern of this shrinkage is shown in Figure 31.4. However, the actual behaviors of shrinkage of clay are not always idealized because it is not possible that all the particles are perfectly separated from one another at initial moisture content and begin to contact with neighboring particles at the same time.

The first quantitative studies on the behavior of clay shrinkage during drying were conducted by Norton [4], Macey [5], and Kamei and Toei [6]. Figure 31.5 [7] shows typical shrinkage curves. The feature of the curves was dependent upon the kind of the clay. Williamson [8] and Cox and Williamson [9] studied the effect of orientation of the particle's crystal in ball clays produced by casting on the drying shrinkage. The total shrinkage and the shrinkage rate in the perpendicular direction to the major face of particles were greater than those in the parallel direc-

tion. According to Noble et al. [10] if the molding pressure for consolidation before drying is increased, not only the minimum specific volumes of clay achieved by drying can be decreased but also the strength of the ceramics after sintering can be desirably improved.

Most of these shrinking behaviors have been determined under the conditions of normal atmospheric drying or during slow drying rate. Hasatani et al. [11] studied the impact of drying conditions on the shrinking behavior of clay molded into sphere or slab. The shrinking behaviors were influenced considerably by the air temperature and the drying rate since the moisture distributions and strain-stress due to the gradient of moisture in the clay might be related to the total volume. The shrunk volume was not always equal to the volume of water removed except during the early period of drying. These results

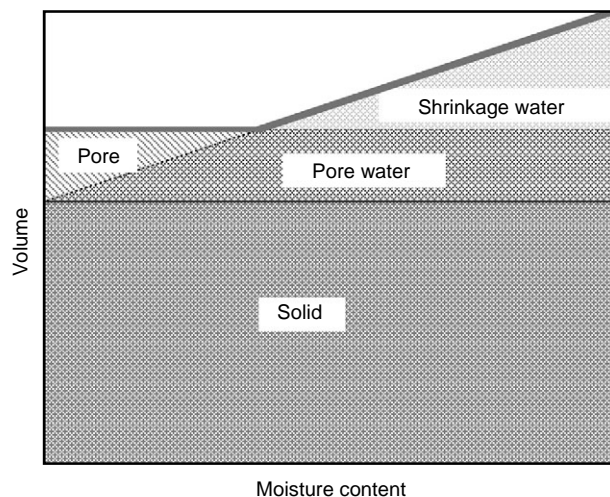


FIGURE 31.4 Shrinking pattern of clay.

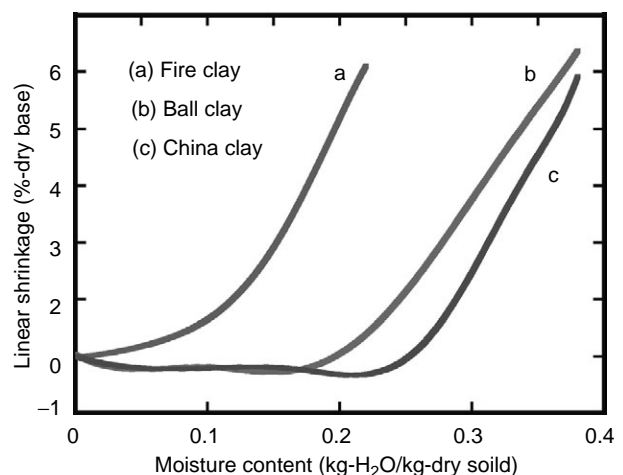


FIGURE 31.5 Typical shrinking curves for some sorts of clay.

suggest that it is not possible to precisely predict the shrinking behavior of molded clay only from the average moisture content and more accurate investigations are required to study its behavior.

31.4 MOISTURE MOVEMENT AND DRYING BEHAVIOR

Several researches have been carried out to understand the mechanism of moisture movement in clay during drying. Newitt et al. [12] and Wakabayashi [13] investigated the moisture movement in clay by liquid diffusion and vapor diffusion, which affect the drying characteristics particularly the falling rate. They concluded that the liquid diffusion dominates the movement until about 20%-dry basis in moisture content for stoneware clay and 30% for the mixture of 80% Kibushi clay and 20% feldspar. Wakabayashi [14] also evaluated the effective moisture diffusion coefficient of some sorts of clay such as Kibushi, Gairome, stoneware, feldspar, and their mixtures. The effective diffusion coefficient is available for the brief description of the moisture movement behavior. The effective diffusion coefficient D can be defined by

$$N = -\frac{D}{v_0} \frac{dw}{dx} \quad (31.1)$$

They correlated D as

$$D = \frac{\alpha K \gamma^2 \rho_p v_0 w^3 \exp(-Kw)}{k \mu S^2 (1 + \gamma w)} \quad (31.2)$$

where k is Carman's constant ($= 5$), N is the mass flux of moisture, S is the specific surface area of clay, v_0 is the specific volume of dry clay, w is the moisture content with dry base, γ is the specific gravity of a clay particle, ρ_p is the density of a clay particle, and μ is the viscosity of liquid water. Constants α and K are given empirically by 2.02×10^{11} Pa and 24.9 kg-solid/kg-water, respectively.

To understand the mechanism further, a microscopic investigation on the moisture content in clay is required. The osmotic suction potential was introduced as the driving force of moisture movement, described in Section 31.3, and was successively applied to the prediction of moisture movement in wet clay [15,16]. The theoretical analysis on the two-dimensional moisture transfer of cylindrical clay was performed taking into account the effects of both osmotic suction and strain-stress caused by the shrinkage [17,18]. However, only the transient mass-transfer equation was analyzed, assuming a constant drying rate on the external surface of the

clay. Comini and Lewis [19] developed the finite element method to solve simultaneous heat and moisture transfer for noncompressible porous media for a complicated geometry of the axial symmetry such as electric insulators. The three-dimensional problem on heat and moisture transfer, involving the drying shrinkage, was analyzed by Hasatani et al. [20]. Their model was not only limited to surface evaporation period but also applied to geometries more complicated than a simple slab shape. In the future, it is expected that drying kinetics would be experimentally studied and the simple analysis be made available for the entire drying periods.

Drying is a macroscopical phenomenon involving simultaneous heat and mass transfer. Suppose that heat conduction and moisture diffusion are dominant during the overall transfer process in a homogeneous medium. The multidimensional conservation equations for an anisotropic medium can then be expressed as

$$c_p \rho_m \frac{\partial T}{\partial t} = -\nabla \mathbf{J}_h + \dot{q} + \Delta H_{v\epsilon L} \frac{\partial C}{\partial t} \quad (31.3)$$

$$\frac{\partial C}{\partial t} = -\nabla \mathbf{J}_m \quad (31.4)$$

The heat flux \mathbf{J}_h and mass flux \mathbf{J}_m in Equation 31.3 and Equation 31.4 are given by

$$\mathbf{J}_h = -k_t \nabla T - k_c \nabla C - k_p \nabla p \quad (31.5)$$

$$\mathbf{J}_m = -D_w \nabla C - D_t \nabla T - D_p \nabla p \quad (31.6)$$

The motion of water is influenced by the water pressure, which is induced by the capillary or osmotic suction resulting from evaporation in porous media. The water flows in the relatively larger pores saturated with water rather than by diffusion process. In such a case, the pressure diffusion term dominates the mass-transfer rate in Equation 31.6. The Soret effect described by the temperature diffusion term is generally considered small compared with other terms. Then Equation 31.6 can be simplified as

$$\mathbf{J}_m = -D_p \nabla p \quad (31.7)$$

The liquid flow rate in the porous media is given by Darcy's law:

$$\mathbf{J}_w = -\frac{k_s}{\mu} \nabla p \quad (31.8)$$

where k_s denotes the permeability. As the mass-transfer rate should equal the water-flow rate if the vapor flow is ignored, the pressure diffusivity D_p is derived from both Equation 31.8 and Equation 31.7:

$$D_p = \frac{k_s}{\mu} \quad (31.9)$$

Hence Equation 31.4 can be rewritten from Equation 31.7 and Equation 31.9, assuming uniform properties in the body,

$$\frac{\partial C}{\partial t} = \frac{k_s}{\mu} \nabla^2 p \quad (31.10)$$

The capillary pressure of water is determined by a balance of the interfacial energies among the three phases: solid, liquid, and vapor, the wetting angle of liquid–solid, and the radii of the pores, which are dependent on the pore structure in the medium and the amount of water existing in the pores. The osmotic pressure is dependent on the pore structure and the liquid–solid interface. Therefore, one must predict the pressure p statistically from the pore structure distribution model.

In order to simplify the model, alternatively, the mass-transfer equation is often expressed by introducing the moisture content w as the driving force for moisture transfer,

$$\frac{\partial w}{\partial t} = \nabla \cdot \left(\frac{D}{\nu_0} \nabla w \right) \quad (31.11)$$

The initial and boundary conditions must be specified depending on the drying system and the surrounding atmosphere to which a medium is exposed.

31.5 DESIGNING PROCEDURE OF DRYERS

Dryers are classified briefly into three types: parallel current, countercurrent, and crosscurrent, by the direction of hot airstream and materials movement. The parallel-current type is introduced when fast drying rate is necessary in the earlier drying period. The countercurrent type is available for the case where drying operation is conducted to maintain a slow rate in the earlier period and sequentially increasing the drying rate at a later period. Batch dryers in a broader sense can be considered essentially as one of the crosscurrent type. In this section, the outline of the dryer-designing procedure is described for the most popular hot air heating.

31.5.1 HEAT AND MASS BALANCE

The first step toward any dryer design is the consideration of heat and mass balance within the dryer. Consider the heat and mass balance when wet materials with the moisture content w_0 (kg-water/kg-dry solid) and temperature T_0 (K) are dried in a dryer

and the dry products with the moisture content w_1 (kg-water/kg-dry solid) and temperature T_1 (K) are obtained. Suppose that the heat loss is negligibly small, the total heat balance equation is expressed by

$$\begin{aligned} G_0 i_0 + F_{d0}(c_s + c_w w_0)T_0 \\ = G_0 i_1 + F_{d0}(c_s + c_w w_1)T_1 \end{aligned} \quad (31.12)$$

where G_0 (kg-dry air/s) is the hot air mass-flow rate with dry base, i (J/kg-dry air) is the wet air enthalpy, F_{d0} (kg-dry solid/s) is the treatment rate of dry materials, c_s (J/(kg·K)) is the specific heat of dry solid, and c_w (J/(kg·K)) is the specific heat of water. The subscripts 0 and 1 denote the conditions at inlet and outlet of dryer, respectively. F_{d0} is given by $F_{d0} = F_{w0}/(1 + w_0)$ from the treatment rate of wet materials F_{w0} (kg-wet solid/s). The wet air enthalpy is correlated by

$$i = c_g T_g + (\lambda + c_v T_g)H \quad (31.13)$$

where c_g (J/(kg·K)) is the specific heat of dry air, T_g (K) is the air temperature, λ (J/kg) is the latent heat of water evaporation, c_v (J/(kg·K)) is the specific heat of steam, and H (kg-water/kg-dry air) is the humidity.

The mass balance deduces the following equation:

$$G_0(H_1 - H_0) = F_{d0}(w_0 - w_1) \quad (31.14)$$

If air temperatures at the inlet and outlet of the dryer and humidity at the inlet are known as operating conditions, G_0 and H_1 can be solved simultaneously from Equation 31.12 and Equation 31.14. The air temperature at the outlet T_{g1} should be chosen such that the humidity obtained at the outlet, H_1 , does not exceed the saturated humidity at T_{g1} . The flow rate of hot air necessary at the normal standard state (101.3 kPa, 298.15 K), V_g (m³_N-wet air/s), is obtained by

$$V_g = \frac{298.15 G_0 R}{101.3 \times 10^3} \left(\frac{1}{m_{\text{air}}} + \frac{w_0}{m_{\text{water}}} \right) \quad (31.15)$$

where R is the gas constant (8.314×10^3 J/(kmol·K)), m_{air} is the average molecular weight of air (28.97 kg/kmol), and m_{water} is the molecular weight of water (18.02 kg/kmol).

31.5.2 DETERMINATION OF DRYER SIZE AND DRYING TIME

A typical dryer used for continuous drying of ceramics is a tunnel dryer seen in Figure 31.6. The length of tunnel-type dryers and drying time are estimated fundamentally by the following procedures.

Now consider the countercurrent dryer modeled as shown in Figure 31.7. In the dryer, there are two

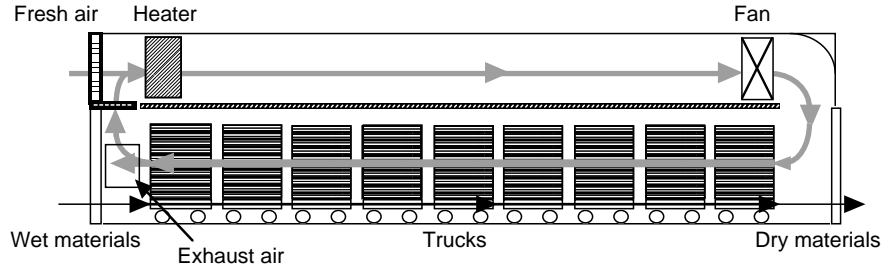


FIGURE 31.6 Configuration of typical tunnel dryers.

zones of the constant drying rate period and falling rate period. In the zone of the constant rate period, the mass balance between z , where z is equal to 0, and z_c becomes

$$F_{d0}(w_0 - w_c) = G_d(H_1 - H_c) \quad (31.16)$$

The mass balance between z , where z is equal to 0, and z is

$$F_{d0}(w_0 - w) = G_d(H_1 - H) \quad (31.17)$$

The differential expression of Equation 31.17 is written by

$$-F_{d0} dw = -G_d dH = R_c a dz \quad (31.18)$$

where w_c and H_c denote the critical moisture content and humidity at $z = z_c$, R_c is the constant drying rate, and a is the surface area of materials existing per unit length of dryer. The constant drying rate is expressed as

$$R_c = k_H(H_s - H) = \frac{h}{\lambda}(T_g - T_s) \quad (31.19)$$

where k_H and h are the mass- and heat-transfer coefficients, respectively; H_s is the saturated humidity; and T_s is the saturated temperature. Substituting Equation 31.19 into Equation 31.18, integration of Equation 31.18 deduces the length of the constant rate period zone.

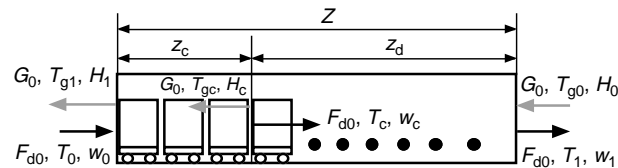


FIGURE 31.7 Model of hot air countercurrent flow for design of continuous tunnel dryer.

$$z_c = \frac{G_0}{k_H a} \int_{H_c}^{H_1} \frac{dH}{H_s - H} \quad (31.20)$$

When H_s is assumed to be constant in the constant rate period zone, Equation 31.20 becomes

$$z_c = \frac{G_0}{k_H a} \ln \frac{H_s - H_c}{H_s - H_1} \quad (31.21)$$

The mass-transfer coefficient k_H can be approximated analogically with the Lewis relation from the heat-transfer coefficient h and specific heat of wet air c_{gw} for the water–air system.

$$k_H \approx \frac{h}{c_{gw}} \quad (31.22)$$

In the zone of the falling drying rate, similar to the constant rate period zone, the following balance equations are obtained:

$$F_{d0}(w_c - w_1) = G_d(H_c - H_1) \quad (31.23)$$

$$F_{d0}(w_c - w) = G_d(H_c - H) \quad (31.24)$$

$$-F_{d0} dw = -G_d dH = R_d a dz \quad (31.25)$$

The reducing drying rate R_d is often correlated from the drying characteristic curves determined experimentally for wet materials in the following form:

$$R_d = R_c f\left(\frac{w - w_e}{w_c - w_e}\right) \quad (31.26)$$

where w_e is the equilibrium moisture content. Then the length of the reducing rate period zone is given by

$$z_d = \frac{G_0}{k_H a} \int_{H_1}^{H_c} \frac{1}{f\left(\frac{w_c - w_e}{w - w_e}\right)} \frac{dH}{H_s - H} \quad (31.27)$$

If the function f is approximated by the linear of the first order, the length z_d for a constant saturated humidity is derived as

$$z_d = \frac{G_0}{k_H a} \frac{w_c - w_e}{w_c - w_e + \frac{G_0}{F_{d0}}(H_s - H_c)} \times \ln \left(\frac{w_c - w_e}{w_1 - w_e} \frac{H_s - H_1}{H_s - H_c} \right) \quad (31.28)$$

As the preheating period is considerably shorter than other periods, the total length of the dryer Z is determined by

$$Z = z_c + z_d \quad (31.29)$$

When the surface area and the dry weight of each molded material are a_m and m_d , the drying time t_d given by the following equation is required to get dry materials with moisture content w_1 .

$$t_d = \frac{m_d a Z}{a_m F_{d0}} \quad (31.30)$$

31.6 DRYERS

31.6.1 TUNNEL DRYERS

The variety of dryers used for drying processes of molded ceramics is limited. A typical continuous type of ceramic drying is the tunnel dryer whose representative configuration is shown in Figure 31.6. Tunnel dryers are adaptable for drying requiring longer periods and are mass-producing, but they require a large installation space. Ceramics put on trucks are moved through the tunnel. Multiway flow of hot air is often introduced to control the drying rate independently during the constant and falling rate periods other than the countercurrent, parallel current, or crosscurrent against the moving direction of ceramics. In the example shown in Figure 31.8, the constant rate period is operated by a parallel flow to heat up rapidly and enhance the drying rate, and the counter flow is employed during the falling

rate period. Humid air during flowing in wet materials recirculates partly to control the humidity by mixing with fresh air. Hot air temperature supplied to the dryer is generally 370 to 470 K. Then the logarithmic mean of temperature difference between air and ceramics in the constant rate period becomes 30 to 70 K. The product of the heat-transfer coefficient and the specific surface area, ha , is between 230 and 350 W/(K·m³-dryer volume). In the truck carrier type, a number of ceramics are placed on a truck and the hot air flow rate is set usually to pass through among the ceramics at 1 to 5 m/s. If the clearance among the ceramics is not sufficient, it results in nonuniform drying and drifting the air flow between trucks and the wall of tunnel. Hence, the arrangement of ceramics on trucks and ventilation of air by blowers should take place carefully.

31.6.2 BAND DRYERS

Band dryers are also of continuous operation type. Materials are conveyed on bands rotating in the drying chamber as shown in Figure 31.9. Band consists of a net or a perforated plate. Hot air flows crossing the band upward or downward because heat- and mass-transfer rates are greater than that from parallel flow. It is possible to control the temperature, humidity, and flow rate of air at each optimum state in some zones separated in the dryer. Although the operation temperatures are almost the same as the tunnel dryers, ha is 45 to 95 W/(K·m³-dryer volume) for the parallel flow.

31.6.3 BATCH DRYERS

A typical configuration of batch dryers is shown in Figure 31.10. Wet materials are placed on trays or shelves, which are set in a dryer's room. Hot air circulates among the shelves in the room by a blower. These dryers are available for drying of various types

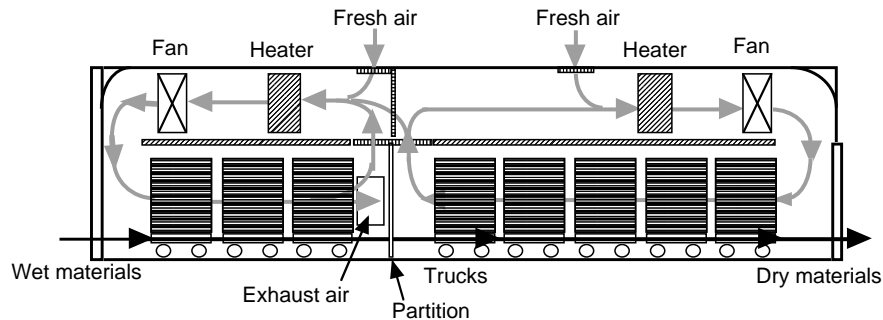


FIGURE 31.8 Concept of the combined-flow pattern of hot air in continuous tunnel dryer.

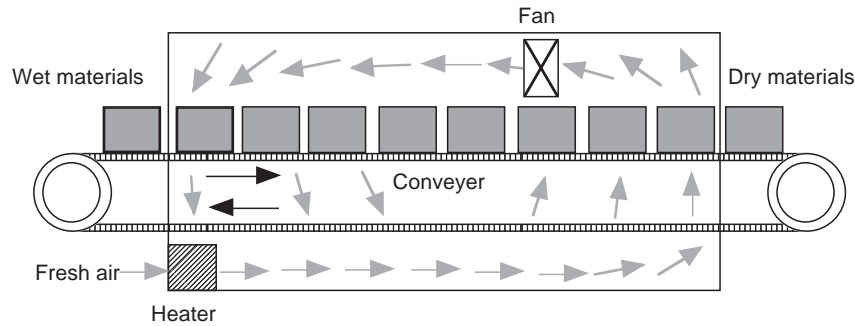


FIGURE 31.9 Configuration of typical band dryers.

of ceramics as drying conditions, i.e., time, temperature, air rate, etc., in each batch operation, can easily be changed and controlled. In most cases, however, manpower is required to pack, install, and remove the materials from the shelves. The drying time must be determined preliminarily from the drying characteristics obtained from a laboratory scale of experiment. As the drying characteristic and the critical moisture content are dependent upon temperature, humidity, and air rate, the preliminary experimental test should be performed with care to adjust to the atmospheric conditions in the dryers. When a part of the hot air recirculates to improve the effective energy utilization and control the humidity in the dryers, the recirculation ratio X is determined to maintain a constant humidity by the following equation:

$$X = 1 - \frac{R/G_1}{H_2 - H_0} \quad (31.31)$$

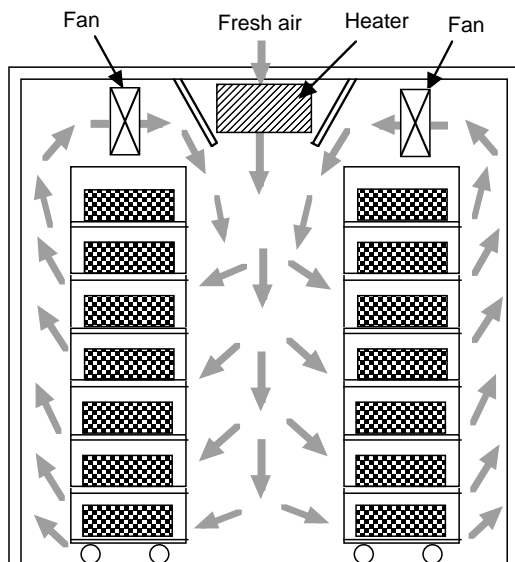


FIGURE 31.10 Configuration of typical batch dryers.

For the exhaust air, V_2 (m^3/s) becomes

$$V_2 = G_1(1 - X) \left(22.4 \left[\frac{1}{28.97} - \frac{H_2}{18.02} \right] \frac{T_2}{273.15} \right) \quad (31.32)$$

where R (kg/s) is the evaporation rate; H_0 and H_2 ($\text{kg-water/kg-dry air}$) are the humidity of fresh air and exhaust air, respectively; and T_2 (K) is the temperature of exhaust air. Dry air mass-flow rate circulating in dryers G_1 (kg/s) is given by

$$G_1 = \frac{V_1}{22.4 \left(\frac{1}{28.97} - \frac{H_1}{18.02} \right) \frac{T_1}{273.15}} \quad (31.33)$$

where V_1 (m^3/s) is the air flow rate in the dryers, H_1 is the humidity of air in the dryers, and T_1 is the temperature in the dryers.

Hot air temperature supplied to the dryers is generally between 370 and 420 K for batch dryers, and the logarithmic mean of temperature difference between air and ceramics in the constant rate period is approximately 30 and 100 K. The operating range of h_a lies between 230 and 350 $\text{W}/(\text{K} \cdot \text{m}^3\text{-dryer volume})$.

A dryer assembled with an air chamber, often known as the air-circulating unit, is shown in Figure 31.11. It is designed to improve the mixing of fresh hot air and recirculating air and ensure uniform air flow through the dryer chamber.

31.6.4 DRYING BY INFRARED HEATING

Materials are heated directly by infrared irradiation. Infrared lamps as well as hot rods and plates with efficient emission performance can be used as infrared sources. When wet materials are heated in the constant drying rate period by infrared radiation, the surface temperature increases to a temperature higher than the wet bulb temperature and the drying rate is enhanced. However, the drying operation must be

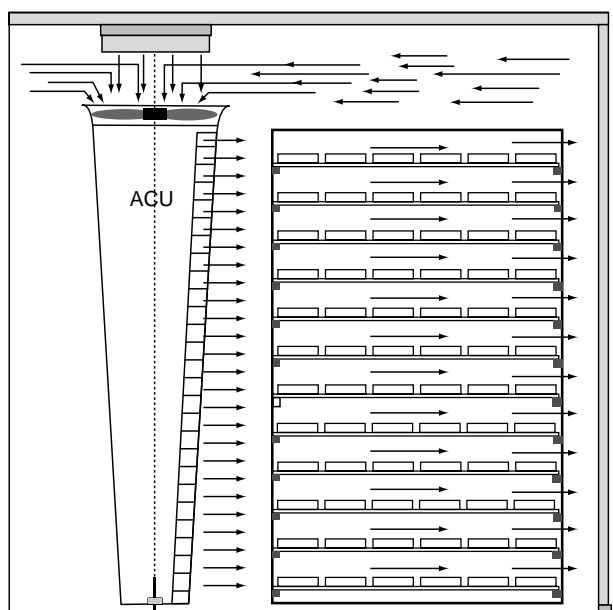


FIGURE 31.11 Advanced batch dryer with the air-circulating unit. (From catalogue of Novokeram.)

performed carefully as a rapid drying rate may result in the failure of ceramics due to shrinkage. An application of infrared heating and combined infrared-convective heating in the falling drying rate period is effective in enhancing the drying rate. It is easy to control the power of infrared sources. Thus, infrared heating is often applied to a complex configuration of ceramics with fins or waves avoiding the generation of cracks in a shorter time. It is necessary to control the power so that the sources facing a simple surface of ceramics are strong whereas those to a complex surface are weak [21].

Infrared heating systems are classified as open type, tunnel type, and closed type. The open type is the simplest system to complement, but its energy loss is the largest. The tunnel type installs infrared sources in tunnel dryers and is operated continuously. The closed type is operated in a closed room, and usually has the highest energy efficiency, as listed in Table 31.2, because the materials are not only heated

TABLE 31.2
Energy Efficiency of Infrared Heating for Various Types

Type	Energy Efficiency
Open	25–35
Tunnel	45–50
Closed	50–65

by direct irradiation of radiative energy from infrared sources but also by the convective heating of air heated by the sources.

31.6.5 DRYING BY MICROWAVE HEATING

The frequency of the microwave used for industrial heating comprises two bands of 2450 and 915 MHz in the broad countries. These frequencies are greatly effective for heating of water. Figure 31.12 shows a typical system of drying by microwave heating. Microwaves are generated in a magnetron, which is introduced into an applicator where the wet materials are heated via transmission through a waveguide. Using the waveguide, an isolator that is an absorber of reflection wave prevents the wave from returning to the generator and breaking it. A power monitor for measuring the output power and a tuner that adjusts to inhibit the reflection of microwave from the applicator are also installed. The merits of microwave heating can be summarized as follows:

- Wet materials are heated internally because microwave energy can penetrate the inside of the body.
- Only wet materials are heated selectively and high energy efficiency is expected.
- Rapid heating is possible.
- The operation and control processes are easy, and drying is possible in any atmosphere including a vacuum.

When a material whose relative dielectric constant is ϵ_r is placed in the electric field E (V/m) and frequency f (Hz), the heating rate per unit volume P (W/m³) is correlated to the equation

$$P = \frac{5}{9} \times 10^{-10} \epsilon_r f E^2 \tan \delta \quad (31.34)$$

where $\tan \delta$ is the dielectric loss tangent. The product $\epsilon_r \tan \delta$ is called as loss the factor. The half power depth D (m) represents the length penetrating the materials from the surface until the power dissipates to half, and is given by

$$D = 0.347 \frac{\lambda}{2\pi} \frac{2}{\epsilon_r (\sqrt{1 + \tan^2 \delta} - 1)} \quad (31.35)$$

where λ (m) is the wavelength.

Uniform drying of molded ceramics such as honeycombs is performed without deformation by controlling the microwave profile along the material length. In the drying of thick ceramics by microwave

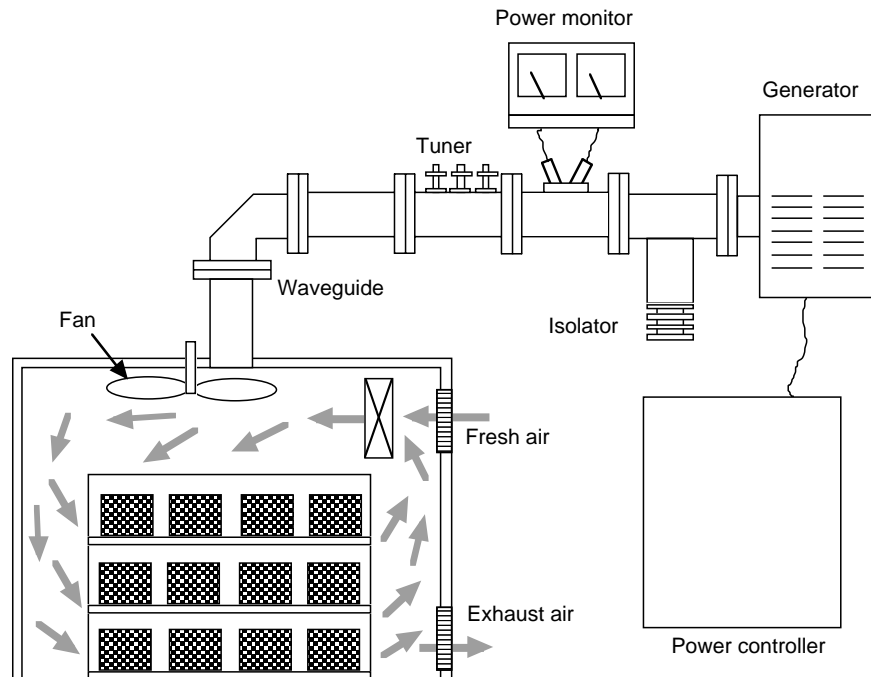


FIGURE 31.12 A standard assembly of microwave drying and batch dryer with the chamber of air-circulating unit.

heating, the material internal temperature may rise rapidly to a temperature higher than the surface temperature. This results in explosive expansion due to the internal vapor pressure. The microwave power should, therefore, be controlled appropriately to prevent overheating the inside of the body.

31.6.6 DRYING BY ELECTRIC CURRENT HEATING

Electric current heating is applied to the drying of cylindrically symmetric ceramics such as insulators. This heating method enhances the drying rate while controlling the crack formation by the effect of internal heating. In general, parts of fins or hoods of insulators dry up remarkably faster than the body, resulting in the formation of cracks due to drying-induced stress. As the electric current selectively heats the body rather than the hoods, drying up of hoods may be inhibited in comparison with the drying by hot air heating. If defects are still found, wrapping around ceramics is further introduced to control the drying rate. The operation takes place usually in charging the voltage of 100 to 400 V between both the edges depending on the size of insulators. The temperature of rods should be increased slowly in order to be maintained almost uniformly throughout the drying period until the maximum temperature is less than approximately 320 K. The typical electric current pattern changes in a drying process are

shown in Figure 31.13. In the earlier period, the current rises gradually due to shrinkage whereas it falls down in the later period because of the reduction of the moisture content. The peak current density is in the order of a few milliamperes per square centimeter. In this method, a special structure for adjusting the electrodes on the insulators may be necessary to ensure a good contact throughout the entire drying period even if the ceramics shrink.

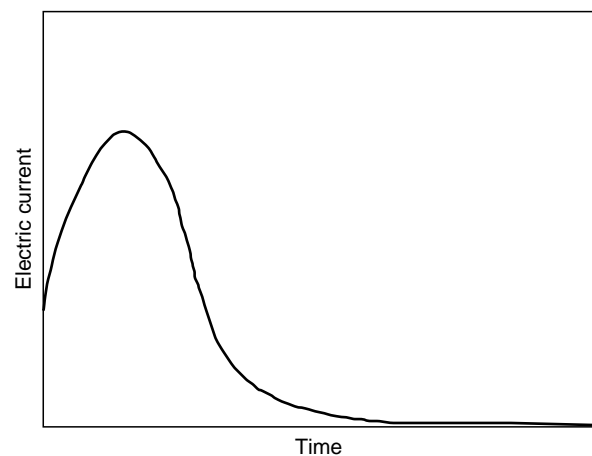


FIGURE 31.13 A typical pattern of current during electric current drying of insulators.

31.7 STUDIES ON FAILURE AND DEFORMATION OF CERAMICS

The generation of failure and deformation of molded clay are the undesirable problems requiring quality control during ceramic production. The mechanical properties of wet clay are important parameters to understand the mechanism of the drying phenomena. Many experimental data are available for a variety of clays [22–26]. The mechanical behavior of clay is generally described by the viscoelasticity or plasticity, which is dependent on the moisture content. In addition to the simulation of heat and moisture transfer, the strain–stress analysis is simultaneously required not only to predict the deformation and the failure of molded ceramics during drying but also to precisely design the mold in order to develop an effective process. Recently, theoretical studies have been conducted for one of the problems including the shrinkage of clay [17,18,20] while numerically simulating the clay-drying problem. However, research taking into account the failure generation and development of an effective drying process to improve the drying rate has been quite minimal. If the drying of molded clay is conducted too quickly, it may result in rapid surface dry-up while keeping a wet state in the inside. A fall in the fast-drying rate often causes the failure or the generation of deformation. These phenomena appear during the low moisture-migration rate in the clay.

Figure 31.14 shows an analytical result of the maximum tensile stress generated within a slab of drying clay with a physical length of 60 mm, width of 60 mm, and thickness of 24 mm [27]. The parameters of drying operation and ceramic properties are

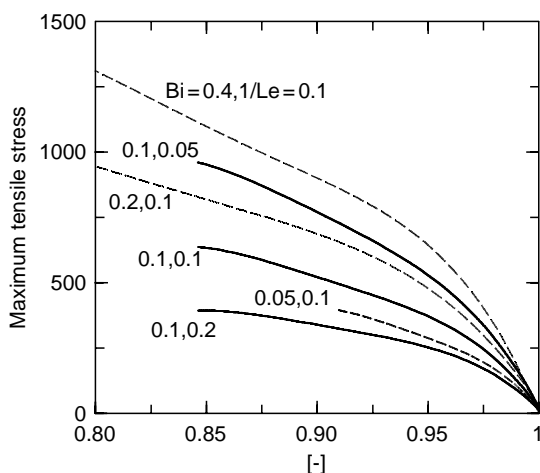


FIGURE 31.14 Maximum tensile stress analyzed during convective heat drying for a slab of $60 \times 60 \times 24$ mm.

represented by dimensionless parameters such as the Biot number (Bi) and Lewis number (Le). Assuming constant properties of the slab, the former number corresponds to the heating rate and the latter to the internal diffusion rate of moisture. Thus a higher Bi applies rapid heating rate whereas a higher $1/Le$ means larger effective diffusion coefficient of moisture in the body. \bar{W}^* denotes average moisture content normalized by initial content. The absolute magnitude of the maximum tensile stress should be considered on an arbitrary scale as it depends largely on the mechanical properties of the slab, but the dimensionless parameters of these properties are not available. Although a higher Bi results in remarkably stronger stress, the stress falls down with a larger $1/Le$. Hence, heat-transfer enhancement in the dryer and uniform drying are not possible during ceramic drying unless the moisture-migration rate in the clay is actively or passively enhanced by an auxiliary drying method such as the superheated steam drying. Therefore both experimental and theoretical studies are expected to make significant contributions to R&D of noble drying methods.

31.8 CONCLUSIONS

The R&D for the ceramic production utilizing advanced drying technology has been proposed. The drying of ceramics must be carefully carried out in order to keep their own molded configuration. If the drying process is hastened, it will result in serious problems such as drying up of the surface, undesirable deformation, and crack generation. Several researchers have studied the mechanism of moisture movement and shrinkage of clay. Some molding processes using the dry method have been developed. It may seem that the research works on the drying of ceramics have been almost complete. However, there are few systematic investigations on the quality control of ceramics during drying and on the development of the precise prediction method for the design of the molding and the dryers. Furthermore, the quality problems arising from drying have not been solved. It should be pointed out that there are still many R&D programs required to improve the drying process of ceramics.

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32 Drying of Peat and Biofuels

Roland Wimmerstedt

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32.1 INTRODUCTION

The use of biomass and peat, for both the industrial sector and the district heating, has greatly increased since the first oil crisis in 1973. The biomass utilized so far is wood and agricultural wastes, including bark, straw, and bagasse. Biomass grown especially for fuel purposes is still a technique in the experimental state. The use of peat as a fuel is limited to countries with domestic resources, such as the former Soviet Union, Canada, the United Kingdom, Ireland, Finland, and Sweden.

The moisture content of the fuel is an essential combustion parameter. Peat produced directly from the bog contains 80 to 95% water (counted on the total mass), which is reduced to about 50% through natural drying on the bog during summer. Biomaterials collected in forests typically have a water content around 50%, which is higher in fall and winter and lower during spring and summer.

The chemical composition of biomass and peat differs considerably from those of coal and oil mainly by the high content of oxygen. The content of volatile material is high. The combustion of biomass and peat can accordingly be divided into a drying stage (evaporation of water), driving off the volatiles that burn in the gas phase, and, finally, burning of the residue (charcoal). The moisture content of the fuel thus influences the drying period and, if the fuel contains water, part of the boiler system must be used to dry the fuel before burning. In Table 32.1, some typical figures for the combustion of a wood fuel containing

different amounts of water could be found. The table is based on a constant flue-gas temperature of 150°C and an entering air temperature of 40°C. The calculations are based on 1 kg of dry substance (DS).

As can be seen from the Table 32.1, the amount of flue gases drastically increases with increasing

TABLE 32.1
Combustion Parameters for Burning of a Moist Wood Fuel

	Moisture Content (%)		
	65	50	15
Water amount (kg/kg)	1.9	1.0	0.2
Excess air level (anticipated)	1.6	1.4	1.2
Higher calorific value (MJ/kg)	20.6	20.6	20.6
Lower calorific value (MJ/kg)	14.4	16.5	18.6
Flue-gas volume (1 bar, 0°C) (m ³ /kg)	10.3	8.8	6.2
Flue-gas loss (sensible heat) (MJ/kg)	2.1	1.8	1.3
Efficiency based on higher value	0.60	0.71	0.84
Efficiency based on lower value	0.85	0.89	0.93
Adiabatic combustion temperature (°C)	900.0	1200.0	1800.0

moisture content. This is due to the evaporated water and the need for higher levels of excess air at higher fuel moisture. The anticipated value of 60% at 65% moisture content is conservative, whereas 20% for almost dry fuel can be regarded as normal in practice. The higher amount of flue gases means a lower temperature in the boiler, as reflected in the adiabatic combustion temperature.

In the case of 65% moisture content, the adiabatic combustion temperature is so low that preheating of the combustion air would be necessary to ensure complete combustion. The combined influence of increased flue-gas volumes and decreased temperature level means that a larger and more expensive boiler is required when firing a fuel with higher moisture content.

The boiler efficiency can be defined in two different ways. In English-speaking countries, the usual way is to consider the latent heat of condensation of the water vapor in the flue gases as boiler losses. The boiler efficiency is thus defined as the quotient between the utilized heat and the gross (higher) calorific value. In other countries, such as Germany and the Nordic countries, the vapor losses are considered when defining the heat value of the fuel. The pricing of a biofuel or peat is thus made on the basis of its net (lower) calorific value and the boiler efficiency is defined as the quotient between the utilized heat and the net calorific value.

Finally, it should be borne in mind that the efficiency in [Table 32.1](#) neither includes the increased amount of unburned material in the ash from the moist fuel nor the increased radiation losses that result from the larger boiler required for the combustion of moist fuel.

32.2 REQUIREMENTS ON MOISTURE CONTENT OF FUELS

It is not obvious that a moist fuel must be dried before firing. On the contrary, this is considered uneconomical in many cases. There are, however, cases where drying is necessary for different reasons, and demands on the fuel moisture content must be raised.

One such case is the gasification of fuels. To ensure a high gas quality, the moisture content of a fuel to be used in a gasification process should not exceed 10 to 15%. Gasification of biomass and peat has hitherto been limited to limekilns in the pulp industry, but commercial-scale gasifiers for district-heating systems with cogeneration of electricity in combined cycle processes are gradually introduced on the market.

Another application where fuel drying is necessary is in the manufacture of fuel as powders, pellets, or

briquettes in “fuel factories,” for which the moisture content must not exceed 10 to 20%. The need for drying is partly due to technological reasons in the manufacturing process and partly due to storing and transport reasons. Storing high moisture-content wood chips can support large populations of fungi, many of which cause allergic reactions in humans [1]. Microbiological processes are also assumed to be the cause of spontaneous ignition in large wood-fuel piles. To avoid significant microbiological degradation during storage, the material should be dried to a moisture content of 20 to 25% [2].

The degree of drying required in direct combustion varies with the type of boiler used. If the fuel is to be used in small scale, 1 to 5 MW, it is preferentially fired as dried material in grate ovens with effective cooling of the grate. Grate-fired boilers of traditional type are dominating in the 5- to 20-MW area. If they are designed for moist fuels, the first part of the grate is designed as a band dryer. Flue gases are recycled to the first part of the grate, where they together with the radiating walls deliver the heat necessary for drying. If this type of boiler is fired with dry fuels, a cooling of the grate by the combustion air must be foreseen. At higher moisture contents the capacity is decreased; hence, if capacity is a crucial factor, a booster effect can be achieved by installing a predryer before the boiler.

In large-scale applications, fluidized-bed boilers are now dominant on the market. Both bubbling fluidized beds (BFBs) and circulating fluidized beds (CFBs) are used for biofuel firing. BFBs were originally limited to a rather narrow specification of fuels but now flue-gas recycling is used in both types of boilers. Especially, CFBs are considered to be tolerant to varying fuel qualities and, if this type of boiler is chosen, there would be no need to combine it with a dryer [3,4].

Suspension firing requires pulverized fuels containing less than 15% moisture to ensure complete combustion in the furnace. Dry fuel also means a high combustion temperature, which in turn means higher capacity in an existing boiler and a smaller and less expensive new boiler. This technique may be used when an oil-fired boiler is used to convert biofuels. It has also been used as a substitute for oil burning in limekilns in pulp mills, where this technique competes with gasification of the fuel. A combination of a dryer and a suspension-fired boiler may be seen as an alternative to installing a fluid-bed boiler.

In conclusion, there are cases where a fuel must be dried and a choice has to be made between different drying methods. In many cases, however, the primary choice is between the firing of wet fuel and the firing of dried fuel (i.e., between drying and not drying).

32.3 THERMODYNAMICS OF FUEL DRYING

Biofuels and peat can be dried according to different thermodynamic principles. The most common method is the use of flue gases. Another is the use of superheated steam. Finally, double-effect drying will be discussed.

Figure 32.1 shows the principles of flue-gas drying in combination with combustion in a boiler [5]. From the figure it can be seen that after the boiler the flue gases are taken through a fuel dryer in which all the fuel from the boiler is dried. The degree of drying is determined by the flue gases entering the dryer. By far the most common method is to dry the fuel using the sensible heat of the flue gases after the economizer down to a dryer exhaust temperature of 100°C . If the fuel has a moisture content of 60% (1.5 kg water/1 kg dry material) and the flue-gas temperature of 170°C , the fuel can be dried to about 55%. A considerable part of the available energy is used to heat the fuel with all its water to the wet bulb temperature of the flue gas.

A complete picture of the energy status for a plant as described in Figure 32.1 can be attained by combining dryer calculations with a combustion analysis. Results from such a calculation are shown in Figure 32.2, based on the assumptions from Figure 32.1 and an original exit flue-gas temperature of 170°C . Figure 32.2 displays the useful energy (i.e., energy that can be extracted in the boiler) counted per wet basis for a fuel originally containing 60% moisture. The point marks the energy status when there is no dryer. To achieve a high dryness, the flue gas must be taken from the boiler at temperatures seen on the lower horizontal axis. If, for example, a dryness of 0.3 is desired, the flue-gas temperature into the dryer must be about 410°C .

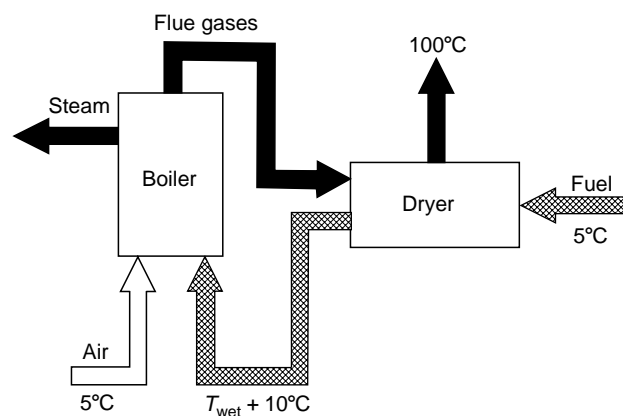


FIGURE 32.1 Principle of a flue-gas dryer in combination with a boiler.

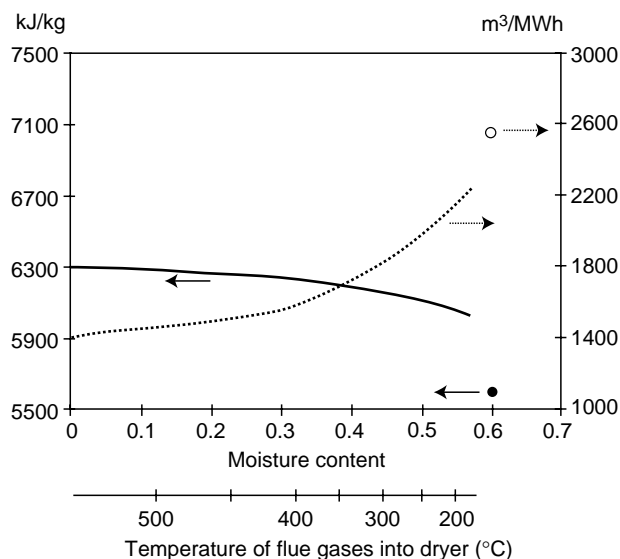


FIGURE 32.2 Useful energy and flue-gas volume when drying to different moisture contents in a flue-gas dryer (inlet fuel with 60% moisture content).

As can be seen, the primary improvement takes place when the exhaust temperature is lowered to 100°C . Further drying leads to marginal savings only, due to the decrease in the amount of excess air. The figure also shows the decrease in flue-gas volume (1 bar, 0°C) per MWh useful energy in the boiler. A number of papers discussing flue-gas drying of biofuels have been published recently [6–9].

Figure 32.3 illustrates the principle of steam drying. The steam necessary for the dryer is taken from the boiler and the evaporated water from the dryer is recovered as low-pressure steam. The economy of the dryer is dependent on whether or not the low-pressure steam can be used in the process. Figure 32.4 corresponds

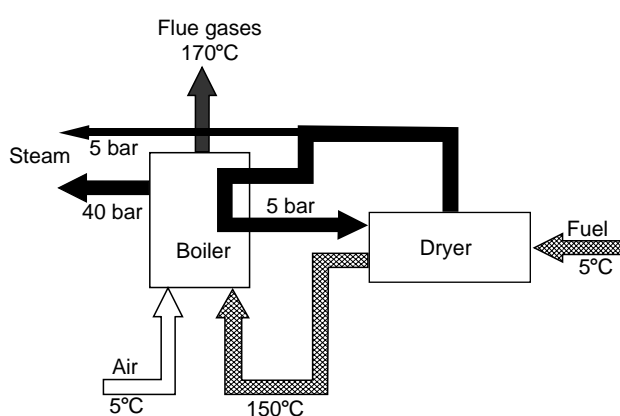


FIGURE 32.3 Principle of superheated steam dryer in combination with a boiler.

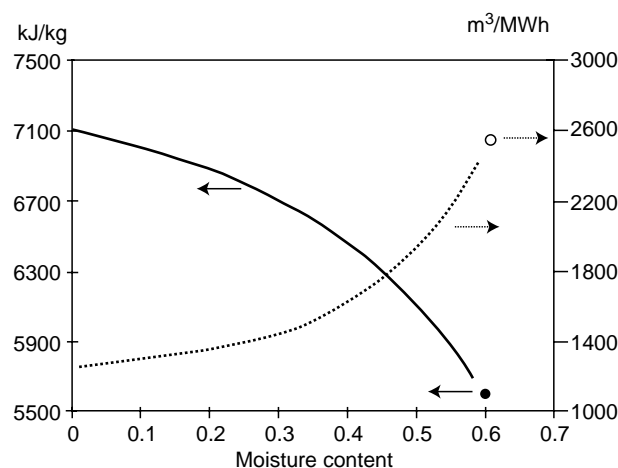


FIGURE 32.4 Useful energy and flue-gas volume when drying to different moisture contents in a steam dryer (inlet fuel with 60% moisture content).

to Figure 32.2. The underlying assumptions are: pressure in the dryer, 5 bar; the produced steam can be used down to condensate at 100°C; and the fuel entering the boiler has a temperature of 150°C. Radiation and leakage losses are estimated to be 5% of the total drying-energy consumption. It can be seen that this type of drying leads to considerable energy savings. The flue-gas volumes are reduced in the same way as for flue-gas drying.

In situations in which there is no use of the low-pressure steam, steam drying can be combined with mechanical recompression of the evaporated steam. A comparative study between dryers working at over-pressure and atmospheric pressure has been done

by Wimmerstedt and Hallström [5]. The energy consumption is lower at reduced pressure, but the capital cost for the dryer as well as the compressor is lower in the elevated-pressure case. To obtain a low energy consumption at elevated pressure, the condensate leaving the plant must be cooled by heat exchange to the moist material so that the temperature of this is increased. This kind of coupling presupposes that the ratio between electricity and fuel price is low.

The third possibility to be discussed here, mainly on grounds of history, is the double-effect dryer. An example of such a dryer is the Peco dryer developed in the 1930s and it has been used extensively for the drying of milled peat. The principle can be seen in Figure 32.5.

The moist peat (50 to 60%) together with ambient air is fed to apparatus II B. The heating medium is water at about 65°C. The air is exhausted from apparatus II A. The peat is then brought to dryer stage I consisting of three drying towers. The heating medium is steam at a pressure of about 4 atm. Air is recirculated in this effect and the humidity is condensed in the heat exchanger, the cooling water of which is used as the heating medium in dryer stage II. The specific steam consumption of the Peco dryer is 0.65–0.70 kg steam/kg evaporated water.

Peco dryers are normally combined with a special peat-fired boiler and a back-pressure turbine providing steam and electricity to the dryer and also electricity to the whole briquetting plant. A Peco dryer was built in Ireland as late as 1982. The high capital cost associated with the dryer and the need for a boiler and electricity-generating plant have, however, led to the Irish peat industry reassessing their choice in favor of rotary-type flue-gas dryers. The flue gases are produced in special flue-gas generators [6].

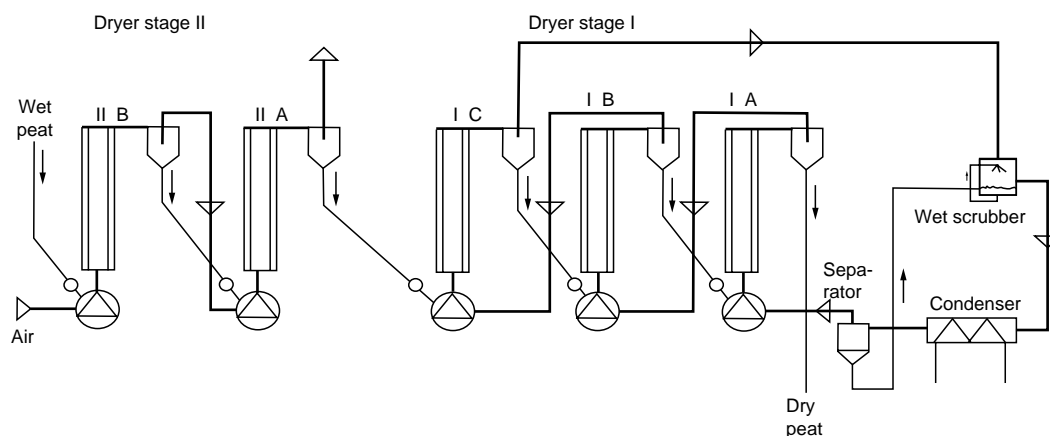


FIGURE 32.5 Two-stage Peco dryer.

32.4 FLUE-GAS DRYERS

Rotary drum dryers have been used for a long time in the drying of biofuels. Typical materials that have been dried are hog fuels, sawdust, and bagasse. Sawdust is also dried for the particleboard industry. Flue gases and fuel normally pass along the drum in a cocurrent flow. This means that the hot flue gases contact the moist material, which reduces fire hazards and emission of organic compounds. The drum rotates at a speed of 2 to 8 rpm. Factory-assembled units have diameters up to 4.5 m and lengths up to 10 m.

The dryers may be of the open-center type or the center-fill type. The former are equipped with lifting vanes or flights on the inside shell to carry fuel up the sides and disperse in the hot gas stream. In the open dryers the diameter is limited to about 2.5 m, and these provide a cheap and reliable system with low horsepower requirement. In the center-fill system, an internal structure helps to pneumatically convey the particles within the dryer. The center-fill dryer processes wet fuel according to its density. Particles of a particular moisture content are automatically maintained at a fixed bed temperature within a section of multizoned rotating cylinder. As moisture is evaporated from the particle it moves further along the dryer. This means that the dryer can handle particles of different sizes that need a broad spectrum of residence times, from a few minutes up to 1 h.

Rotary dryers are thus very flexible and come to extensive use in the particleboard industries and fuel factories. Another advantage is that the flue-gas

velocities through the dryer can be optimized to achieve a low fan horsepower requirement. To obtain a good thermal efficiency, part of the exiting gas is usually recycled over the dryer. This ensures a high moisture content of the exiting gas, which also is essential when the waste heat is utilized in, for instance, district-heating systems [3]. In Figure 32.6, a typical biofuel-drying system utilizing a rotary dryer is shown.

Cascade dryers came to extensive use, especially in the Nordic countries, during the 1970s and 1980s. Moist material is fed into the dryer with a flue-gas stream with high velocity. In the drying chamber, the flue gases cause the fuel to whirl around in a cascading bed. To maintain the cascade during partial load, flue gases are recirculated. The fine particles leave the dryer with the exiting gas and are separated in a cyclone. Coarse material is removed from the drying chamber by an overflow. The residence time in the dryer can amount from 1 to 2 min.

An advantage of the cascade dryer over the rotary drum dryer is the reduced space requirement, which may be critical when installing a dryer in combination with an existing boiler. Applications have been mainly as predryers in combination with wood-fuel boilers in saw and pulp mills.

Pneumatic conveying dryers or flash dryers have been used for a long time in the drying of milled peat, for instance, in the former Soviet Union and Finland. These units are directly fired and the peat is conveyed by the flue-gas flow through two or more towers, or open pipes, in total about 30-m long. The gas velocity is

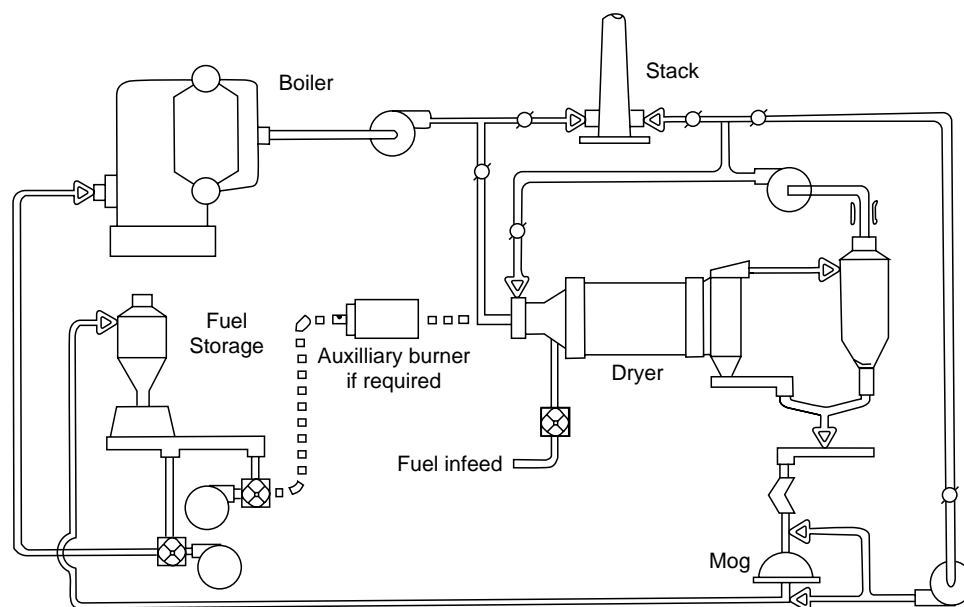


FIGURE 32.6 Biomass-fired system utilizing a rotary dryer. (From Rader Companies Inc.)

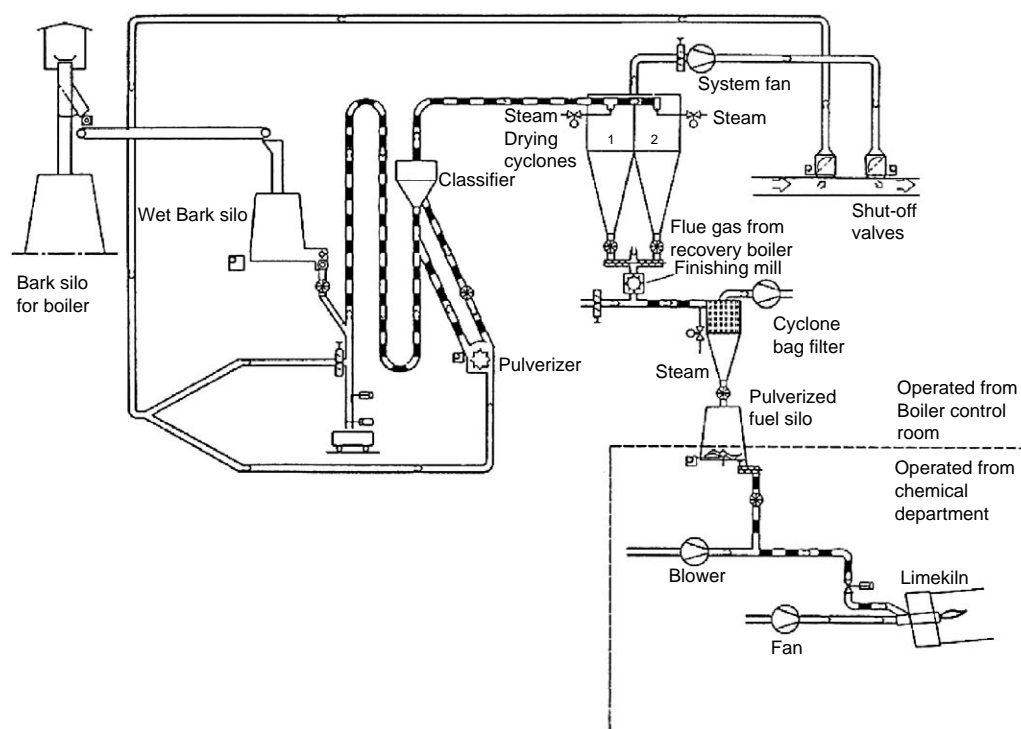


FIGURE 32.7 Limekiln fired by pulverized biomass produced in a BIOMASSTER. (From ABB-Fläkt.)

15 to 35 m/s. Some of these units also include a hammer mill and a short tower. The flue-gas flow is restricted so that only small, dry particles can be conveyed.

Pneumatic conveying dryers have been developed for simultaneous pulverization and drying of bark or hog fuel, for instance. The resulting powder with a moisture content of 10 to 15% can be used in a suspension-fired boiler or a limekiln in a pulp mill. In Figure 32.7, a bark dryer that delivers pulverized fuel to a limekiln is shown. Wet biomass is metered into the system via a rotary valve. It is conveyed by the flue gases, which in this case are delivered from a recovery boiler, up and down in two drying towers. Predrying of the material takes place here. This predrying reduces the surface moisture and the pulverizing process in the hammer mill is thus facilitated. In the classifier, coarse material is brought back to the hammer mill where extensive drying takes place as new, hot flue gases are added here. The material passes the milling circuit until it attains a particle size that allows it to pass the classifier. After this, the material passes through a final flash dryer and then dry fuel is separated from the flue gases in a cyclone.

The power demand for the mill and fan is reported to be 60 to 120 kWh/t DS [5] and the average residence time is about 30 s. Advantages of the system are its compactness, versatility (can be used for different

fuels), and low manpower requirements. A disadvantage is the high power requirement. Other flue-gas dryers that have been commercially used for peat drying are fluid-bed dryers and a combination of a fluid-bed and flash dryer (whirly bed dryer).

32.5 STEAM DRYERS

Steam drying of peat and biofuels reached a breakthrough during the 1980s. The commercial-scale applications reported so far are either of the indirectly heated flash type or of the directly heated fluid-bed type. Figure 32.8 shows a flow scheme of a flash dryer combined with a suspension-fired boiler in a district-heating application. Moist hog fuel is disintegrated and fed into a system with circulating low-pressure steam (2 to 5 bar). The material is conveyed with the steam through the dryer. The drying towers are essentially tubular heat exchangers with the drying material and low-pressure steam inside tubes of 75- to 150-mm diameter. Heat is transmitted from the outside of the tubes through condensing steam or from the cooling of hot water to the superheated transport steam. The temperature driving force is of the order of 40°C or higher. The amount of transport steam that corresponds to the change in moisture content of

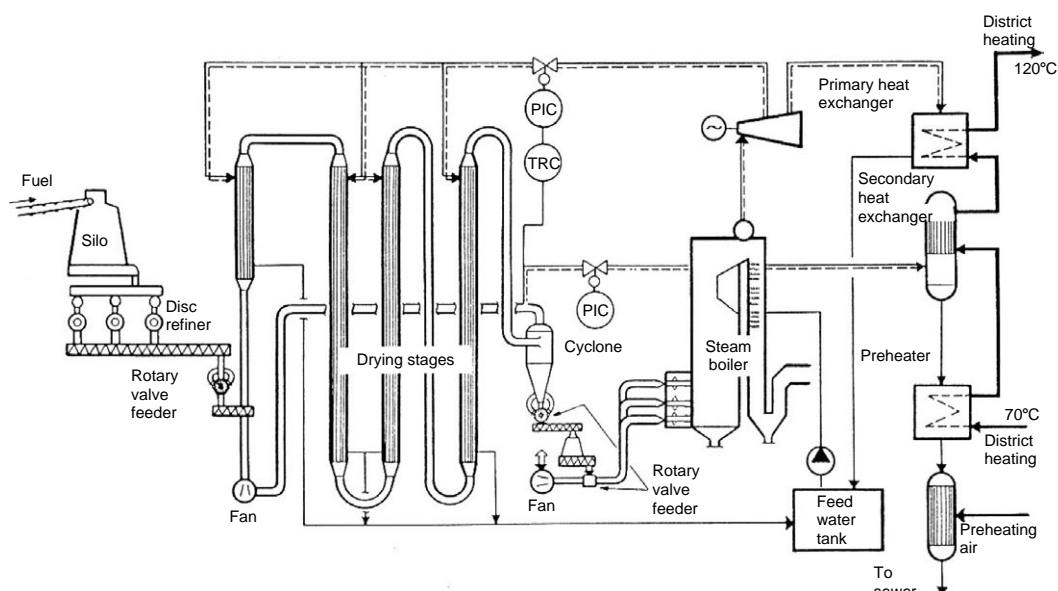


FIGURE 32.8 Steam dryer for biomass installed in a district-heating plant. (From MoDo-Chemetics AB.)

the fuel is extracted at the system pressure. The steam is separated from the material in a cyclone and then transported to a condenser where water for district-heating purposes is warmed. The product condensate is further cooled and then led to a sewer. The steam velocity in the tubes is of the order of 30 m/s. As the steam has a higher density than the flue gases due to higher pressure, the power demand for the fan is normally higher than for a flash dryer of the flue-gas type. The electricity consumption for the disintegration and transportation of a hog fuel that is dried from 60 to 10% moisture content is quoted as being 140 kWh/t DS [10]. Further development of this type of dryer includes an enlarged presuperheater and decreased number of drying towers, which contributes to a decreased electricity consumption [3].

During the 1990s, two steam dryers of the fluid-bed type were taken into operation in Sweden. These were the same type of dryers that have successfully been developed by NIRO for drying of sugar beet pulp [11]. The principle of this dryer is shown in Figure 32.9. The dryer is cylindrical with drying cells around the periphery. The drying steam is circulated up through the fluid bed and down again through a superheater installed in the center of the vessel. The residence time varies from a few seconds for small particles that are entrained in the drying steam to about 10 min for coarse material that is transported in the bed.

The capacity of steam dryers is determined by the heat transport, and thus by the driving temperature force between the condensing heating steam (which is

the normal heating medium) and the circulating steam. This means that the pressure of the condensing steam is crucial to the cost of the dryer. For example, the capacity of a dryer with an operating pressure of 4 to 5 bar is doubled when the pressure of the heating

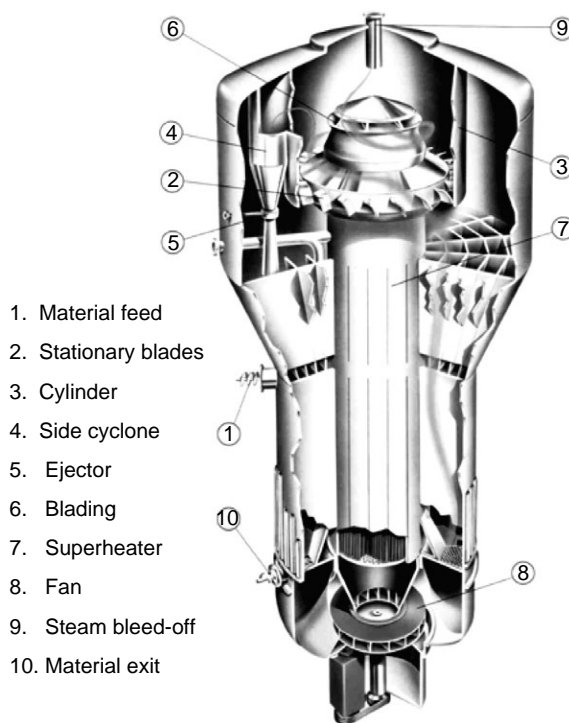


FIGURE 32.9 Fluid-bed steam dryer of the NIRO-type.

steam is raised from 12 to 25 bar. This is true for both flash and fluid-bed dryers. The conclusion is that the price information for a steam dryer must always be related to both the capacity and the heating steam pressure.

When steam drying is used in situations where there is no demand for waste energy in the form of low-pressure steam, mechanical vapor recompression (MVR) can be applied as mentioned earlier. Such a plant has been in operation at Härjedalen, Sweden, since 1988, where peat is dried in a flash-steam dryer. The production of briquettes is based on air-dried peat with 60% moisture and amounts to 300,000 t DS per year with 10% moisture. The dryer unit consists of two identical lines with a capacity of 20 t of DS. The dryer in each line has five heat exchangers in series with a tube length of 20 m. The total heat-transfer area is 2700 m². Turbo-type compressors are used with a compression ratio of 1:4.7 from 3 to 14 bar; the system pressure is 3.6 bar giving a temperature driving force of 60°C. The total electricity consumption is 270 kWh/t DS [12].

32.6 ENVIRONMENTAL ASPECTS

Increased use of biofuels and peat, which can be facilitated by efficient drying methods, means in itself a mitigation of the greenhouse effect. A further advantage is that no emission of sulfur oxides takes place. The emission of NO_x is strongly dependent on the combustion temperature. A dried fuel would hence mean increased emissions of NO_x. This may be partly counterbalanced by the decreased amount of excess air and better possibilities for control of the combustion process achievable with a dried fuel. The low temperature at the combustion of very moist fuels might lead to increased amounts of unburned hydrocarbons in the flue gases.

Many sources report on fewer particulate emissions when burning dry fuels. This is especially true for old grate-fired boilers. This is again achieved by the higher combustion temperature, which enables the smaller particles to burn faster and more completely. Drying of bagasse prior to firing has been found to reduce particulate emissions from boilers by roughly one half [7].

Drying of biofuels and peat, however, also leads to emissions of organic compounds. In principle, all volatile organic compounds in the material to be dried might vaporize during drying. The amount of volatile material is 60 to 80% in wood and 50 to 70% in peat. In most cases, however, drying takes place at such conditions where only the most volatile compounds can vaporize; but, nevertheless, biofuel drying is contributing to the release of volatile organic compounds

(VOC). Of most concern is the emission of terpenes. Terpenes are hydrocarbons present in conifer oleo-resins. They could also be found in many hardwoods, especially of tropic origin. Natural emissions take place everywhere in our forests and the quantity released naturally is much higher than the anthropogenic emissions. The problem with the latter is the very high local concentrations that can be achieved. Terpenes have a high reactivity in the atmosphere and may contribute to the formation of ground-level ozone.

In the last decade, there has been quite an extensive research in this area and there is now a good understanding of the problem [13–16]. The emitted compounds, at normal drying temperatures, are volatile extractive compounds, mainly terpenes, carboxylic acids, and light aldehydes and alcohols. The emitted amounts are principally controlled by the material temperature, and also by the type of material. Sawdust originating from pine has, for instance, a much higher terpene content compared with sawdust from spruce. The storing time of the material after harvesting is important for finely divided materials like sawdust and flakes as the terpenes are easily released from those materials. It is practically impossible to dry a material without releasing about 80% of the original terpenes. Experiments with batch fluid-bed drying of sawdust showed a very typical behavior. The terpene release showed one dominating peak directly after the first air contact, during the rest of the constant rate drying the release was on a low, stable level. A second peak, much smaller than the first, could be seen during the final part of the drying with increasing material temperature [13].

The nature of the emission is, as mentioned, controlled by the material temperature. This temperature is normally below 100°C for flue-gas dryers whereas a typical value for a steam dryer is 140°C. On the other hand, the gas temperature is much higher in flue-gas dryers. Small overdried particles might therefore be exposed to very high temperatures. The occurrence of blue haze in rotary flue-gas dryers is a result of pyrolysis of small particles.

Most of the VOCs released during drying in a steam dryer could be found in the condensate; the flow of noncondensibles is rather small and is normally taken to incineration in the boiler. Also, the effluent from a traditional flue-gas dryer could be taken to a condenser for heat recovery and gas cleaning. The remaining inert-gas flow is, however, much higher in this case. If further cleaning is required by the national legislation, this could be achieved by using a wet electrostatic precipitator followed by a regenerative thermal oxidation unit. Also closed-loop drying systems are being developed [3]. In these systems, the circulating dryer gas in a rotary dryer is

indirectly heated by hot flue gases from the boiler in a heat exchanger. The bleed from the circulation is incinerated in the boiler. The circulating gas stream has a very high dew point and heat could be recovered from the bleed via a condenser. This type of dryer is essentially a steam dryer working at atmospheric pressure and thereby avoiding problems associated with feeding the materials into a pressurized system.

32.7 OPERATION EXPERIENCES

In Sweden, a number of plants for drying of peat and biofuels are in operation. In a recent study [17], operation experiences from eight plants were collected and evaluated. Four of the plants had steam dryers and the other four had flue-gas dryers. The total availability of the flue-gas dryers was typically 80% or higher whereas the corresponding values for steam dryers were 35 to 90% with an average around 70%. The highest value was for the peat dryer mentioned earlier, which was installed already in 1986, whereas the recently installed dryers had a very low availability. Common problems of all dryers, steam dryers as well as flue-gas dryers, were corrosion and erosion. The condensate in biofuel dryers is acidic and corrosive. All parts of the dryers, where condensation can take place, must be either lined with or manufactured in stainless steel materials. In the original design, carbon steel had been the normal construction material. The drying material is also erosive due to sand particles and other inorganic materials. Especially, bends and pipes with high flow velocities are easily eroded and in one case lining with wear-protection plates is part of the regular maintenance.

Common problems with the steam dryers are related to the feeding-in and feeding-out devices. Original cell feeders did not work satisfactorily and had been substituted by plug-screw feeders. In some cases, these must also be continuously exchanged due to wearing.

The dried product is highly inflammable and fires could occasionally take place, and in some cases water nozzles have been installed as a fire extinguisher. In steam dryers it is well over 100°C during drying, and opening the dryers with air supply must be avoided until the material has cooled off.

The general impression from the experiences is that maintenance must be carefully planned for these type of dryers. A tight dialogue with the manufacturer and taking advantage of experiences from earlier installations seem highly recommendable.

32.8 EQUILIBRIA AND KINETICS

Equilibrium moisture contents of wood have been presented in a number of published papers. The interest has been caused by the drying of timber. Wood is a moderately hygroscopic material at moisture contents below about 0.3, at which the free water has been removed and the residue can be regarded as cell-bound water. During the last few decades, considerable interest has been devoted to the drying of timber in superheated steam and a number of papers on equilibrium moisture content at high temperatures have been published [18–20]. In Figure 32.10, a typical set of data are presented [20]. Figure 32.10a shows the original way of presentation and Figure 32.10b shows the presentation for which the data has been recalculated and presented as isobars. As is

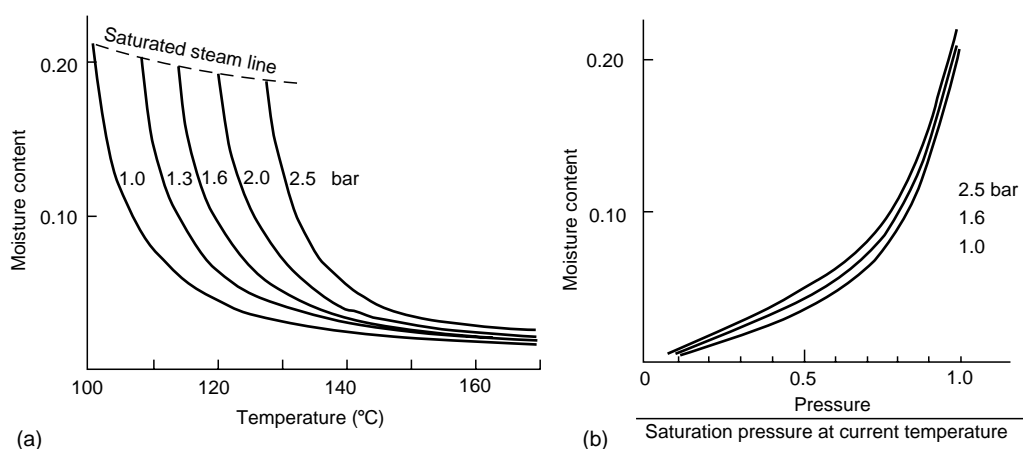


FIGURE 32.10 Equilibrium moisture content of wood at different temperatures and pressures.

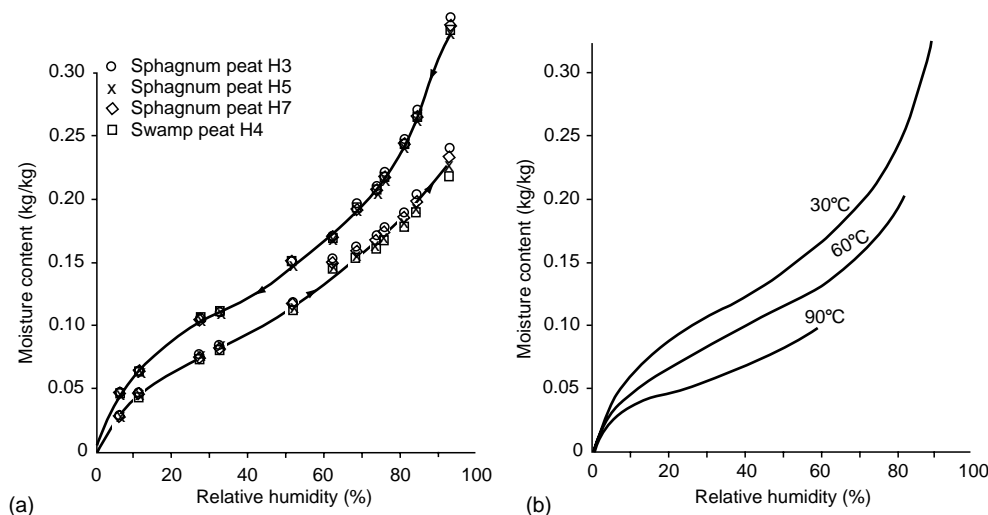


FIGURE 32.11 Equilibrium moisture content of peat.

evident, the latter way of presentation gives a much smaller parameter influence.

A number of equilibrium moisture data have been found in the literature for peat also. Larsson and Wimmerstedt [21] presented a data for peats of different origin and humification. Results from the investigation are presented in Figure 32.11. The sorption and desorption isotherms at 30°C and for different peats are shown in Figure 32.11a. As can be seen, the similarity between different peats is pronounced. As can be expected from a biological material, a considerable hysteresis effect is demonstrated. Figure 32.11b presents the "average" isotherms valid for desorption of different peats. A pronounced temperature dependence over the whole relative-humidity range can be seen. The shapes of the isotherms are similar to those of wood and imply that most of the water is not strongly bound. It can be concluded that the hygroscopicity is of rather limited importance for the drying process.

Rather few papers deal with the kinetics of biofuel drying and most designs seem to be "experience-based." Bagasse is reported to be easily dewatered with exit temperatures for commercial rotary dryers approaching the wet bulb temperature [7]. A study of drying rates of milled peats [21] in a fluid-bed bench-scale dryer showed no influence on the origin of the peat. On the other hand, different size fractions of the same peat gave quite different results. The intraparticle resistance is pronounced and, of course, controlled by the particle size.

Similar results are reported from the drying of bark and peat in superheated steam in a pilot-plant pneumatic conveying dryer [22]. The results are presented as a convective apparent heat-transfer coefficient,

defined with the assumption that the temperature of the particle surface coincides with the saturation temperature of the transport steam. This transfer coefficient shows a clear dependence on the moisture content of the particles and the particle sizes. Fyhr [23] presented a model for a pneumatic conveying steam dryer. The dryer model consists of two submodels, one for the single particle and the other for the hydrodynamics of gas and particles in the dryer.

Hermansson et al. [24] reported the results from bench-scale drying of bark and wood chips in superheated steam in a fixed bed. The results are presented as a thermal efficiency defined as the ratio between the time-averaged steam temperature decrease over the bed and the maximum obtainable temperature decrease. This efficiency proved to be almost independent of pressure and temperature. When the mass load of the bed exceeded 30 kg/m² the thermal efficiency was above 85%, even at mass fluxes as high as 0.6 kg/m²/s.

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33 Drying of Fibrous Materials

Roger B. Keey

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33.1 NATURE OF FIBERS

Fibers are regarded as very elongated particles. Staple fibers that are spun into yarns have length/diameter ratios greater than 10,000. The fiber lengths of cotton are of the order 25 to 75 mm; wool fibers may exceed 100 mm and are variable in length even when shorn from the same sheep; flax fibers may be available in lengths up to 1 m [1]. The corresponding diameters of textile fibers range between 3 and 500 μm .

Fibers vary in cross-sectional shape, both naturally and by design. Wool fibers are essentially round and cotton fibers are elliptical. Synthetic fibers made by melt spinning can be of a desired shape. Artificial fibers that are spun from solvents in air or from an aqueous medium are usually irregular in shape because of the skin-core effect [2]. Rayon, for example, can have both regular and irregular cylindrical forms composed of hollow as well as solid fibers.

The cross-sectional shape influences the way the fibers pack together in yarns. Silk fibers, because of their triangular section, can pack compactly to give small-diameter, dense yarns. Natural fibers that grow in short lengths and are spun into staple yarns are very rarely straight. Cotton and woolen yarns naturally have a spiraling crimp. Textile fabrics

are composed of interlocking threads in a gridlike pattern produced by weaving or knitting individual strands.

Wood consists of a large number of fibers, together with cells of other types, bonded together by lignin to form a solid and rigid structure. The fibers are hollow, cylindrical structures, typically 1 to 4 mm in length, with walls composed mainly of cellulose and its associated polysaccharides. Most woods shrink and swell with moisture content, but the dimensional change is much smaller along the fiber length compared with the changes across the fiber by a factor of 50 to 100. Pulping, by chemical or thermochemical means, delignifies the structure, releasing individual fibers. Paper results from screening, draining, and drying the macerated mix to give sheetform material of intertwined fibers. Wood fibers are also hot pressed with resins, mainly urea formaldehyde, into reconstituted timber products such as fiberboards and hardboards.

33.2 MOISTURE IN FIBERS

The amount of moisture adsorbable by the fibrous material varies markedly, as shown in [Table 33.1](#). Hydrophilic fibers of natural origin can take up

TABLE 33.1
Smoothed Values of Dry-Basis Moisture Content (kg/kg) for the Adsorption of Water Vapor at 30°C onto Fibers

Fiber	Relative Humidity, $\varphi = p/p^0$		
	0.2	0.5	1.0
Casein	0.0615	0.1115	1.05
Cotton	0.0305	0.0565	0.23
Cotton, mercerized	0.042	0.0775	0.335
Nylon 6.6, drawn	0.0127	0.0287	0.05
Orlon (50°C)	0.0031	0.0088	0.05
Rayon, cuprammonium	0.0515	0.0935	0.36
Terylene yarn	0.014	0.037	0.03
Viscose yarn	0.0555	0.101	0.46
Wood pulp	0.034	0.062	0.25
Wool	0.062	0.11	0.38

Source: Data from Currie, J.A. 1969. Thermodynamic properties and irradiation studies of high polymers, Ph.D. thesis, Northwestern University, Evanston, IL.

considerable amounts of moisture, whereas some artificial fibers are barely hygroscopic. The variation of equilibrium moisture content with relative humidity at constant temperature is shown in Figure 33.1 for a number of fibers.

At low relative humidities ($0 < p/p^0 < 0.35$), water is adsorbed monomolecularly by many natural fibers. The equilibrium moisture content X_e then relates to the fraction of available sites taken up, i.e.

$$X_e = \phi X_1 \quad (33.1)$$

where X_1 is the moisture content for a fully completed monomolecular layer. Cassie [3,4] considers that water can be sorbed into hydrophilic fibers in one of two states: either strongly localized or liquid-like. Through a statistical thermodynamic analysis this concept leads to the Brunauer–Emmett–Teller (BET) relationship:

$$\varphi = X_e/X_1 = C\varphi/(1 - \varphi)(1 - \varphi + C) \quad (33.2)$$

in which φ is the relative humidity or relative vapor pressure p/p^0 and C is a coefficient that relates to a partitioning between bound and liquid water. Windle [5] has extended Cassie's analysis to include three distinguishable forms of sorbed water. Other elaborations include the possibility of multiple molecular sorption layers [6].

Schuchmann et al. [7] suggest that Equation 33.2 can be put in a more general form

$$X_e = ax/(1 + bx)(c + x) \quad (33.3)$$

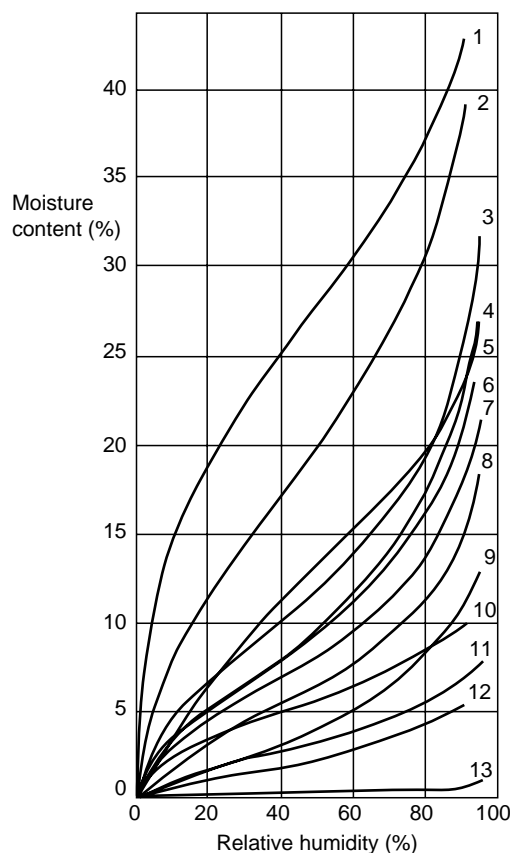


FIGURE 33.1 Sorption isotherms for textile fibers: (1) beryllium alginate, 25°C; (2) calcium alginate, 25°C; (3) viscose, cellulose acetate, cupraammonium rayon, and woolen yarn, 25°C; (4) casein fiber, 25°C, wool, 35.8°C; (5) jute; (6) mercerized cotton, 20°C; (7) flax, 30°C, hemp; (8) steeped cotton, 20°C; (9) acetate rayon, 25°C; (10) linen; (11) perlon, nylon, 25°C; (12) cellulose acetate; and (13) Pe–Ce rayon 20°C. (From Krischer, O., and Kast, W. 1978. *Die wissenschaftlichen Grundlagen der Trocknungstechnik*, 3rd ed., Springer-Verlag, Berlin, p. 57.)

where a , b , and c are empirical coefficients considered to be “shape characteristics.” The dependent variable x is the relative humidity when the parameter is low, and thus Equation 33.3 reduces to Equation 33.2 with $c = 1$. At very high relative humidities,

$$x = -\ln(1 - \varphi) \quad (33.4)$$

This substitution enables relative humidity data for a variety of food materials (including apple fiber) to be satisfactorily correlated up to $\varphi = 0.98$.

Another modified BET equation has been derived by assuming that moisture adsorption occurs at randomly located, equal-size active sites and there is no interaction between adsorbate entities [8]. These assumptions lead to the expression

$$X_e/X_1 = C\varphi/(1 - k\varphi)(1 - \varphi + C\varphi) \quad (33.5)$$

When $k = 1$, this expression reduces to Langmuir's equation for monomolecular adsorption. If $0 < k < 1$, there is a finite maximum hygroscopic moisture content. While many textile fibers approach such a moisture content asymptotically at high relative humidities, some man-made fibers such as nylon and viscose appear to have well-defined maximum hygroscopic moisture contents [9]. In many cases, the coefficient k is greater than 1. Jaafar and Michalowski [8] interpret this behavior as the thermal effect of adsorption being equal to the heat of condensation only after a multimolecular layer has been formed.

As implied by the proposal of Schuchmann et al. [7], many adsorption isotherms may be normalized by plotting the equilibrium moisture content X_e against the free energy change of sorption ($-RT \ln \varphi$) [10,11]. It is believed that desorption isotherms can be similarly correlated, particularly at moderate relative humidities when multimolecular adsorption is the dominant mechanism of attachment as the moisture content is directly related to the thickness of the adsorbate layer. Tests with particulate materials have shown that there is a linear relationship between the free energy change and $\ln X_e$ over a sixfold range in the latter [12]. This correlation implies an explicit relationship for the equilibrium moisture content of the kind

$$X_e = A \exp(BT \ln \varphi) \quad (33.6)$$

where A and B are empirical coefficients. Equation 33.6 predicts that there is a maximum hygroscopic moisture content (when φ becomes unity) that takes the value A . In those cases when this moisture content is ill defined and the equilibrium value increases rapidly with relative humidity, a form of Henderson's equation [13] is preferred by Papadakis et al. [14]:

$$X_e = A - [B/T] \ln(1 - \varphi) \quad (33.7)$$

Henderson's equation is based on the ratio of the amount of adsorbed moisture per unit wetted surface to the energy change on adsorption, this ratio being assumed to be a function of the equilibrium moisture content. Papadakis et al. [14] find that Equation 33.7 fits data for two kinds of cellulose over a range of relative humidity from 0.113 to 0.946 and a range in temperature from 20 to 93°C. An important finding of their work is that a correlation may fit data over a limited range in relative humidity very well, but can give misleading results if extrapolated beyond the tested range, particularly to higher relative humidities when different moisture-retention mechanisms take place. Walker [15] makes the same point with regard

to water in wood. Strictly, the fiber saturation point corresponds to the maximum hygroscopic moisture content when the cell walls of the fibers are fully saturated. However, at relative humidities above 0.98, the lumens and pits of the tracheid fibers begin to fill by capillary condensation, causing a sharp upward break in the sorption curve.

Since adsorbed moisture evaporates more readily at higher temperatures, the equilibrium moisture content becomes smaller with increasing temperature at a given relative humidity. Desorption isotherms, for instance, of a never dried softwood such as sitka spruce suggest that the fiber saturation point falls from about 31% at a temperature of 25°C to 23% at 100°C [16]. Under kiln conditions, the equilibrium moisture content becomes very low: at a dry bulb temperature of 120°C and a wet bulb depression of 30°C, Hilderbrand gives this moisture content as being only 3%. (Indeed, in using an oven-drying test to determine the moisture content of a fibrous mass, it is frequently assumed that the residual "moisture" is negligible. Difficulties in using weight loss methods to determine the moisture contents are discussed in Reference 17.) Shubin [18] presents a useful chart of equilibrium moisture content for wood covering vapor pressures to 1 MPa and temperatures to 180°C.

Adsorbed moisture can be held very tenaciously by natural fibers. Nuclear magnetic resonance studies have shown that adsorbed water on green and remoistened wood can exist in two states, with an immobile monolayer bonding directly to the cell walls of the fibers [19]. For most woods the differential heat of sorption is about half that of vaporization, falling to about one quarter of that value when a complete monolayer is formed at about 4 to 5% moisture content [15].

The heat of sorption is the difference in specific heat content or enthalpy between the bound moisture and that freely attached at the same temperature and total pressure. This enthalpy difference is normally derived from a form of the Clausius–Clapeyron equation on the assumption that the moisture vapor phase acts like an ideal gas and the molal volume of the condensed phase is negligible compared with that of the vapor. These considerations lead to the expression

$$\Delta H_w = -R \frac{\partial \ln \varphi}{\partial (1/T)_x} \quad (33.8)$$

It follows that the heat of wetting can be found by plotting $-\ln \varphi$ against $1/T$ if sorption data are available at various temperatures. However, at low equilibrium moisture contents (for wood fibers <7%), this procedure is inaccurate and direct calorimetry is preferable [20].

The free energy change $-RT \ln \varphi$ is sometimes used as an approximation to the heat of wetting. For this approximation to be valid, the equilibrium relationship must obey a degenerate Antoine expression

$$\ln \varphi = a/T + b \quad (33.9)$$

where a is a constant and b is zero. In many cases, b will be nonzero, and a function of moisture content. The difference between the free energy change and the enthalpy associated with adsorbed moisture has been associated with the entropy changes accompanying the dimensional changes as fibers swell on the uptake of moisture [21].

Moisture can have a profound effect on the mechanical properties of fibers. Water imbibed by the cell walls of fibers causes swelling and moisture loss causes shrinkage. Loading, including the development of drying stresses, introduces a creep strain that may not fully recover when the load is released. This is known as *permanent set*. With hydrophilic materials, moisture is found to reduce stiffness and increase creep, possibly as a result of plasticization. Variations in moisture content enhance creep. Changes in the rigidity of wool fibers undergoing both adsorption and desorption of moisture have been reported by Mackay and Downes [22]. The creep of aromatic polyamide fibers at constant moisture content is found to follow the logarithmic relationship [23]

$$\varepsilon(t) = a + b \log t \quad (33.10)$$

where $\varepsilon(t)$ is the strain, a function of time t . If the fibers are exposed to an environment in which the relative humidity is cycled between limits, then considerably enhanced strain rates are found [24]. At 60°C, recycling between 5 and 95% relative humidity produces the same strain in 2 days as that expected in 5 y when the material is kept at the higher relative humidity.

Wood fibers shrink anisotropically on drying below their saturation point. Walker [15] cites some possible reasons for this behavior, including the possibility that microfibrils in the cell wall restrain the cell wall matrix of lignin and hemicelluloses, and differences in the behavior of earlywood and latewood between growth rings. The transient effects of moisture on the strength of products composed of wood fibers are reviewed by Back et al. [25]. The effect of drying on the properties of wet wood pulp fibers is summarized by Kumar and Mujumdar [26]. Drying weakens the fiber mat causing a substantial reduction in breaking length. However, a study of the drying of

simple, virgin fibers shows an improvement in tensile strength, but predried fibers of less hemicellulose content do not [27]. Optical properties are also affected by drying, with a decrease in light-scattering ability.

Fibrous materials are dried commercially in superheated steam as well as air [28]. Materials include cellulose, corn gluten, and sugar beet pulp. Superheated-steam drying has the advantages of lower energy use compared with air drying, absence of oxidation, and less contamination of the product. Bernardo et al. [29] provide some data on the relative coloration of layers of sugar beet fiber when dried in air at temperatures up to 105°C compared with drying in superheated steam at temperatures up to 150°C. The white color of the fibers dried with hot air is preserved, but yellowing of the fibers occurs with superheated steam at dry matter contents >80%. This reaction may be caused by an initial rehydration, producing traces of melamines.

33.3 MOISTURE MOVEMENT IN SINGLE FIBERS

From thermodynamic reasoning we expect the movement of water through a single fiber to occur at a rate that depends on the chemical potential gradient. For movement in one direction, the flux of moisture may be written as

$$J = Bc \frac{\partial \mu}{\partial x} \quad (33.11)$$

where B is some coefficient, c is the total concentration, and μ is the chemical potential gradient. Equation 33.11 can be reexpressed in terms of the concentration gradient by

$$J = - \left(\text{BRT} \frac{\partial \ln a}{\partial \ln c} \right) \frac{\partial c}{\partial x} \quad (33.12)$$

on introducing the activity of the sorbed moisture. The term in parentheses is the diffusion coefficient, and Equation 33.12 is a form of Fick's first law of diffusion.

In the hygroscopic moisture regime, when the activity is not equal to the concentration, the diffusion coefficient can become highly concentration dependent. Further, at low moisture contents, sorbed water may form strong bonds with a hydrophilic fiber, so that simple diffusion can no longer occur. The concept of sorptive diffusion has been introduced to describe the way moisture might migrate under these conditions [30]. Only those molecules with kinetic energies greater than the activation energy of the moisture-fiber bonds can shift from one site to another. The driving force for sorptive diffusion is considered to be the sorptive pressure, which acts

over two-dimensional regions in a similar way to vapor pressure acting in three-dimensional space.

However, diffusion coefficients based on concentration gradients are still commonly employed as a means of describing the rates of moisture movement. In general, the diffusion coefficient will be a function of both concentration and temperature. At constant temperature, the diffusion coefficient is only independent of concentration if the moisture isotherm is linear or the moisture unbound. Isotherms for many fibers are approximately linear over the range of relative humidity from 20 to 80%, as the data in [Figure 33.1](#). Thus, in a number of practical applications, the assumption of a concentration-independent diffusivity can lead to useful results.

Fick's first law can be expressed in terms of the dry basis moisture content gradient by noting

$$J = -(D) \frac{\partial}{\partial x} (\rho_s X) \quad (33.13)$$

where ρ_s is the density of the solid matrix. This density is a function of the moisture content as the fiber swells or shrinks in response to the moisture present. Generally, only the bone-dry fiber density $\tilde{\rho}_f$ is known, and Equation 33.13 is transformed into Equation 33.14:

$$J = -(D) \tilde{\rho}_f \frac{\partial X}{\partial x_f} \quad (33.14)$$

The diffusion coefficient is now concentration dependent, reflecting dimensional changes in the fiber. Such changes can be accommodated by appropriate definitions of length coordinates to avoid shrinkage effects within the diffusion coefficients [31].

Notwithstanding a variety of cross-sectional shapes that are found, a fiber may be viewed, to a first approximation, as a long circular cylinder, either hollow or solid. Under these conditions, the moisture concentration is a function of fiber radius r and time t only, and mass balance over an elemental isotropic volume yields a form of Fick's second law of diffusion:

$$\frac{\partial c}{\partial t} = \frac{1}{r} \frac{\partial}{\partial r} \left(r D \frac{\partial c}{\partial r} \right) \quad (33.15)$$

Coumans and Thijssen [32] derive a form of Fick's second law that can describe the drying of either solid or hollow cylinders exhibiting linear volumetric shrinkage. The space coordinate is taken as the relative radial distance at any time, while the diffusion coefficient is assumed to increase as a power law function of volumetric moisture content. This implies that the diffusion vanishes as the fiber approaches equilibrium.

Crank [33,34] provides numerous solutions to Fick's second law for a variety of boundary conditions. Some of the more important solutions follow.

33.3.1 STEADY-STATE BEHAVIOR

At steady state, when the diffusion coefficient is independent of position, Equation 33.15 reduces to

$$\frac{\partial}{\partial r} \left(r \frac{\partial c}{\partial r} \right) = 0 \quad (33.16)$$

which has the general solution

$$c = A + B \ln r \quad (33.17)$$

where A and B are constants to be determined from the boundary conditions.

One solution, which sometimes corresponds to an early period in the drying process, is the condition that the upper surface at $r = R_0$ is kept at a constant concentration c_0 and at the outer surface ($r = R$) evaporation takes place in the atmosphere for which there is an air-surface equilibrium concentration of c_e . A mass balance at this outer surface ($r = R$) gives

$$-D \frac{\partial c}{\partial r} = \beta(c - c_e) \quad (33.18)$$

where β is a mass transfer coefficient. This boundary condition leads to the solution

$$c = \frac{c_0[1 + \alpha \ln(R/r)] + \alpha c_e \ln(r/R_0)}{1 + \alpha \ln(R/R_0)} \quad (33.19)$$

in which $\alpha = \beta R/D$ and is a Sherwood number.

The loss of moisture per unit length of fiber is given by

$$\begin{aligned} W &= -(2\pi R) D \left(\frac{\partial c}{\partial r} \right)_R \\ &= 2\pi D \left(\frac{\alpha}{1 + \alpha \ln R/R_0} \right) (c_0 - c_e) \end{aligned} \quad (33.20)$$

which reduces to the expression

$$W = (2\pi R) \beta (c_0 - c_e) \quad (33.21)$$

for thin-walled fibers.

33.3.2 UNSTEADY-STATE BEHAVIOR

The expression

$$c = u \exp(-Da^2 t) \quad (33.22)$$

is a solution of Equation 33.15 for constant diffusivity provided u is a function of the radial dimension r only satisfying

$$\frac{d^2u}{dr^2} + \frac{1}{r} \frac{du}{dr} + a^2u = 0 \quad (33.23)$$

which is a Bessel's equation of order zero. Solutions of Equation 33.23 may be found in terms of Bessel functions, chosen to satisfy initial and boundary conditions.

Crank [33,34] has presented a number of such solutions. If the cylindrical fiber is initially at a uniform concentration c_0 , and there is a surface condition

$$-D \frac{\partial c}{\partial r} = \beta(c_s - c_e) \quad (33.24)$$

where c_s is the actual surface concentration at some time t and c_e is the final concentration at the surface at equilibrium, then the required solution is

$$\frac{c - c_0}{c_e - c_0} = 1 - \sum_{n=1}^{\infty} \frac{2LJ_0(ra_n/R)}{(a_n^2 + \alpha^2)J_0(a_n)} \exp(-a_n^2Dt/R^2) \quad (33.25)$$

In this expression, $J_0(x)$ is a zero-order Bessel function of the first kind, α is the nondimensional quantity $\beta R/D$, and the coefficients a_n are roots of the equation

$$aJ_1(a) - \alpha J_0(a) = 0 \quad (33.26)$$

in which $J_1(x)$ is a first-order Bessel function. The fractional reduction in moistness, M_t/M_∞ , compared with the amount lost after infinite time when the fiber reaches equilibrium is given by

$$\frac{M_t}{M_\infty} = 1 - \sum_{n=1}^{\infty} \frac{4\alpha^2 \exp(-a_n^2Dt/R^2)}{a_n^2(a_n^2 + \alpha^2)} \quad (33.27)$$

Newman [35] provides tabular values of M_t/M_∞ , which Crank [33] has plotted in graphical form, reproduced here as Figure 33.2.

Another useful solution of Bessel's equation occurs with the case of a cylindrical fiber initially at uniform concentration c_0 and subjected to the condition that there is a constant transfer rate at the surface (corresponding to constant drying rate conditions). This boundary condition may be written as

$$-D \frac{\partial c}{\partial r} = J_0 = cst \quad (33.28)$$

Macey [36] gives the solution at large times for the fractional reduction in moisture content as

$$\frac{M_t}{M_\infty} = 1 - \frac{A}{R} \left[2t + \frac{r^2}{2} - \frac{R^2}{4} \right] \quad (33.29)$$

where $A = \partial c / \partial r$ at the surface, $r = R$. Values of the concentration distribution given by Crank [33] are plotted in Figure 33.3 with the Fourier number Dt/R^2 as a parameter.

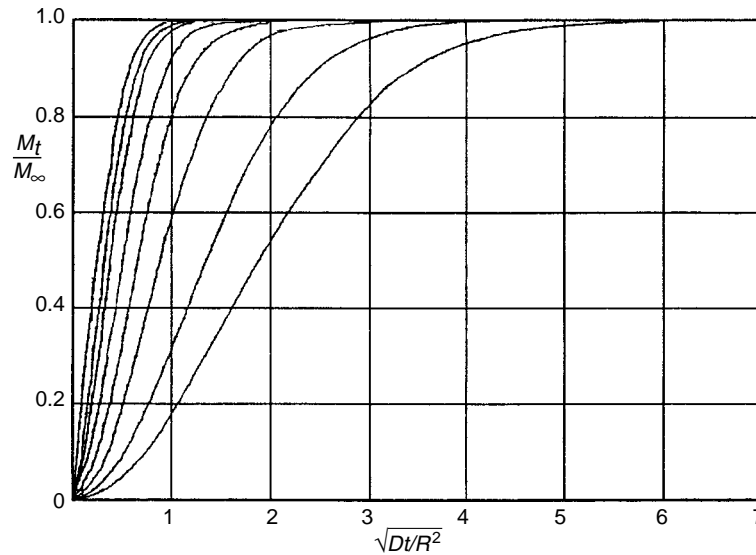


FIGURE 33.2 Fractional loss of moisture (M_t/M_∞) from a cylindrical fiber, as a function of $(Dt/R^2)^{1/2}$. The parameter is the Sherwood number $\alpha = \beta/D$. (From Crank, J. 1956. *The Mathematics of Diffusion*, Oxford University Press, Oxford and Newman, A.B. 1931. *Trans. AIChE*, 27:203–220.)

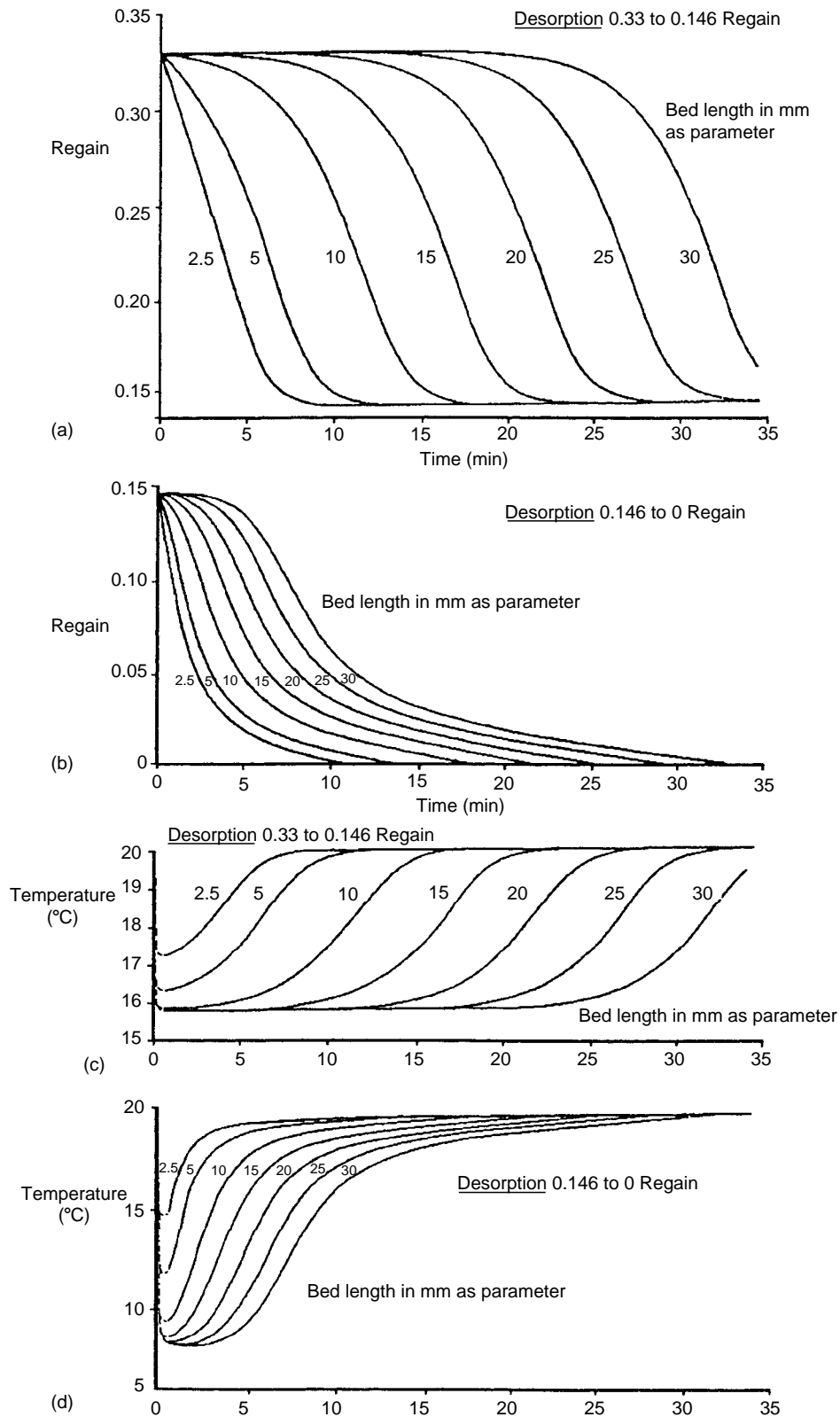


FIGURE 33.3 Response of a 30-mm thick wad of wool to environmental changes: (a) moisture content (regain) changes from 0.33 to 0.146 kg · kg⁻¹; (b) moisture content (regain) changes from 0.146 to 0 kg · kg⁻¹; (c) temperature changes on drying from 0.33 to 0.146 kg · kg⁻¹; and (d) temperature changes on drying from 0.146 to 0 kg · kg⁻¹. (From David, H.G., and Nordon, P. 1969. *Text. Res. J.*, 39(2):166-172.)

The moisture diffusion coefficient within a single wool fiber is of the order $10^{-11} \text{ m}^2 \cdot \text{s}^{-1}$ [34]. When there is no surface resistance, Figure 33.2 indicates that for fiber diameters, typically $20 \text{ }\mu\text{m}$ for wool, the fiber will respond to a change in moisture conditions very quickly, being about 98% complete within 25 s. Experimentally, King and Cassie [37] have measured the moisture uptake *in vacuo* by a sliver of Merino wool (having a fiber diameter of $21 \text{ }\mu\text{m}$). A temperature of 65°C was observed after 30 s but, by correcting the effect of the heat of adsorption, the authors conclude that the fiber would reach equilibrium in a time less than 15 s. Clearly, a mass of fibers will react rapidly to its surrounding environment, and local hygrothermal equilibrium may often pertain.

33.4 RESPONSE OF A FIBER MASS TO ENVIRONMENTAL CHANGES

Henry [38,39] first provided a theoretical framework to describe the response of a fiber mass to step changes in humidity or temperature in the surrounding environment. He assumed that the amount of moisture held by unit mass of fiber was a linear function of the water vapor concentration c_v in the air space between the fibers and the temperature T :

$$X = X_0 + \sigma c_v + \omega T \quad (33.30)$$

in which X_0 and the coefficients σ and ω are constants. These coefficients have the meaning of $(\partial X / \partial c_v)_T$ and $(\partial X / \partial T)_c$, respectively. Although Henry assumed constant values for these coefficients, they have been found to be strong functions of moisture content and temperature for both cotton [39] and wool [40]. The coefficient X_0 may be regarded as datum moisture content. Local hygrothermal equilibrium is assumed. The equations for the diffusion of moisture vapor and temperature through the mass in the longitudinal direction reduce to the pair of expressions

$$\begin{aligned} D \frac{\partial^2 c_v}{\partial x^2} - \frac{\partial}{\partial t} (c_v - \lambda T) &= 0 \\ \kappa \frac{\partial^2 T}{\partial x^2} - \frac{\partial}{\partial t} (T - \nu c_v) &= 0 \end{aligned} \quad (33.31)$$

in which

$$\lambda = -\gamma\omega / (1 + \gamma\omega) \quad \text{with} \quad \gamma = (1 - \varepsilon) / \varepsilon\rho_s \quad (33.32)$$

and

$$\nu = -\eta\sigma / (1 + \eta\omega) \quad \text{with} \quad \eta = \Delta H_W / C_F \quad (33.33)$$

It follows from the previous equations that $\gamma \ll 1$ and thus λ is also small: the concentration changes are scarcely influenced by temperature. Further, $\eta \gg 1$ and $|\nu| \rightarrow |\sigma/\omega|$, so the temperature changes depend upon the moisture concentration shifts.

As air is passed through a bed of moist hygroscopic fibers, the temperature fronts are seen to sweep through [41]. Associated with these temperature fronts are moisture concentration fronts, of which the second is the greater. Both fronts travel at widely different, but constant, velocities. The first front travels at a fraction of the air velocity, while the second front moves at a velocity of several orders of less magnitude. The slow second front is normally the one of principal concern in drying technology, being associated with majority of the moisture content change.

Nordon [42] models the passage of these fronts on the assumption that the moisture transfer in the fiber is very fast compared with the diffusion within the interfiber space. The moisture content and temperature profiles calculated by him for the drying of a thick wad of wool are presented in Figure 33.3a and Figure 33.3b for the exposure of fully saturated material to air of 65% relative humidity, and that of material equilibrated to perfectly dry air. The corresponding moisture contents are 0.33 kg/kg at $\varphi = 1$ and 0.146 kg/kg at $\varphi = 0.65$. Nordon's calculations apply to the case in which the wad is relatively extensive (a thick bed or a slow airflow rate) and would not be applicable to high-intensity drying of thin webs. In that case, the number of transfer units (in the airflow direction) is very small and the drying air no longer emerges saturated for most of the drying time, as in the example evaluated by Nordon.

For the purpose of calculating the propagation of the changes in humidity and the moisture content, the effects of the first, essentially thermal front can be neglected. The lowest temperature attainable in desorption is greater than the adiabatic saturation temperature, but only by a small amount. The drying of the bed as a whole can show a long constant-rate period that is only slightly smaller than the constant-rate period when the fibrous material containing free water is dried. This behavior is the characteristic of drying webs with large extensiveness (large number of transfer units).

Nordon's predictions have been subsequently confirmed in tests on changes in moisture content and temperature in response to hygrothermal changes in the surrounding air, and in the changes that accompany the Hoffmann pressing and heat through fabrics during changes in moisture content [43].

In another work, Nordon and David [44] have modified their analysis to take account of the two-stage sorption behavior of a textile material and moistness dependence of the transfer rate. Use of

this elaboration does not alter the essential conclusions of Henry's linearized analysis. A wad of textile material exposed to a sudden rise in ambient relative humidity experiences the passage of the fast temperature front with a rise in temperature, and a slow front with a drop in temperature of equal magnitude. Of the two corresponding humidity fronts, only the slow one is obvious as the fast one is very small and superimposed on the major one.

Henry's linearized analysis, with averaged values for the coefficients σ and φ , has been used by Walker [99] to estimate changes in temperature and weight of wool bales, both open and enclosed, when exposed to an environmental change in temperature and humidity. The method is able to follow the response in core temperature and bale moisture to the new equilibrium level. Walker also used the technique to determine the equilibration of overdried fleece wool in an undisturbed bin. After 1 day, the moisture uptake is calculated to penetrate 150 mm into the fibrous mass, reaching 700 mm after 10 days and 1.1 m after 100 days.

More recent work by Cudmore et al. [45] highlights difficulties in using Henry's methods in analyzing the behavior of modern high-density bales of scoured loose wool in which there may be pockets of relatively wet material, and significant unbound moisture transport may prevail. When fiber is packed at high density, its equilibrium moisture content and the apparent moisture diffusion coefficient appear to fall, and data obtained with materials packed loosely may no longer apply. Moisture diffusion in the bale approaches that in a single fiber. Under these conditions, Cudmore et al. describe the moisture redistribution in terms of Fickian movement between slabs at different moisture content, employing diffusion coefficients that are a function of temperature and packing density [46]. Temperature changes were experimentally found to be insignificant.

The influence of the openness of a fiber matrix on the effective moisture diffusion is also observed in the diffusion of water vapor through wood pulp and paper sheets.

Nilsson et al. [47] find that the apparent moisture diffusion coefficient falls from 5.4×10^{-6} to $2.1 \times 10^{-6} \text{ ms}^2 \cdot \text{s}^{-1}$ as the density increases from $500 \text{ kg} \cdot \text{m}^{-3}$ (softwood pulp) to $1530 \text{ kg} \cdot \text{m}^{-3}$ (coated and calendered paper).

33.5 CONVECTIVE DRYING OF FIBROUS MASSES

Moisture migration in fibrous and porous media can occur in a number of ways: (a) by liquid diffusion along the fibers due to moisture and temperature

gradients; (b) movement due to capillarity and gravity within interfiber spaces; and (c) vapor diffusion due to variations in moisture vapor pressure throughout the mass. Knudsen flow or effusion exists when the mean free path of the vapor molecules is of similar dimensions to the space between the fibers. This is unlikely under most commercial drying situations. Surface diffusion of sorbed moisture may also occur, but such movement may not significantly influence the transport of moisture as the migrating material may simply recirculate around a single air-filled pocket [48]. Transport of sorbed moisture through fibers, however, does appear to take place [49].

It is normally assumed that large fiber masses, fiberboards, and webs are macroscopically homogeneous so that it is possible to apply conservation and constitutive equations over sufficiently small control volumes to obtain smooth profiles of temperature and moisture, as implied in the work described in Section 33.4. Should no detailed information on these properties be needed, it is possible to fit a simple diffusional equation to the drying process, often with a concentration-dependent diffusion coefficient [50]. In the hygroscopic moisture region, tests with glass fiberboards suggest that temperature gradients can make a significant contribution to the total moisture transfer, with a thermal gradient coefficient in the order of $2 \times 10^{-9} \text{ kg} \cdot \text{kg}^{-1} \cdot \text{K}^{-1}$ [51].

The relative success of the concept of the characteristic drying curve has led to the investigation of whether the concept describes the drying of beds of loose fibers. This concept derives from van Meel's idea [52] that for a given material the rate of drying, relative to the value when only the external boundary layer controls the process, is only a function of the volume-averaged moistness, expressed as the relative free moisture content. In other words, the drying kinetics may be described by a function of the kind

$$f = f(\Phi) \quad (33.34)$$

where

$f = N_v/N_w$ and $\Phi = (X - X_e)/(X_{cr} - X_e)$, subject to the boundary conditions

$f = 1, \Phi = 1$ at the critical moisture content

$f = 0, \Phi = 0$ at equilibrium with

$f = 1$ for $\Phi \geq 1$ (unhindered drying period), and thus

$0 \leq f \leq 1$ over the range, $0 \leq \Phi \leq 1$ (falling rate period).

With some hygroscopic fibers, however, an initial, constant drying rate period is uncertain, giving rise to doubts about the estimation of an appropriate critical moisture content. Walker's data [40], for drying wool in cans under constant external conditions, show a

maximum drying rate, but no period in which the drying rates remained constant. Moreover, a mass of loose fibers do not constitute a simple capillary-porous body, and the mechanisms that give rise to a constant-rate period in a porous network [53] do not arise in a touching fibrous assembly. Recent experiments [54] on the transport of water and *n*-propanol through samples of Kraft pulp and filter paper confirm lack of importance of capillarity. Even with knitted fabrics, drying has been envisaged in terms of evaporation at all depths in the material and vapor diffusion there from [55].

If there is no critical point, Langrish et al. [56] suggest that it might be possible to derive normalized drying rate curves based on the initial rather than the critical moisture content. A functional relationship

$$g = g(\Gamma) \quad (33.35)$$

where $g = f/f_0$ and $\Gamma = \Phi/\Phi_0$ will be found if the original characteristic drying curve takes the simple algebraic form,

$$f = A\Phi^n \quad (33.36)$$

in which A and n are coefficients. It is also sometimes possible to estimate an apparent critical point from data obtained wholly in the falling rate period in drying thin layers of material [12].

Tubbs [57] finds a characteristic drying curve for loose wool hung in minibales in an airstream over a limited range of humidity potential and air velocity. Later work by Keey and Wu [58] on through-circulating thin layers of saturated wool over a temperature range from 60 to 85°C, indicates that the concepts hold approximately for these conditions, with $n = 0.6$. The fractional dryness of the surface will be a limiting factor in the cross-circulation drying of thin veneers and webs, as noted in the experiments of Peck et al. [59] who find $n = 2/3$ in the drying of thin slats of balsa wood. The internal resistance to moisture movement is essentially negligible with porous materials less than 6 mm in thickness.

A single characteristic drying curve is unlikely to be found in the drying of bulk fibrous material, such as timber boards. However, Keey and Pang [60] note that the high-temperature drying of softwood boards of a given thickness can be described by two common curves for commercial kiln ranges in temperature and humidity. One curve relates to the period when an evaporative front at the boiling point sweeps through the material; the other to the period when cell wall and bound water diffusion takes place below the fiber saturation point.

With bulky stuff, such as bound bobbins of cloth and yarn, capillary transport of moisture is possible.

Nissan et al. [61] have investigated the drying of a 127-mm wide piece of Terylene cloth (a polyester fiber) wound on a spool to give about 25-mm depth of material and the drying of a similar wound bobbin of wool composed of bulked yarn in heavy and felted cloth. A critical moisture content at $0.73 \text{ kg} \cdot \text{kg}^{-1}$ was found with the Terylene bobbin, but for the wool the value was $2.96 \text{ kg} \cdot \text{kg}^{-1}$. A feature of these and other tests reported by Bell and Nissan [62] was the appearance of a pseudo wet bulb temperature, a quasisteady temperature determined by the thermal balance between the inward transfer of heat and the outward evaporation of moisture from the body of the winding, as illustrated in Figure 33.4. Such results suggest that the drying can be modeled in terms of an evaporative plane receding from the exposed surface.

Gummel [63,64] has examined the through-circulation drying of textile and paper webs. These are regarded as regular porous networks, with the threads composed of multiple strands of individual fibers (Figure 33.5). Some values of characteristic dimensions reported by Gummel are given in Table 33.2. The textiles were through-circulated in the range of temperatures from 20 to 70°C, and the paper tissue from 20 to 90°C at air velocities between 0.06 and $1.5 \text{ m} \cdot \text{s}^{-1}$. The results for textiles could be expressed in the form of normalized drying curves, with a critical point of $0.41 \text{ kg} \cdot \text{kg}^{-1}$ for the textile fabrics. The normalization was less convincing for the data involving paper tissue, with highly variable critical moisture contents in the range of 2.6 to $4.0 \text{ kg} \cdot \text{kg}^{-1}$ being recorded.

In a later work, Albrecht [65] showed that characteristic drying curves could be drawn up for the through-circulation drying of cottonlike fabrics over a range of incident velocities from 0.1 to $0.6 \text{ m} \cdot \text{s}^{-1}$. The critical moisture content was defined by the intersection of the constant rate and linearized falling rate curves. One has

$$-\frac{dX}{dt} = KX = \frac{N_V}{\rho_s/a} \quad (33.37)$$

where K is an empirical drying coefficient. It follows that the critical moisture constant becomes

$$X_{\text{cr}} = \frac{N_W a}{\rho_s K} \quad (33.38)$$

with

$$K = -\frac{d}{dt}(\ln X) \quad (33.39)$$

and N_W is the maximum unhindered drying rate.

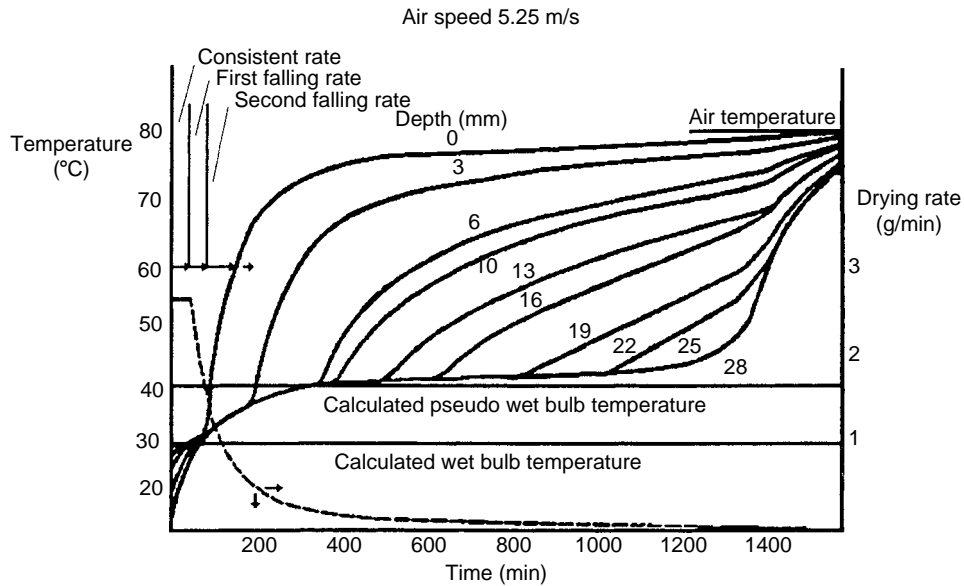


FIGURE 33.4 Temperature profiles on drying a 28-mm diameter bobbin of wool at a dry bulb temperature of 80°C, a wet bulb temperature of 30°C, and an air velocity of 5.5 m · s⁻¹. (From Bell, J.R. and Nissan, A.H. 1959. *AIChE J.*, 5:344–347.)

More advanced models of the drying of fibrous webs take into account the interaction of water and the solid matrixes [66]. For example, dry paper fiber consists of about 100 lamellae, but is not porous [67]. Water can diffuse into the fiber and dissolve within the cellulose and hemicellulose, causing swelling across the fiber but not along it. Completely wet paper consists of moisture between fibers and bound water within the solid matrix.

Only when the moisture content reaches the fiber saturation point does the intrafibrous moisture vanish and the paper begins to shrink.

Pores within the fibrous mass may not lead to the exposed surface, but may have dead ends or simply be occluded. Methods of estimating capillary motion in such porous structures are considered by Neiss and Winter [68].

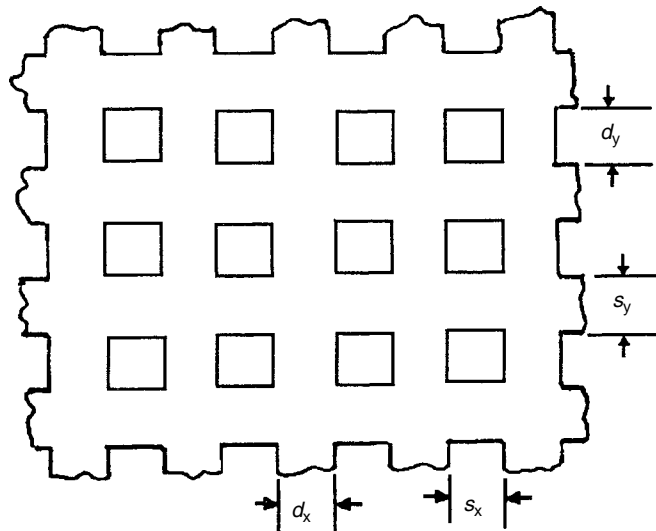


FIGURE 33.5 Model of a textile web (d = yarn diameter; and s = thread spacing). (From Gummel, P. 1977. *Durchströmungsrocknung. Experimentelle Bestimmung und Analyse der Trocknungsgeschwindigkeit und des Druckverlustes luftdurchströmster Textilien und Papiere*, Dr. Ing. thesis, University Karlsruhe, TH.)

TABLE 33.2
Characteristic Dimensions of Some Textile and Paper Webs

Material	<i>i</i>	Number Threads (mm ⁻¹)	Thread Diameter (μm)	Pore Width (μm)	Thickness (μm)	Weight (g · m ⁻²)
Polyester PES626	<i>x</i>	2.68	324	49	424	194.6
	<i>y</i>	1.97	324	184		
Polyester PES611	<i>x</i>	2.80	300	57	439	192.8
	<i>y</i>	1.98	300	205		
Wool fabric	<i>x</i>	2.15	326	130	644	204.9
	<i>y</i>	1.41	352	357		
Acrylic-wool fabric	<i>x</i>	2.11	323	151	590	203.3
	<i>y</i>	1.28	448	332		
Paper tissue						
Light				200	257	50.0
Heavy				150	258	23.1

Source: From Pander, J.R. and Ahrens, F.W. 1987. *Drying Technol.*, 5(2):213–243.

Besides any capillary transport of liquid moisture and vapor diffusion, there is sorptive or bound moisture diffusion within the fibers when they are less than fully saturated [49]. At relative humidities below 0.8, the sorptive transport coefficient diminishes exponentially with decreasing values of relative humidity.

The use of such a detailed mechanistic picture of the pore structure in modeling the convective drying of paper is described by Harrman and Schultz [69]. The alternative approach is to consider the capillary-porous web as a continuum having equivalent thermodynamic properties to the pore system [70]. This approach has recently been described in detail by Lampinen and Ojala [71].

33.6 THROUGH-CIRCULATION OF LOOSE FIBERS AND WEBS

Kröll [72] reviews the kinds of commercial through-circulation dryers that can treat loose fibers and fibrous webs, and Watzl [73] presents a more recent overview of these kinds of dryers with particular

reference to energy conservation and environmental protection.

Originally, cloth was stretched out on a wooden framework or tenter in the open air to dry, the edges being firmly held by hooks. To be on “tenterhooks” has entered the English language as a figurative expression for being held in suspense under tension. Today the tenteryards have been replaced by enclosed dryers through which the cloth is moved over pegs. Loose fibers can be conveyed by perforated bands through which the drying air is circulated. Most modern drying systems, however, incorporate rotating perforated drums that take up less floor space than horizontal band dryers. Single-drum dryers can be adapted for the drying of paper tissue and carpet lengths, while multiple-drum units can handle loose fibers such as fleece wool. Table 33.3 gives an indicative comparison of the major dryer types. Cylinder machines offer advantages in space needs and thermal economy.

The evaporative capacities in Table 33.3 are somewhat higher than the corresponding values given by Stewart [74] for fleece wool dryers, namely, 10 to 15 kg · m⁻² h⁻¹ for belt (or brattice) dryers and 20

TABLE 33.3
Comparison of Equipment to Dry Loose Fibers and Webs

Type	Specific Evaporative Energy Use (kJ · kg ⁻¹)	Evaporation Rate (kg · m ⁻² h ⁻¹)	Relative Floor Space
Flat tenter	4610	30	1
Perforated band	3600	30	0.85
Drum tenter	3440	44.8	0.4
Sieve drum	3250	44.8	0.4

Source: From Watzl, A. 1991. *Melliand Textilberichte*, 72(6):470–479.

TABLE 33.4
Average Drying Rates of Selected Materials in Through-Circulated Perforated-Drum Dryers

Material	Air Temperature (°C)	Moisture Content Dry Basis (%)		Mean Drying Rate ($\text{kg} \cdot \text{m}^{-2} \text{h}^{-1}$)
		Inlet	Outlet	
Wool, spun	80	50	20	13.0
Wool, squeezed	80	60	20	15.8
Cotton, spun	90–100	60	8	18.2
Cotton, squeezed	90–100	100	8	24.2
Rayon, spun	110	90	11	21.9
Rayon, squeezed	110	180	11	25.2
Sisal	120	80	12	25.8
Jute	110	90	15	18.5

Source: From Kröll, K. 1978. *Trockner und Trocknungsverfahren*, 2nd ed., Springer-Verlag, Berlin.

to $30 \text{ kg} \cdot \text{m}^{-2} \text{h}^{-1}$ for suction-drum dryers. However, Stewart's figures probably relate to older dryers of lesser performance. The electrical power use for brattice dryers is less than that for units involving rotating drums, being about $250 \text{ kJ} \cdot \text{kg}^{-1}$ (evaporation) for the band driers and $400 \text{ kJ} \cdot \text{kg}^{-1}$ for the drum dryers. However, the latter is still only about 12% of the thermal load.

These data are consistent with guaranteed operative capacities (presumably minimum values) of one manufacturer reported by Kröll [72] for suction-drum dryers and are reproduced in Table 33.4.

Gardiner and Dietl [75] describe the single-drum dryers for thin permeable paper sheets in which drying rates in the range of 50 to $200 \text{ kg} \cdot \text{m}^{-2} \text{h}^{-1}$ were measured with inlet air temperatures of 170 to 450°C . The drying is accomplished in a matter of seconds. There is sufficient pressure drop across the perforated cylinder

and pressing sheet to hold it to the drum, as the sheet is pulled through, without leaving marks. The principle of the process is illustrated in Figure 33.6.

Multiple-drum dryers for handling loose stuff may be arranged with rotating cylinders in a horizontal or vertical array, although the former is more common. The vertical arrangement has advantages if the material has to be shifted from one level to another. A comparison in layout between brattice and suction-drum dryers is shown in Figure 33.7.

Up to 20 drums may be used in series, although commonly about 5 would be employed. Normally, drums are supplied in the range of 1.4 to 2.0 m diameter, with widths up to 6.0 m . The principal operational problem with these dryers for loose fibers relates to the difficulty of securing a feed of uniform openness and thickness. Scoured fleece wool, for example, after passing through the final squeeze roll at

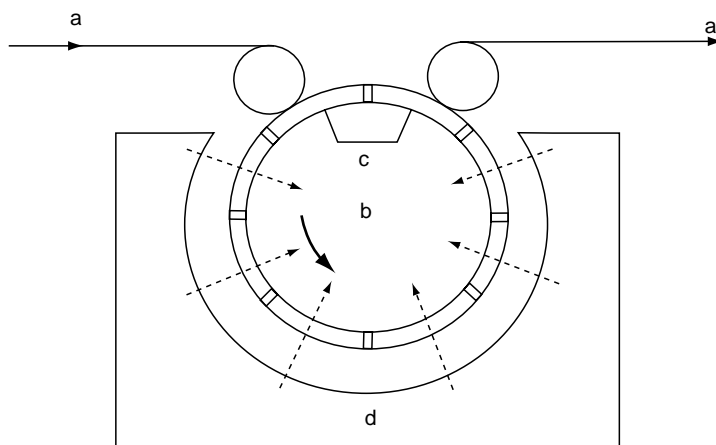


FIGURE 33.6 Perforated cylinder for drying thin paper sheets: (a) paper sheet; (b) sieve drum; (c) exhaust; and (d) hot air distributor. (From Kröll, K. 1978. *Trockner und Trocknungsverfahren*, 2nd ed., Springer, Berlin.)

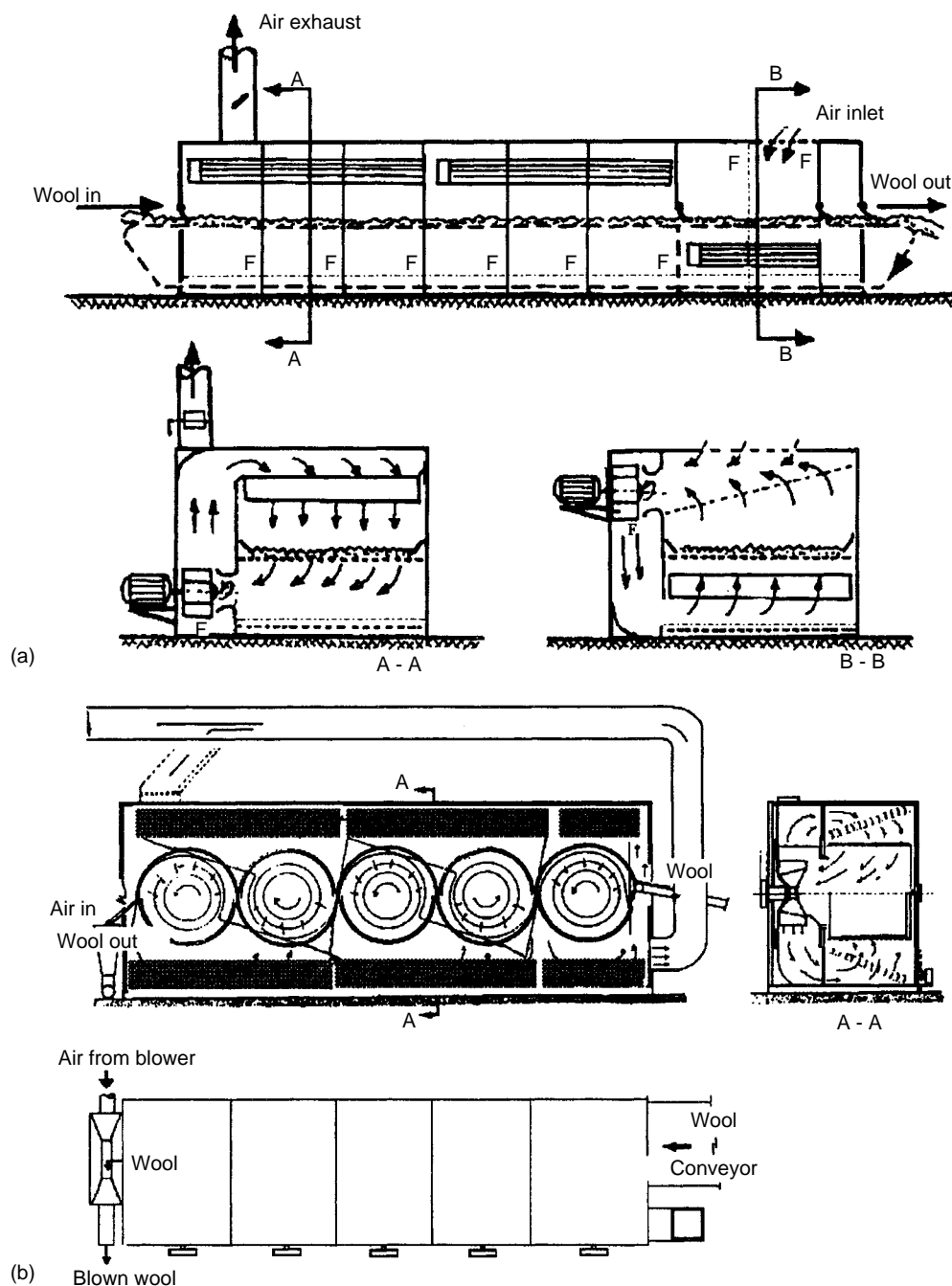


FIGURE 33.7 Dryers for loose wool: (a) brattice (band) dryer; and (b) suction-drum dryer. (From Stewart, R.G. 1983. *Woolscouring and Allied Technology*, Wool Research Organization New Zealand, Christchurch, New Zealand.)

the outlet of the scouring bowl, is then combed out into a spiked feeding conveyor to present to the dryer a fairly uniform, tangled mat of fibers.

Band dryers and tenters lose significant amounts of heat through the cooling of the band or chain as it passes out of the hot zone and through the extensive chamber wall area. Drum dryers are thermally more efficient, as Table 33.3 indicates. Conventionally, these dryers are fitted with internal steam, finned heaters,

which are prone to collect fluff and loose fiber. Direct firing of the inlet air has the advantage of avoiding such heating elements that need periodic cleaning and may be the site of incentive pockets of stuff.

If the standard drum-drying system does not provide adequate capacity, the newly developed, high-capacity drum with a fractional free area of 96% may be specified as a possible alternative unit having evaporative capacities up to $400 \text{ kg} \cdot \text{m}^{-2} \cdot \text{h}^{-1}$ [76].

33.7 VARIATION OF PROCESS CONDITIONS AND MOISTURE IN THROUGH-CIRCULATED DRYERS

A detailed analysis of the through-circulation of loose materials is given by Keey [17]. The analysis is also valid for webs of fibrous materials. Consider an elemental volume within fibers on a perforated band as shown in Figure 33.8.

A moisture balance between that lost by the fibers and that gained by the drying air yields

$$G \frac{\partial Y_G}{\partial y} = -\rho_B(1 - \varepsilon)u_S \frac{\partial X}{\partial z} \quad (33.40)$$

where the symbols have the meanings illustrated in Figure 33.8. The movement of solids may be directly related to the band speed if the fibers do not redistribute on the band during drying:

$$t = z/u_S = z/u_B \quad (33.41)$$

The longitudinal moisture content gradient is found by considering the rate of drying of the fibers within an incremental volume in the layer:

$$-\rho_S(1 - \varepsilon)u_S \frac{\partial X}{\partial z} = f\beta\phi_M a(Y_W - Y_G) \quad (33.42)$$

where β is the mass transfer coefficient, ϕ_M is the humidity potential coefficient, and $(Y_W - Y_G)$ is the humidity difference between the wet bulb and the bulk gas values.

Equation 33.40 and Equation 33.42 can be conveniently recast into dimensionless form

$$\frac{\partial \Phi}{\partial \theta} = \frac{\partial \Pi}{\partial \xi} \quad (33.43)$$

and

$$-\frac{\partial \Phi}{\partial \theta} = f\Pi \quad (33.44)$$

in which one has a characteristic moisture content

$$\Phi = \frac{X - X_e}{X_{cr} - X_e} \quad (33.45a)$$

a relative time of drying

$$\theta = \frac{N_{W0}A}{F(X_{cr} - X_e)}t \quad (33.45b)$$

a relative humidity potential

$$\Pi = \frac{Y_W - Y_G}{Y_W - Y_{G0}} \quad (33.45c)$$

the extensiveness of the fiber mass

$$\xi = \beta\phi_M a y / G \quad (33.45d)$$

In the foregoing definitions, subscript cr refers to the transition from unhindered to hindered drying in the falling rate period, e to equilibrium, 0 to the air inlet face of the fiber mass, G to the bulk air, and W to the wet bulb conditions. The holdup of solids on the band is F.

The solution of Equation 33.43 and Equation 33.44 for a first-order drying process (linear falling rate period) is given in detail by Keey [17]. The drying can be divided into three stages:

1. The moisture contents everywhere are above the critical point. The drying rates fall off in the airflow direction, resulting in a moisture content profile normal to the band.
2. Part of the fibrous mass, adjacent to the band, is dried below the critical moisture content. Some enhancement of drying rates within this occurs, changing the shape of the moisture content profile somewhat.
3. All of the materials are below the critical moisture content, and the moisture content differences normal to the band gradually diminish, although the relative difference $\Delta\Phi/\langle\Phi\rangle t$, where $\langle\Phi\rangle$ is some average value, does not.

The variation of the characteristic moisture content Φ and the drying rate f as a function of relative time θ is shown in Figure 33.9 for the case when $\Phi_0 = 2$ and $\xi_{\max} = 1$.

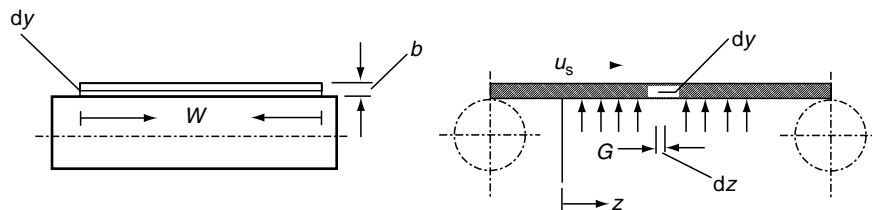


FIGURE 33.8 A perforated band dryer.

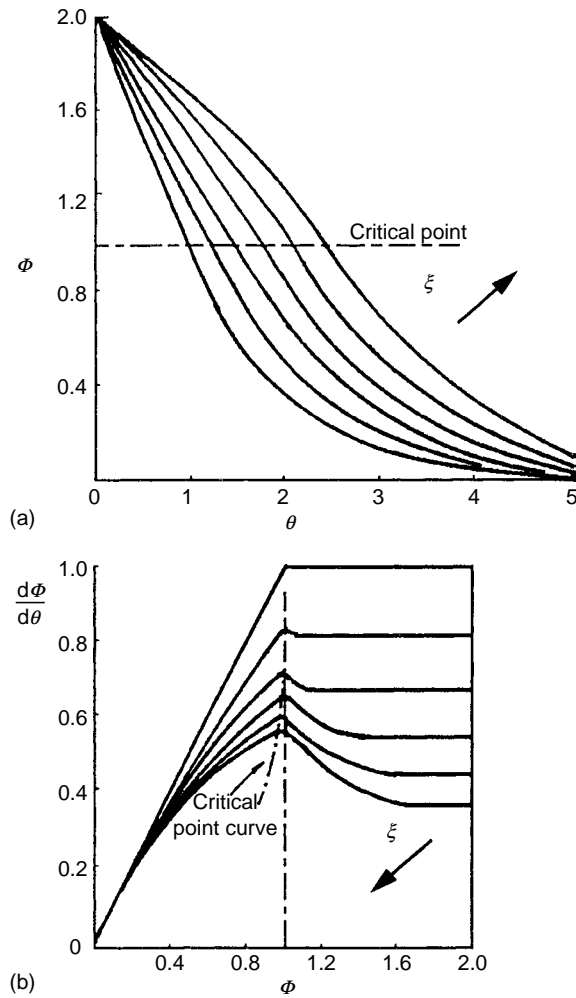


FIGURE 33.9 Through-circulation of a fibrous mass on a band for an initial moisture content $\Phi_0 = 2$ and a mat thickness $\xi_{\max} = 1$: (a) variation of moisture content Φ with relative time (or distance) θ along the band; (b) variation of drying rate $d\Phi/d\theta$ with relative time (or distance) θ . (From Keey, R.B. 1992. *The Drying of Loose and Particulate Materials*, Hemisphere, Washington, DC.)

The moisture content variations that arise at right angles to the direction of the band movement can lead operators to overdry stuff so that the absolute residual moisture content differences are acceptable. The relative moisture content differences are never reduced. However, by periodically changing the airflow direction along the band, the streamwise variation in moisture content can be restricted to a considerable degree. Suction-drum dryers automatically incorporate this feature, and suitable baffling and placing of fans can also affect airflow reversal in brattice (horizontal band) dryers, as illustrated in Figure 33.10.

The effect of a single airflow reversal at $\theta = 1$ for the case illustrated in Figure 33.9 is depicted in

Figure 33.11. At a contact time of $\theta = 2$ the maximum moisture content difference $\Delta\Phi$ has been reduced to one fifth the value that is found without any reversing of the airflow, with the wettest material now being in the middle of the mass and the outer portion of the drier. In the suction-drum dryer, the radial moisture content profiles are essentially negligible after the fourth drum (three airflow reversals) [58].

It is difficult to ensure even feeding of loose fibers to a dryer as combing and other prefeeding techniques are only partially effective in untangling fibers that have interlocked. The effect of such unevenness in mat thickness and/or openness has been investigated by Keey and Wu [77], who consider that an actual suction-drum dryer can conceptually be replaced by an equivalent unit consisting of a number of narrow dryers in parallel, each being fed uniformly with material. Experimental transverse profiles of thickness can be discretized so that the individual feeding rates are determinable. Gas dynamic considerations yield the corresponding gas flow through each of the narrow dryers to maintain across the dryer a uniform pressure drop that is known. The resulting nonuniformity in the extent of the drying implies that the material would be underdried unless additional heating were supplied or the capacity reduced or both. Experimental observations suggest that the relative thickness on the band is given by

$$\delta = \frac{y - y_0}{s - y_0} = \exp(-k/w) \quad (33.46)$$

over the region $P_0 < P(w) < 1$, where w is essentially the relative effective width of the dryer and is evaluated as

$$w = \frac{P(w) - P_0}{1 - P_0} \quad (33.47)$$

with $P(w)$ being the probability that the relative depth is δ or less. The coefficient k has been called the “unevenness factor,” and is a measure of the variation in thickness. A representative value of k for one suction-drum dryer handling loose wool is 0.2 [78].

33.8 AIR IMPINGEMENT OF FIBROUS MATERIALS

The drying of heavy fabrics, including broadloom carpets, can be assisted by the use of air jets. The design of air-impingement systems to improve the efficiency of drying textile fabrics is reviewed by Gottschalk [79] with reference to one commercial arrangement. The

of paper or textiles, impingement drying rates can be enhanced further by drawing some of the hot drying gas through the web.

Martin [86] provides design correlations for multiple-slot and round jets besides recommendations for the spatial arrangement of jet nozzles on the basis of maximizing heat transfer per unit fan energy. The optimal ratio of the pitch of the nozzles compared with the distance above the surface (0.7), which is recommended by Martin, is close to the reported critical value at which jet-to-jet interactions start influencing the heat transfer at the stagnation point under the jet axis [83].

33.9 DRYING OF PULP AND PAPER

All paper and wood-based boards are made from a suspension of fibers in water, with relatively small quantities of nonfibrous additives and fillers to give particular finishes to the dried product. Most of the water is drained away in a combination of rolls, foils, and suction boxes. In papermaking, the pulp is admitted to a wire section through a flow box, from which it leaves having a water-to-fiber ratio of about 6:1. Hence, the pulp passes to a press section in which the water is squeezed into endless belts during traverse between loaded nips, so removing about 70% of the water. A diagrammatic layout is shown in Figure 33.12.

The web then passes to the drying section, from which the sheet emerges at about 8% moisture content (dry basis), essentially in equilibrium with the ambient atmosphere. Normally, the drying section consists of a series of rotating cylinders, internally heated by steam, grouped in double banks of about 10 units each. The wet sheet passes alternately between the upper and lower ranks of cylinders in an endless web. Contact is assisted by pressing dry felts, which also retain shrinkage and hinder wrinkling of the

sheet (Figure 33.13). The sheet passes through the machine at high speed, the peripheral velocity of the rotating cylinders reaching $10 \text{ m} \cdot \text{s}^{-1}$ or higher [87], to be taken up by an end reel when dried.

Semipermeable tissue can be dried by through-circulation over a single cylinder [84,87]. Much lightweight paper is dried in this way. The very large steam-heated cylinder, up to 6 m in diameter, is run at extremely high speeds (more than $20 \text{ m} \cdot \text{s}^{-1}$), and may be worked with a small number of fore and aft cylinders. Gardiner and Dietl [75] give details of the performance of these single-cylinder dryers.

Nissan [88] has divided the drying over each drum of the multiple-cylinder drying section into a cycle of four phases:

1. The paper sheet contacts the outer surface of the cylinder, but is uncovered by the felt.
2. The sheet is pressed onto the cylinder's surface by the felt.
3. The sheet remains in contact with the cylinder, but the felt has now left.
4. The sheet, no longer in contact with the cylinder, traverses freely to the next drum in the adjacent bank.

To obtain the moisture profile in the machine direction, Nissan et al. [89] have made several assumptions about the drying process, principally that the drying was a first-order process (linear falling rate period) and that the pressing felt in phase 2 of the drying cycle reduced the evaporation rate to one tenth that in the sheet's free traverse between cylinders under similar temperature driving forces. Over each of the periods when the sheet touches the cylinder, it is assumed that the sheet temperature is constant (or linearly varying about an arithmetic mean), both in the plane of the sheet and normal to it. These concepts lead to a relatively

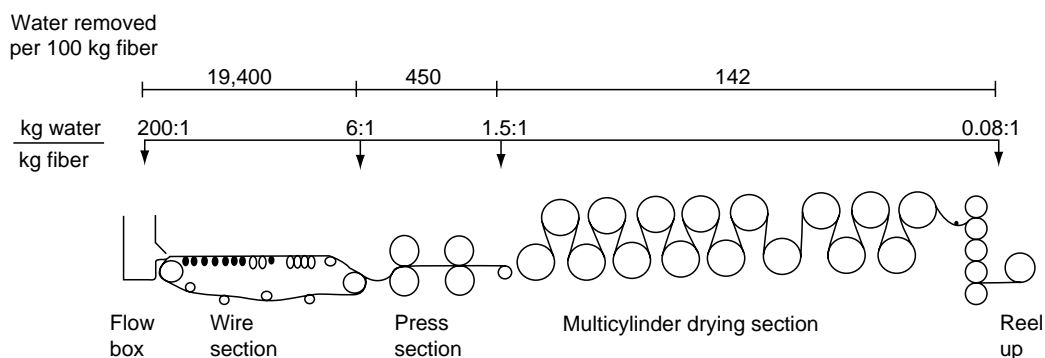


FIGURE 33.12 Layout of a papermaking machine. (From Kirk, L.A. 1984. *Advances in Drying*, 3:1–37.)

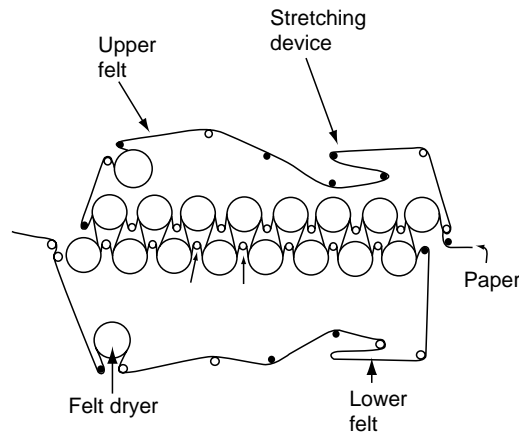


FIGURE 33.13 Multicylinder drying section showing felt runs. (From Kirk, L.A. 1984. *Advances in Drying*, 3:1–37.)

simple relationship for the sheet temperature after a time interval t :

$$T_{*t} = T_S - (T_S - T_0) \exp[U\Delta/M_S(C_S + C_L X)] \quad (33.48)$$

where T_0 is the initial sheet temperature at the beginning of the considered phase, T_S is the steam temperature, U is the overall heat transfer coefficient, M_S is the amount of dry solids per unit sheet area, C_S is the specific heat of dry fiber, C_L is the specific heat of moisture, and X is the dry basis moisture content. Over phase 4, the temperature changes are obtained by evaluating the sum of convective, evaporative, and radiative heat losses.

These ideas led to the prediction of a saw-toothed temperature profile, tracking the successive heating and cooling of the sheet as it passes over and under the trains of cylinders. Most of the moisture loss is predicted to take place during the free traverse (phase 4), but a significant fraction occurs over phase 3 when the felt has left the cylinder (Figure 33.14).

However, later experiments have indicated that a considerable fraction of water is removed in phase 2 [90], probably as liquid into the felt. Subsequent analogue computations by Deploy [91] suggest that the felt does indeed hinder evaporation, with the wet web leaving the cylinder at a higher temperature than when the cylinder was unfelted. The net effect is that a felted cylinder has slightly less moisture loss than an unfelted one for the same initial conditions.

Bell et al. [90] explain the differences in the relative importance of each phase of the cycle in terms of the difference in the felting action between various ma-

chines. Nissan [92], in considering the way the felt removes water in a papermaking machine, attributes such differences to the composition of the felt and the extent to which the felt is heated and dried between cylinders.

More advanced models of the drying process are discussed by Kirk [87]. More recent reviews include those written by Nederveen et al. [93] and Wilhelmsson et al. [94]. The latter authors identify 20 models that have been proposed to simulate multicylinder paper dryers. In some of these models, the physical transfer processes are simulated in detail, while others adopt a less detailed viewpoint and rely on more empirical coefficients to fit the data.

In practice, the majority of the evaporation (about 80%) occurs in the draws, with a disproportionately large number of cylinders needed for the relatively small amount of evaporation in the falling rate period. This feature is a general observance for progressive drying, as illustrated by the moisture profiles in Figure 33.9. Wilhelmsson et al. [94] note that about two thirds of the heat demand in a multicylinder machine is used directly in evaporation, while about 22% is lost from the cylinders to the environment in various ways. About 6% is lost from the web by convective heat exchange with the air.

Ventilation of the sheet as it passes through the drying section is important as an inadequate air supply can result in the conversion of a uniform cross-web moisture profile into one with a markedly wetter mid-section at the take-up reel. The efficiency of various pocket ventilation systems is investigated by Kirk [95].

Comparative values of drying rates and steam use with multicylinder dryers for paper are given in Table 33.5. The data probably do not reflect modern practice, with its higher machine speeds and better

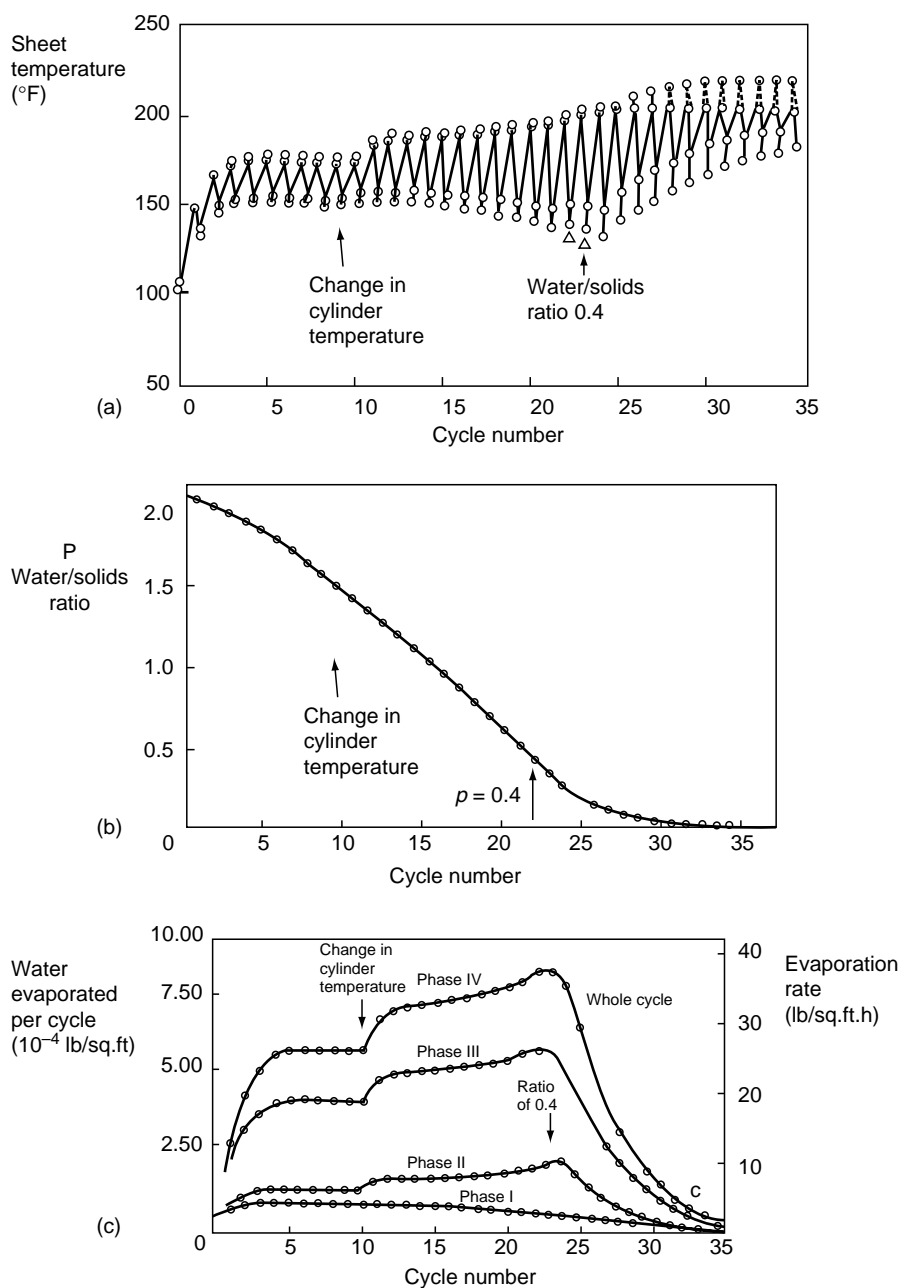


FIGURE 33.14 Calculated variation in sheet properties in the drying section of a papermaking machine: (a) temperature profile, $^{\circ}\text{C} = (^{\circ}\text{F} - 32) \times 5/9$; (b) water/solids ratio; (c) water loss rate, $\text{kg} \cdot \text{m}^{-2} = 4.882 \text{ lb} \cdot \text{ft}^{-2}$. (From Nissan, A.H., Kaye, W.G., and Pilling, D.E. 1958. *Trans. IChemE*, 36:107–114.)

energy savings, but merely give some indication of the relative performance of these kinds of drying units with various grades of paper.

Hardboard, which is produced from coarse thermomechanical pulp, and similar dense products are dried batchwise by pressing. The fiber mat is fed to a press, where the mass is compressed over a heated platen. The main effect of press drying is the thermal

softening of lignin and cellulose components. If temperatures 25 to 50°C above the glass-transition point of lignin is reached, then bonding can take place. This condition entails a temperature in the order of 150°C or above in the wet web. With a further increase in temperature, the lignin viscosity slowly declines and lesser pressures are needed to attain a given board density [96].

TABLE 33.5
Drying Rates and Steam Use in Multicylinder Dryers for Paper

Material	Basis Weight (g · m ⁻²)	Mean Drying Rate (kg · m ⁻² h ⁻¹) ^a	Energy Use (kg steam/kg dry s)	No. of Sections	No. of Cylinders/Section
Condenser paper	8–12	3–5	5–7	1	1
Vellum	40–60	5–7	4–6	1	4–6
Writing paper	50–70	16–28	1.9–2.5	3	8–10
Newsprint	50–52	20–23	1.8–2.2	3	8–16
Packaging	60–120	20–24	2–3	3	8–10
Carton	200–700	14–20	2.2–3.0	3 ^b	6–8

^aCalculated in terms of the web-covered surface (about 60% of the total surface).

^bUpper and lower cylinders separated.

Source: From Meinecke, A. 1974. *Wochenblatt f. Papierfabr.*, (102):41–52.

33.10 SUPERHEATED-STEAM DRYING

Superheated-steam drying is a specialized technique that uses superheated steam rather than moist air as the drying medium. The advantages are that the medium no longer provides an environment in which heated loose fibers can bum or potentially form explosive mixtures and there are economies in energy use. The heat used in vaporization can be partially recovered elsewhere in a process by the condensation of the vapor. On the other hand, a potential drawback is that the elevated temperatures that are associated with superheated-steam drying may cause thermal damage to the fibers. For example, both Nicholls [100] and Sweetman [101] found that thermal damage to wool fibers in the presence of steam increased with temperature, exposure time, and the pH of the examined material. Presence of water in the material enhanced the degradation. These tests, however, exposed the material for longer than one might expect in any drying installation. Other tests [102] with the through-drying of small samples of wool fabric showed that very little yellowing of the fibers occurred at temperatures below 150°C, but very wet samples (at initial moisture contents above the critical value) showed a significant color change at the beginning of the process within the first 30 s.

At a sufficiently high temperature, the evaporation rate from a wetted surface into its pure superheated vapor is higher than into perfectly dry air. The temperature at which this transition rate occurs is called the inversion value. Experimental values for this temperature range between 160 [103] and 230°C [104] in comparing steam with air. The inversion occurs because of the difference in properties between steam and air (which have different temperature coefficients) and the absence of a gas side mass transfer resistance when

a liquid evaporates into its pure vapor. At low temperatures, the evaporative process is heat transfer controlled and evaporation into perfectly dry air is the faster process; whereas at higher temperatures, the mass transfer resistance becomes relatively significant for evaporation into air and the superheated-vapor process then becomes faster. The actual magnitude of the inversion temperature depends upon the geometry of the evaporating surface and the temperature and humidity changes in the direction of the gas flow.

As pointed out by Schwartz [102], the inversion temperature is also a function of gas composition. There is a different “point” inversion temperature in the way the comparison is evaluated. A comparison of the evaporation into perfectly dry air with that into moist air with diminishing humidity from the limit of pure steam ($T_{0,1-x}^i$) is different from comparing the evaporation into pure steam with that into moist air with increasing humidity from the limit of dry air ($T_{1,x}^i$). Figure 33.15 shows the results of calculating the inversion temperatures for evaporation within a wetted-wall column with a gas core of 29 mm diameter in turbulent flow. The point inversion temperatures range from 155°C for ($T_{0,1}^i$) and 300°C for ($T_{1,1}^i$) respectively. The two kinds of point inversion temperatures have a common value of 198.6°C, representing the condition, ($T_{0,0}^i$) = ($T_{1,0}^i$), when evaporation into steam is compared with evaporation into pure air from an infinitesimally small surface.

Bond et al. [103] discuss the drying of paper by impinging jets of superheated steam. Svensson [105] discusses applications in the drying of wood pulp while Amoux et al. [106] consider the use of superheated steam in the drying of softwood biomass.

Stubbing [107] describes the use of drying at atmospheric pressure in a steamy environment, which he terms airless drying, with the tower density of the

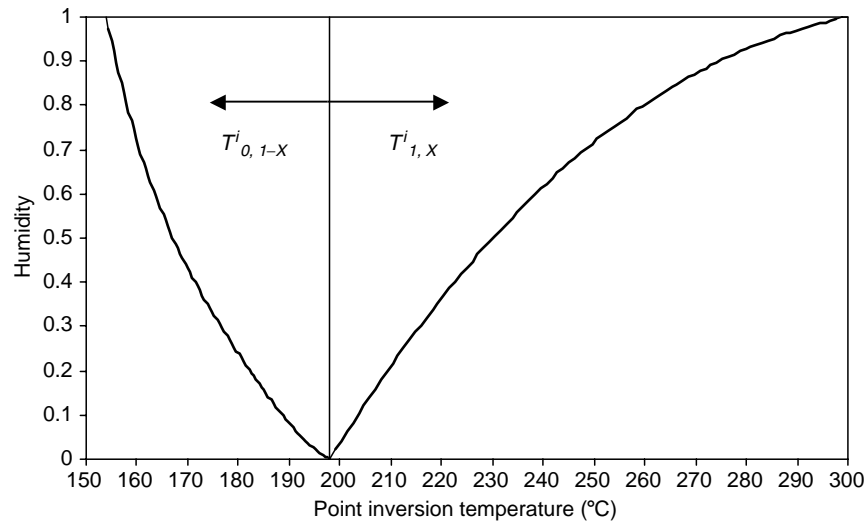


FIGURE 33.15 Variation of the point inversion temperature as a function of air humidity in a wetted-wall column. ($T_{0,1-X}^i$ is temperature on comparing evaporation into perfectly dry air with evaporation into moist air of diminishing humidity; $T_{1,X}^i$ is the temperature on comparing evaporation into pure steam with evaporation into air of increasing humidity. (From Schwartz, J.P. 1999. *Evaluation of the Superheated Steam Drying Process for Wool*, D 82, Diss. RWTH Aachen, Shaker Publisher, p. 89.)

steam relative to the air being used to seal the chamber, so that the air is progressively displaced. A possible application in the drying of cloth is suggested.

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NOMENCLATURE

a, b, c	coefficients
a	activity
a	exposed surface per unit area
A	coefficient
B	moisture flux coefficient
c	moisture concentration
c_e	equilibrium moisture concentration
c_0	initial moisture concentration
c_s	surface moisture concentration
c_v	moisture vapor concentration
c_w	moisture liquid concentration
C	coefficient
C_L	specific heat of liquid moisture
C_P	heat capacity at constant pressure
C_S	specific heat of dry solids
d	yarn diameter
D	moisture diffusion coefficient

f	relative drying rate function
F	solids holdup
g	reduced relative drying rate function
G	specific dry gas rate
J	total transfer flux
J_0	zero-order Bessel function
J_1	first-order Bessel function
k	mass transfer coefficient
k_i	coefficients
K	drying coefficient
L	characteristic length
M_s	mass of sheet
M_t	mass of moisture at time t
M_∞	mass of moisture at infinite time
N_v	drying flux
N_w	drying flux in first drying period (constant-rate period)
N_{w0}	drying flux in first period at the air inlet
p	moisture vapor pressure
p^0	saturation vapor pressure
$P(w)$	probability of depth having a value w or less
r	radial distance
R	radius of fiber
R	universal gas constant
s	a parameter
s	spacing between threads
t	time
T	temperature
T_A	ambient temperature
T_F	fiber temperature
T_0	initial temperature

T_S	steam temperature
T_V	moisture vapor temperature
T_W	moisture liquid temperature
u	moisture concentration function
u_B	conveying band speed
u_G	gas velocity
u_S	solids velocity
U	overall heat transfer coefficient
w	width function
x	relative humidity function
x	distance
X	moisture content (dry basis)
X_{cr}	critical moisture content
X_e	equilibrium moisture content
X_1	moisture content for complete monolayer
y	normal distance
Y	humidity
Y_G	bulk gas humidity
Y_{G0}	bulk gas humidity at air inlet
Y_W	wet bulb humidity
z	longitudinal distance
Z	total longitudinal distance

Greek Symbols

α	a Sherwood number
β	mass transfer coefficient
γ	voidage function
Γ	relative free moisture content (based on initial value)
δ	nondimensional layer depth
ε	strain
ε	voidage
η	volumetric heat of wetting
θ	relative extent or time
κ	thermal diffusivity
λ	a function
ν	a function
ξ	relative distance or NTU
Π	humidity potential
ρ	density
ρ_B	bulk density
ρ_F	fiber density
ρ_G	gas density
ρ_s	solids density
σ	change of moisture content with vapor concentration
ϕ	a ratio
ϕ_M	humidity potential coefficient
Φ	characteristic moisture content (based on critical value)
Φ_0	initial value of characteristic moisture content
ω	change of moisture content with temperature

φ	relative humidity
ΔH_V	latent heat of vaporization
ΔH_W	heat of wetting

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34 Drying of Textile Products

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34.1 INTRODUCTION

The word “textiles” comes from the Latin *textilis*, meaning “woven”; but in textile science, textile is defined as any product made from fibers. Thus textiles refers not only to woven fabrics but also to nonwoven fabrics, knitted fabrics, tufted fabrics such as carpets and bedspreads, and specially constructed fabrics [1]. Figure 34.1 depicts the major segments and linkages of the textile industry, from fibers to products. The textile mill portion of the textile complex includes many chemical wet processes such as slashing, dyeing, printing, latex bonding, and finishing. In many of these processes, drying is required to remove the excess moisture in the porous materials to produce the desired product. For example, the typical steps used to produce latex-backed tufted carpet are shown schematically in Figure 34.2.

As there are many textile products ranging from yarns to carpets which have to be processed, many different drying processes are used by the textile industry. To complicate things further, various processes are used for the same product. Typical drying systems used by the textile industry for drying fabrics and tufted carpets are discussed in this chapter.

The term drying is commonly used to describe the process of thermally removing the volatile substances from a product [2]. In textiles, the term is more generally used to mean the dewatering of a product. Mechanical dewatering is generally much less expensive than thermal drying. Thus, as much water as possible is usually removed mechanically.

Approximately 25% of the energy used in wet processing is consumed in drying [3]. In addition to energy cost, drying is a common bottleneck in the wet processes. Thus, improvement of the drying rate and

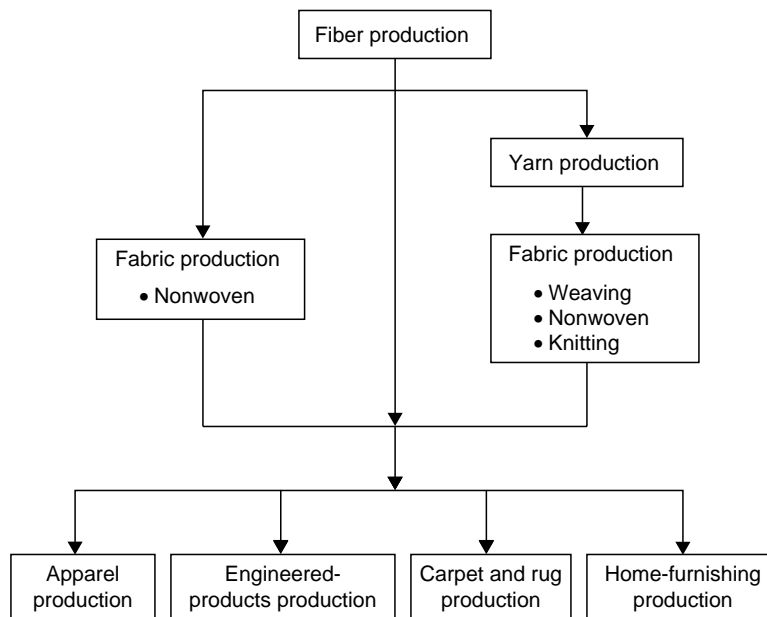


FIGURE 34.1 The textile industry.

energy efficiency of the drying system and selection of the proper type of dryers are of great importance to the textile industry.

In this chapter, moisture regain (%), defined as the weight of water in the textile to the dry weight of

textile, is used to indicate the amount of moisture contained in the textile.

34.2 CLASSIFICATION OF DRYING STEPS FOR TEXTILE PRODUCTS

The drying or dewatering processes can be classified either as predrying and final drying or as mechanical and thermal drying. Predrying is accomplished using either mechanical or thermal processes whereas final drying is achieved using thermal processes. Predrying is the lowering of the moisture regain to some predetermined level. This is followed by the final-drying step in which the moisture regain is lowered to the desired level. For example, fabrics that are heat set in the drying process will be predried by some means to a moisture regain of approximately 30%. They will be transported through a convection oven using a tenter frame that fixes the width of the fabric, giving it a dimensional stability. Mechanical dewatering is usually used to predry fabrics when feasible because it requires much less energy expenditure per mass of water removed (Table 34.1) [4]. However, the level to which the moisture regain can be lowered by mechanical methods is usually too high, and thermal processes are required to obtain the final moisture regain.

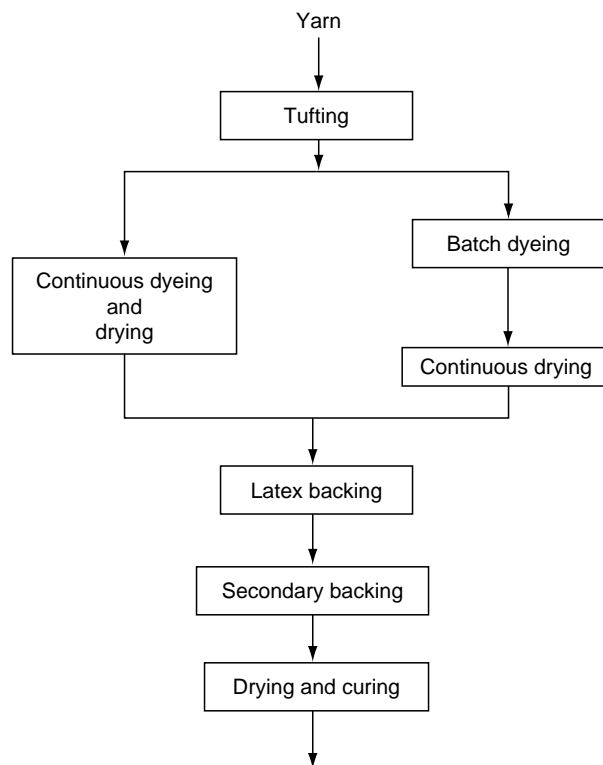


FIGURE 34.2 Process diagram of a typical carpet mill.

34.2.1 PREDRYING AND FINAL DRYING

Drying is sometimes separated into two steps: (1) predrying and (2) drying. This may be done for either economical or technical reasons. Predrying may be

TABLE 34.1
Typical Energy Requirements for Common Textile Drying Equipments

Equipment Type	Energy Requirement (kJ/kg) ^a
<i>Mechanical dewatering</i>	
Squeeze roll	58
Vacuum extractor	700
<i>Thermal drying</i>	
Steam can	4700
Convection dryer	7000
Radiant predryer	9300

^aKilojoules per kilogram of water removed.

Source: From Georgia Institute of Technology, Report No. ORO-5099-T1, U.S. Department of Energy, 1978, p. 15. With permission; Badin, J.S. and Lowitt, H.E., Report No. DOE/RL/01830-T56, U.S. Department of Energy, January 1988. With permission.

carried out by methods that are efficient and require low energy input. This will be followed by final drying either because the moisture regain that can be obtained with the predrying method is limited or because better product quality can be obtained by final drying using some other method. Sometimes predrying is used for technical reasons. For example, in continuous dyeing, fabric saturated with dye solution must be dried to fix the dye. The moisture regain must be lowered uniformly to about 30% to prevent dye migration. Predrying is often accomplished using gas infrared (IR) systems and final drying is achieved using either steam cans or convection ovens.

34.2.2 MECHANICAL DEWATERING

Several methods are available for mechanically reducing the amount of water in textile materials. The most common techniques are mangling or squeezing, vacuum extraction, and centrifuging. The technique selected is influenced by the type of material to be processed and on whether a batch or continuous process is to be used. Many delicate fabrics are damaged by squeezing between rollers or spinning in a centrifuge. Since vacuum slots cause less damage, they are often used with delicate fabrics. Centrifuging is normally limited to batch processes. Mangling followed by vacuum extraction is often used for continuously processing fabrics and carpets. Mangling removes free water very efficiently, and vacuum extraction lowers the moisture regain to levels unachievable with mangling alone.

The moisture regain achievable with mechanical methods depends significantly on the type of process used and on the type of textile being dried. For

example, the reduction in moisture regain obtainable with squeezing rollers varies with both the fiber type and the textile construction. For structures composed of hydrophilic fibers such as cotton, the moisture regain can typically be reduced from about 150% to 60–100%. On the other hand, for structures composed of hydrophobic fibers such as polyester, significantly lower moisture regain can be obtained.

34.2.3 THERMAL DRYERS

Thermal dryers are sometimes used for predrying, but are almost always used for final drying because of the limitations of mechanical dryers. After mechanical predrying, much of the remaining water is chemically bonded to the fiber and must be evaporated. This is accomplished using several types of thermal system such as heated cans, convection ovens, and radio frequency (RF) dryers. Recently, microwave dryers have been designed for drying textiles and carpet tiles [5].

34.3 PROPERTIES OF TEXTILES

The physical, thermal, and sorptive properties of textiles are very important in calculating drying-energy usage, modeling the thermal drying processes, and determining the dryer operation conditions. The physical and thermal properties of several fibers are given in Table 34.2 [6]. The equilibrium moisture regain vs. relative humidity of several fibers is given for absorption and desorption in Table 34.3 [7]. As illustrated in

TABLE 34.2
Physical and Thermal Properties of Fibers

Material	Density (g/cm ³)	Specific Heat (J/kg/K)	Thermal Conductivity (W/m/K)	Thermal Diffusivity (cm ² /s)
<i>Natural Fiber</i>				
Cotton	1.52	1250	0.07	0.000368
Cotton bats	0.08	1300	0.06	0.00577
Wool	1.34	1340	—	—
Wool bats	0.5	500	0.054	0.00216
<i>Synthetic Fibers</i>				
Nylon 6, 66	1.14	1419	0.25	0.001545
Polyethylene	0.97	1855	0.24	0.001334
Poly(ethylene terephthalate)	1.37	1103	0.14	0.000926
Polypropylene	0.93	1789	0.12	0.000721
Polyacrylonitrile	1.18	1286	—	—
<i>Others</i>				
Carbon fiber	1.8–2.1	710	05–500	0.10–3.97

Source: From Warner, S.B., in *Fiber Science*, Englewood Cliffs, NJ: Prentice-Hall, 1995. With permission.

TABLE 34.3
Moisture Regain of Fibers for Moisture Absorption and Desorption at 21°C

Fiber	Relative Humidity (%)									
	10	20	30	40	50	60	70	80	90	95
Cellulose acetate	0.52 ^a /0.80	0.87/2.70	1.67/4.47	2.73/5.40	3.53/6.50	4.27/7.81	5.71/8.83	6.67/10.59	7.96/13.54	9.12/14.01
Cotton	1.40/2.28	2.08/4.38	3.72/5.67	4.86/6.06	5.86/7.61	6.95/8.36	7.90/9.71	9.45/11.72	11.04/13.90	12.74/14.12
Kevlar aramid	0.88/1.45	1.73/2.69	1.98/3.01	2.58/3.96	3.58/5.38	3.82/5.55	4.12/6.15	4.68/6.73	5.51/6.97	5.49/7.14
Nomex aramid	1.15/2.66	2.34/3.88	2.66/4.48	3.27/5.01	4.38/5.68	4.70/5.76	5.00/6.03	5.48/6.27	6.15/6.75	6.45/6.80
Nylon 66	0.44/0.70	0.78/1.77	1.49/2.36	2.17/2.69	2.70/3.33	2.70/3.87	3.77/4.40	4.45/5.30	5.01/5.53	5.47/6.02
Poly(ethylene terephthalate)	0.02/0.04	0.07/0.14	0.12/0.20	0.28/0.29	0.34/0.36	0.39/0.43	0.43/0.47	0.50/0.55	0.53/0.55	0.53/0.55
Silk	1.14/3.45	3.25/6.20	5.20/7.70	6.76/9.72	7.91/11.71	8.11/11.51	10.67/12.26	11.96/14.85	13.54/17.36	15.74/20.97
Viscose rayon	2.38/3.96	4.87/8.28	7.03/10.62	8.76/12.42	10.45/13.65	12.20/14.67	14.39/16.13	16.22/18.32	18.36/20.57	20.65/25.00
Wool	2.14/3.68	3.58/8.13	6.67/10.80	8.28/12.60	10.02/14.78	12.39/16.08	13.62/17.71	15.33/19.33	17.26/20.20	19.34/21.09

^aFirst number is for moisture absorption and second number is for moisture desorption.
Source: From Fuzek, J.F., *Ind. Eng. Chem. Prod. Res. Dev.*, 24, 140–144, 1985. With permission.

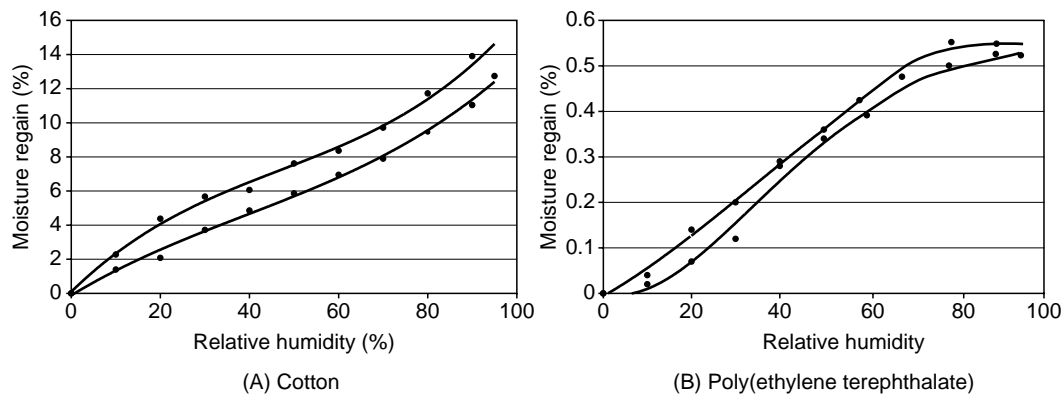


FIGURE 34.3 Moisture regain vs. relative humidity for absorption and desorption of cotton and poly(ethylene terephthalate) fibers. (From Fuzek, J.F., *Ind. Eng. Chem. Prod. Res. Dev.*, 24, 140–144, 1985. With permission.)

Figure 34.3, comparison of the equilibrium moisture regains for absorption and desorption shows that the hysteresis [8] is large for hydrophilic fibers, such as cotton, and low for hydrophobic fibers, such as polyester.

34.4 TYPICAL DRYERS FOR TEXTILES

The typical dryers used for textiles and carpets are summarized in Table 34.4. These dryers are briefly described below.

34.4.1 MECHANICAL DEWATERING

The most common procedure for mechanically dewatering in continuous processes is the squeezing using nip rollers followed by vacuum extraction. In processing batch materials, centrifuging is commonly used.

34.4.1.1 Squeezing

The first step in the dewatering process is usually squeezing or mangling, shown schematically in Figure 34.4. It is accomplished by passing the textiles between a pair of rollers and is by far the least expensive method for dewatering fabrics. The energy requirements of water removed are only approximately 58 kJ/kg. However, squeezing can reduce the moisture regain only by 60 to 100%, depending on the properties of the fibers and the fabric construction.

34.4.1.2 Vacuum Extraction

The next step in dewatering textiles and carpets is usually vacuum extraction as shown schematically in

TABLE 34.4
Typical Dryers Used in Textile and Carpet Industry

Application	Drying Step	Type
<i>Yarn preparation and weaving</i>		
1. Slashing	Predrying	Squeezing
	Drying	Steam cans
<i>Fabric finishing</i>		
1. Preparation	Predrying	Squeezing/vacuum extraction/steam cans
	Drying	Steam cans or convection oven
2. Dyeing and printing		
Dyeing	Predrying	Squeezing/vacuum extraction/steam cans
	Drying	Infrared oven
		Steam cans or convection oven
Printing	Drying	Convection oven
3. Finishing	Predrying	Squeezing/vacuum extraction/steam cans
	Drying	Steam cans or convection oven
<i>Floor covering (tufted carpet)</i>		
1. Drying after dyeing	Predrying	Squeezing/vacuum extraction
	Drying	Convection Oven
2. Application of secondary backing	Drying	Convection Oven

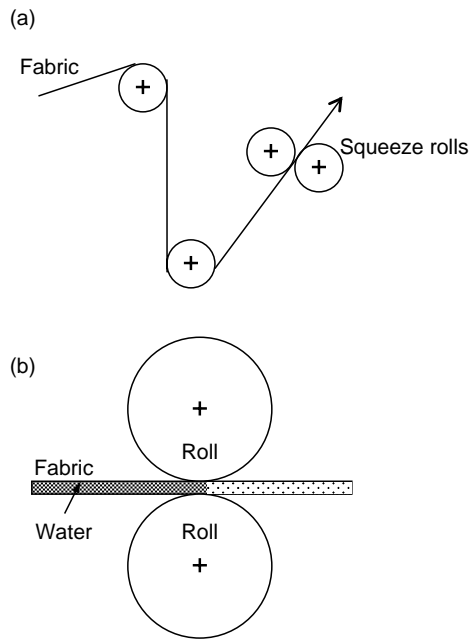


FIGURE 34.4 Dewatering using squeeze rolls.

Figure 34.5. With this method, water is extracted from the textile in open width as it passes over a slotted or perforated box in which vacuum is maintained by a pump. The final moisture regain depends on operating parameters, such as initial moisture regain, vacuum pressure, and production speed, and is also highly dependent on textile construction and fiber hydrophilicity. The energy requirements (700 kJ/kg of water removed) are high compared with squeezing, but much lower than for thermal methods. The typical moisture regain achievable for fabrics made from several types of fibers is shown in [Table 34.5](#) [9]. It is about 30 to 50% for tufted nylon carpets, depending on the vacuum system and carpet construction.

34.4.1.3 Centrifuge

Centrifuging involves rotating the textile at high speed to remove water, and is normally limited to batch processing of materials such as skeins of yarns, packages of yarns, and small rugs. The moisture regain of various kinds of fibers after centrifuging is shown in [Table 34.6](#).

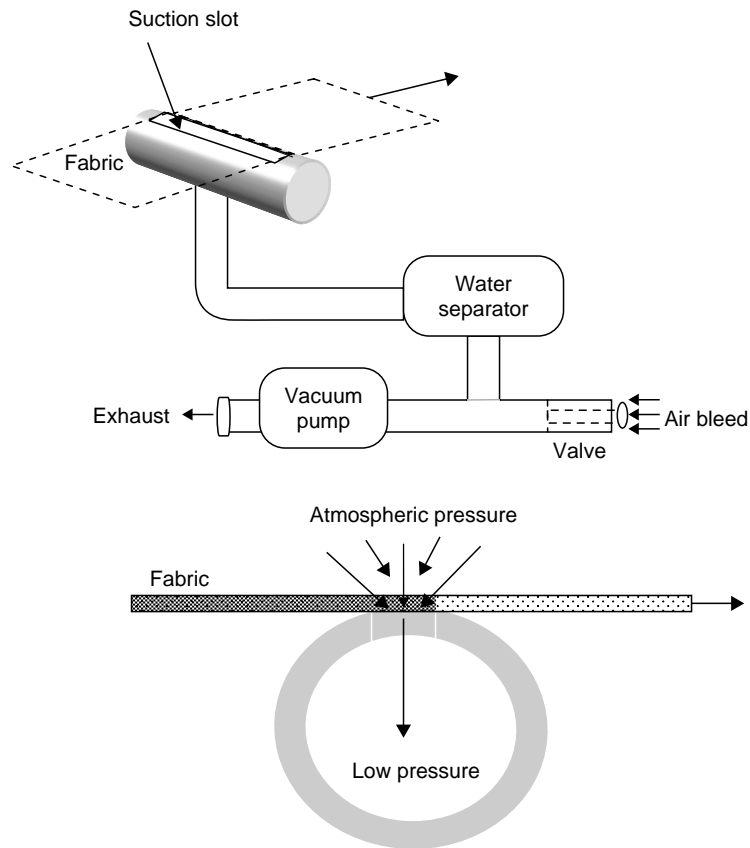


FIGURE 34.5 Schematic of a vacuum extraction.

TABLE 34.5
Moisture Regain after Vacuum Extraction

Fabric	Vacuum (Hg Vac)	Moisture Regain (%)
100% Polyester	15"	10–15
100% Polyester	10"	25–30
80% Polyester/20% cotton	15"	25–30
65% Polyester/35% cotton	15"	35–40
100% Cotton	17"	50–55
50% Rayon/50% cotton	17"	55–65

Source: From Ostervold, J.A., *America's Textiles International*, 11, 16j–16l, 1982. With permission.

34.4.2 CONDUCTION DRYERS

Conduction drying involves placing the surface of the material in direct physical contact with a heated surface. An advantage of conduction dryers is that heat-transfer rates achievable are usually higher than for convection drying. A disadvantage is that the direct contact with the solid surface may cause damage to the textile.

34.4.2.1 Hot-Cylinder Drying

Conduction drying is usually carried out using hot cylinders, which are rotating metallic cans that are heated using steam or special heat-transfer liquids. When steam is used, the drying system is often referred to as steam cans, which are shown schematically in Figure 34.6. The energy requirements removed are typically around 4700 kJ/kg, which is usually lower than that of other thermal dryers.

Steam cans are used in predrying and final drying of fabrics. They are often used to predry fabrics that

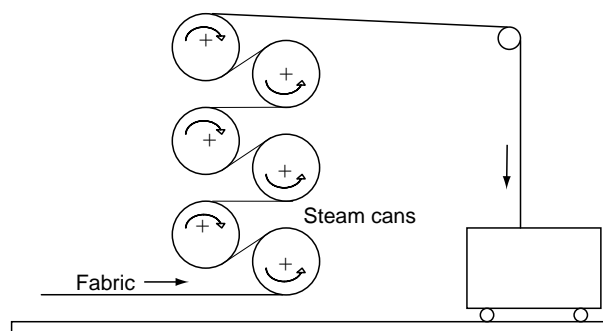


FIGURE 34.6 Schematic of a steam can dryer.

must be dimensionally stable after final drying. The steam cans are used to reduce the moisture regain to about 30%, and convection ovens with tenter frames are used for final drying. In some cases where contact with the hot surface does not negatively affect fabric properties, all of the thermal drying is achieved with the steam cans.

Although the direct contact of the fabric with the hot surface gives high heat-transfer rates, it can cause problems. One problem is that the pressure between the can and the fabric can distort the surface of delicate fabrics. Another problem occurs when a fabric containing materials such as dyes, print paste, and adhesives touches the hot surface. Sometimes materials from the fabric will transfer to the hot surface, and then subsequently transfer to regions of the fabric, causing unwanted effects.

34.4.3 CONVECTION DRYERS

Convection dryers are the most common type of dryers used for drying textiles and carpets. The drying medium is usually hot air though steam can be used if the temperature can be raised sufficiently high without damaging the textile [10,11]. The manner in which air is applied to the product greatly affects the heat-transfer and drying rates that are achievable in convection ovens. There are three types of convection dryers: (1) circulating-air, (2) through-air, and (3) impingement.

34.4.3.1 Circulating-Air Dryers

In circulating-air dryers, drying is achieved by transporting the textile through hot air that circulates through the oven. Products that have varying sizes and weights such as towels and washcloths are looped over hangers that slowly pass through the oven. Drying rates are usually low, and residence times in this type dryer (sometimes referred to as loop dryers) are usually in the order of 40 to 45 min.

TABLE 34.6
Moisture Regain after Centrifuging

Material	Moisture Regain (%) ^a
Acetate	31
Cotton	48
Nylon	16
Silk	52
Viscose rayon	103

^aMoisture regain after rotating for 5 minutes using a centrifuged force 1000 times the gravitational field.

Source: From Morton, W.E. and Hearle, J.W.S., *Physical Properties of Textile Fibers*, 3rd ed., Manchester, UK: The Textile Institute, 1993. With permission.

34.4.3.2 Through-Air Dryers

Through-air (flow-through) dryers have been widely used to remove moisture from various textile products for over 50 years. The principle of through-air drying is to evaporate moisture by forcing the air to flow through the porous material. The basic concept is illustrated in Figure 34.7. As through-air drying greatly increases the contact surface area between the hot air and the wet surface, it provides high overall drying rate.

There are two types of through-air dryers often used in the textile industry: (1) suction or perforated drum drying systems and (2) convection ovens with tenter frames for fixing the transverse dimension of the product while it is transported through the oven.

Suction or perforated drum dryers are often used to dry fabrics, particularly nonwovens, and sometimes unbacked tufted carpets. These dryers usually consist of two or more perforated drums mounted horizontally in a compartment (Figure 34.8). Several two-drum compartments are typically linked together to form a complete machine. Fans draw air from the interior of the drums producing suction on the surface area in contact with the material. This suction holds the material to the surface of the drum permitting hot air to pass through the material being dried. A portion of the drum has no suction, which permits the material to transfer to the next drum without interference.

The second type of through-air dryer commonly used to dry fabrics and unbacked tufted carpets is shown schematically in Figure 34.9. It is a convection oven with a tenter frame that controls the transverse dimension of the product while it transports the fabric through the dryer. Fans blow hot air in the chamber on one side of the fabric creating a pressure drop across the fabric, causing hot air to flow through the fabric. Some of the warm moist air exiting the fabric is usually reheated and recycled through the fabric.

In the through-air drying process, most of the energy is required for heating air and blowing it through the wet material. For a given blower configuration, the velocity of the air flowing through the material depends on the air permeability of the

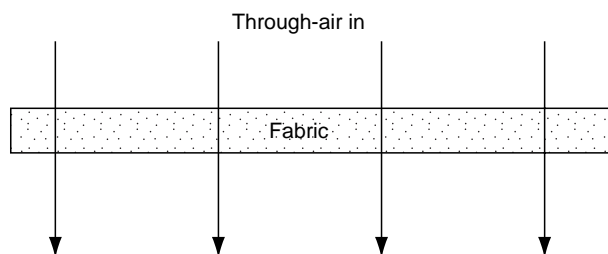


FIGURE 34.7 Schematic of a through-air drying system.

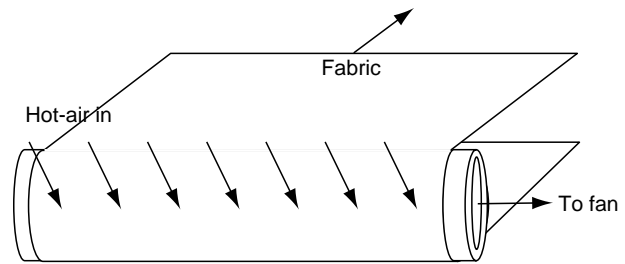


FIGURE 34.8 Schematic of a suction drum dryer.

material. Air permeability is a measure of how easily air can flow through a unit area of porous material at a given pressure drop. As the air permeability of the material increases, the airflow rate increases resulting in a higher drying rate. In this case, through-air drying is a good option; however, if air permeability of the material is too low, the electrical energy for driving the fan to obtain a required airflow rate would be too high. In this case, impingement drying is often used instead of through-air drying.

34.4.3.3 Impingement Dryers

Convection impingement drying systems are often used to dry textiles with low air permeability, for example, latex compounds used to back tufted carpets. This technology utilizes columns of high velocity hot air directed at the product surface via nozzles mounted above and below the product, as shown schematically in Figure 34.10. Due to the complex fabrication and high air-handling costs of impingement dryers, they are preferably used for thick fabrics and fabrics with backings such as latex.

Typically, the impingement air hits perpendicularly or near perpendicularly onto one or both of the product surfaces. If air impinges onto both surfaces, a

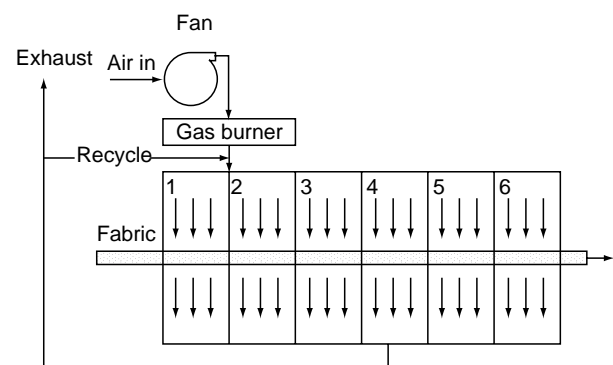


FIGURE 34.9 Industrial through-air dryer.

tenter frame is often used to control the transverse dimension and to transport the product through the dryer. If air impinges onto only one product surface, the product is typically supported on a mesh screen, conveyor, or roller.

The designing of impingement dryers may involve selecting nozzle configuration and geometry, determining velocity and temperature of impingement air, nozzle-target spacing, calculating drying rate, estimating air-recycle ratio, etc. The details of designs of impingement dryers are well described in another chapter of this handbook.

34.4.4 INFRARED DRYERS

Thermal-radiation emitters, referred to as IR emitters, are used primarily to predry textile and carpet products. Thermal radiation is a mode of heat transfer characterized by energy transport in the form of electromagnetic waves. It is the energy emitted by the body solely by virtue of its temperature. Although high-temperature emitters emit thermal radiation in the IR, visible, and ultraviolet regions of the electromagnetic spectrum, almost all of the radiations are in the IR region (Figure 34.11). Thus the dryers using the thermal-radiation emitters are referred to as IR heaters.

IR ovens are typically used to predry materials that are finally dried in tenter-frame convection ovens. They are also used to augment the existing ovens where additional production is needed and space is limited. IR ovens are not normally used to accomplish final drying because IR energy is absorbed by dry textiles and there is a danger of scorching or burning the product.

The thermal radiation heat-transfer rate, q_{rad} , emitted by the IR emitter is equal to $A\varepsilon\sigma T^4$, where A is the emitter surface area, ε is the emissivity of the emitter, σ is the Stefan-Boltzmann constant ($5.67 \times 10^{-8} \text{ W/m}^2/\text{K}^4$), and T is the temperature of the emitter. As the energy emitted varies with the fourth power of temperature, the heating power

densities of IR emitters vary greatly with emitter temperature. IR systems with high-temperature emitters are available that can provide much higher energy densities than conduction and convection dryers.

The depth of IR penetration into the textile depends on the wavelength. The shorter the wavelength, the deeper is the penetration. However, even at short wavelengths, the maximum depth is typically no greater than about 1.5 mm. Thus, IR drying is generally used for thin materials such as fabrics in dyeing, finishing, or coating processes.

Two types of IR emitters (electric and gas-fired) are used in IR dryers for textiles. Both types of IR emitters can be used to produce medium-wavelength (2 to 4 μm) and long-wavelength (4 to 10 μm) IR radiation; however, only the electric IR emitter can generate the short-wavelength (0.76 to 2 μm) IR radiation. For the electric IR drying systems, 100% of the input energy is utilized to heat the emitters and up to 86% is converted to IR radiation. For the gas IR drying system, 25 to 50% of the input energy typically goes into the flue gas and is exhausted. Thus, the gas IR drying systems usually produce less radiation per unit of input energy. However, cost per unit of input energy for electricity can be 3 to 5 times that of gas. Based on the efficiency and the unit price of energy, the energy cost for drying must be calculated for each application.

IR drying is accomplished using various setups. An IR emitter suspended over the textile or mounted vertically with the textile moving parallel to it transfers heat that may be adequate in some cases. However, placing the IR emitter in a properly designed enclosure (Figure 34.12) enables it to be more energy efficient, create better temperature uniformity across the web, put less load on the plant air-conditioning system, and be safer for personnel [12]. A low-velocity air impingement between the IR emitters is useful in removing the moisture-laden air from the surface of the textile.

34.4.5 RADIO FREQUENCY AND MICROWAVE DRYERS

Although RF and microwave dryers have been used by the textile industry, the market penetration of RF dryers is much greater. The cost of the equipment has been a barrier to the use of both types of dryers, and uniformity of treatment has been a problem associated with microwave drying. As RF and microwave systems are operating on electric power, the unit of energy cost is usually higher than for conventional dryers. However, the energy efficiency of RF and microwave systems are normally much higher than for conventional system, which tends to offset the higher unit of energy cost.

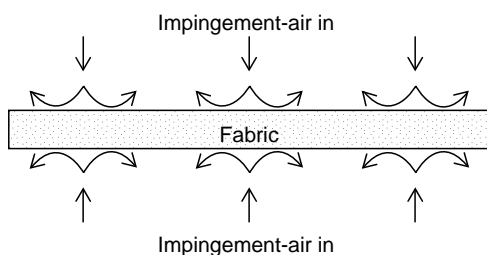


FIGURE 34.10 Schematic of a double-impingement drying system.

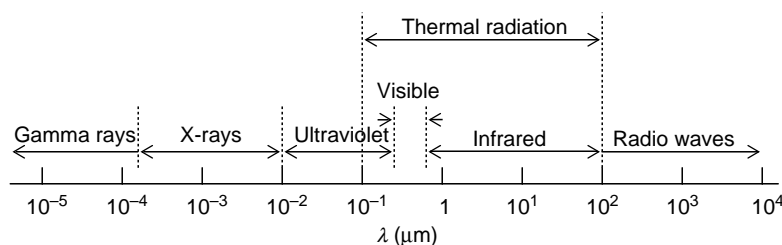


FIGURE 34.11 The electromagnetic spectrum.

RF and microwave dryers are based on a thermal effect known as dielectric heating. When a dielectric, i.e., a material which is an electrical insulator, is placed in an alternating electric field, successive distortion of the molecules causes heating. Although there are several heat dissipation mechanisms, only two are of major importance to heating textiles. One is ionic conduction, where ions are accelerated by the electric field and collide with other molecules, giving up kinetic energy. The other is dipolar rotation, where dipoles or induced dipoles in the molecules making up the material tend to align with the electric field. When the field alternates, the dipoles tend to rotate to follow the electric field. The rotating dipoles interact with the surrounding material and the intermolecular friction results in heat being given off [13].

The frequencies for RF heating are between 1 and 100 MHz while those for microwave heating are between 300 MHz and 300 GHz. However, Industrial, Scientific, and Medical (ISM) bands for industrial heating have been established by international agreement. These are 13.56, 27.12, and 40.68 MHz for RF heating and 896, 915, and 2450 MHz for microwave heating [14].

Since 1978, RF dryers have been used by the textile industry primarily to dry bulky textile materials such as hanks, muffs, tops, and yarn packages. Four types of systems are currently being manufactured: (1) batch units; (2) continuous belts; (3) combination RF and air/vacuum system where water is vaporized and removed at a temperature lower than 100°C; and (4) air, RF assisted (ARFA). When compared with conventional dryers, RF dryers usually have the advantage of providing a more uniform final drying and reduced chance of overdrying. The energy requirements of RF dryers depend on process parameters such as fiber type, substrate, and process: for example, 0.53 kWh/kg of water removed for drying 100% polyester packages and 4.91 kWh/kg of water removed for drying loose-stock cashmere [15].

Until recently, the use of microwave systems for drying textiles has been extremely limited. The major limitation of microwave heating has been a problem with uniformity of treatment. Nonuniform treatment can lead to hot spots, resulting in overheating in some areas and underheating in others. Recent research has led to a new approach to microwave drying of textiles. The use of waveguides to serpentine the microwave

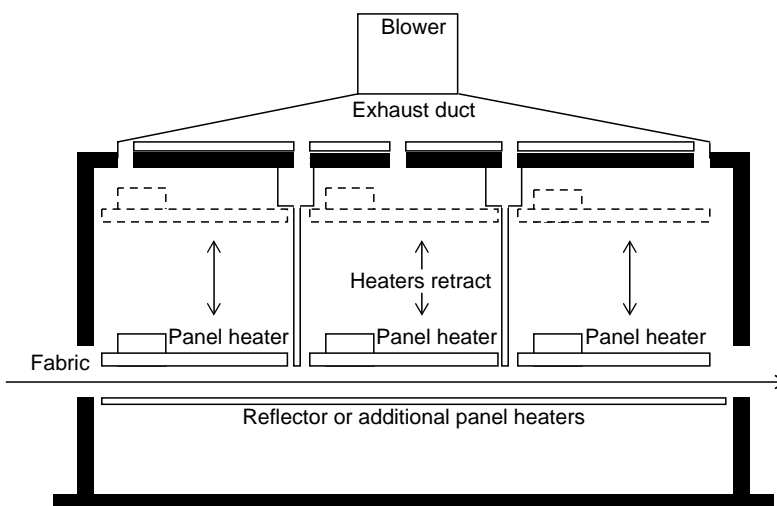


FIGURE 34.12 Schematic of an infrared oven.

energy back and forth across the material being treated gives the needed improvements in uniformity. Microwave drying systems are now being tested for drying a range of textile products including tubular knits, sheets of yarns, terry towels, and carpet tiles [5,16].

34.5 CONCLUSION

As there are many textile products that are to be processed, many different drying systems are used by the textile industry. To complicate things further, various processes are used for the same product. Typical drying systems used by the textile industry for drying fabrics and tufted carpets are briefly discussed in this chapter. No attempt is made to give full design details. Also the mechanical, electrical, or control aspects of these dryers are not discussed. The reader can refer to other chapters of this handbook for more details and additional references.

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35 Drying of Pulp and Paper

Osman Polat and Arun S. Mujumdar

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35.1 INTRODUCTION

Ts'ai-Lun apparently produced a sheet of paper in about A.D. 100 in China and became the first recorded papermaker in the world. However, it took about 1000 y for this new art to reach Europe. In the medieval era, the progress of papermaking was very slow and the major ingredient of paper was old rags. By the beginning of the 19th century, the progress of this industry was enormously accelerated. The first practical paper machine was produced in the early 1800s; then continuous drying techniques were introduced to the industry by means of cylinder drying in 1817 by John Dickinson; later, in the mid-1840s, the extensive use of wood as a cellulose-fiber source began by the advent of the first wood grinder. Today, papermaking has become one of the major industries in the world. The production of paper increased enormously, over 60 million tons per year in the United States alone. The machine speeds also increased up to 10 to 15 m/s and even higher for tissue products, to keep pace with the increased production rates.

Papermaking is essentially a massive dehydration operation. A fiber–water suspension with initial consistencies of 0.2 to 1.0% (consistency = grams of fiber per gram of fiber–water suspension) is delivered to a screen, where, with the application of vacuum, much of the free water is drained off and the consistencies rise to about 18 to 23%. Then, more of the free water is removed by mechanical “squeezing” at the press section. The sheet is then transferred to the drying section, with a consistency of 33 to 55% to remove the remaining excess water to obtain the final product with 6 to 9% moisture content. In the United States, the production of over 60 million tons of paper per year entails the removal of over 80 million tons of water by thermal dryers. Considering that for a typical newsprint machine the water removed in the dryers is less than 1% of the original water, one can easily realize the amount of water that must be removed in the papermaking process.

Although drying removes the least amount of water in absolute terms, it still remains the most costly and energy-intensive step in the papermaking

process. Thus, improvement of the efficiency of water removal before drying and improvement of the drying system and its thermal efficiency without adversely affecting the product quality are of great importance to the pulp and paper industry.

In this chapter, pulp drying and paper drying are discussed under separate headings.

35.2 DRYING OF PAPER

35.2.1 DRYING PROCESS

Paper is hygroscopic. The transfer of moisture between the paper and the surrounding atmosphere takes place unless the sheet is in equilibrium with the surroundings. However, the amount of water present in the paper at equilibrium conditions depends upon whether it has been taken up or given off by the paper. This hysteresis phenomenon is known for many other hygroscopic substances. The sorption behavior of a paper sample (Figure 35.1) shows that an equilibrium moisture content reached by wetting and drying will be different at the same humidity. There is no satisfactory theory to explain the hysteresis. Luikov [1] suggested two explanations for this phenomenon. One is that the hygrothermal equilibrium sets in slowly, as a result of which the observed equilibrium is not a true equilibrium. The other hypothesis suggests that evaporation and condensation phenomena are irreversible. In drying (desorption), full wetting of the capillary walls occurs. On the other hand, during wetting (sorption), the capillary walls are gradually covered with a layer of liquefied vapor, but the meniscus is not formed until the adsorption layer is sufficiently thick to close the pore at the narrowest point.

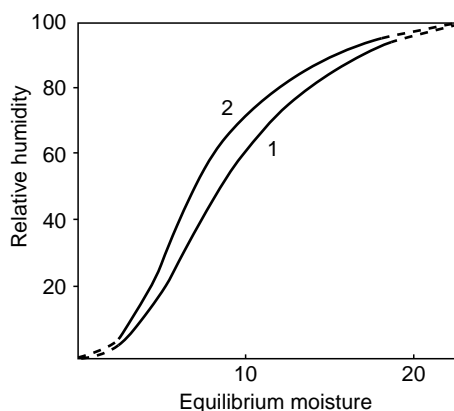


FIGURE 35.1 Sorption behavior of paper sample (1, drying; 2, wetting).

The drying cycle of paper is divided into three fairly distinct stages, as for most materials. The initial warming-up stage is followed by the constant-rate stage, which is followed by one or more falling-rate stages. This idealized concept is shown in Figure 35.2.

The heat is supplied to the sheet to increase its temperature up to a certain value at which the heat demand for evaporation and losses comes into equilibrium with the heat supply. At this point, constant-rate drying begins. During this period, water evaporates from the paper surface and the diffusion of moisture from inside the sheet is rapid enough to keep up with the vapor-removal rate from the surface. When the rate of diffusion cannot keep up with the rate of evaporation, the drying plane recedes into the sheet and a falling-rate stage begins. In his drying studies on various paper and paperboard products, Montgomery [2] observed that the drying rate is fairly constant up to 10 to 15% moisture content, after which it decreases rapidly. The falling-rate period can be divided into three phases. The behavior of moisture movement in these phases is not yet well understood. It is generally accepted that the capillary action and diffusion inside the fibers are responsible for the first and second falling-rate periods in drying, respectively. At the end of these two stages, the sheet is almost “oven-dry.” The last stage is to break up the strong chemical bonds and to remove the final

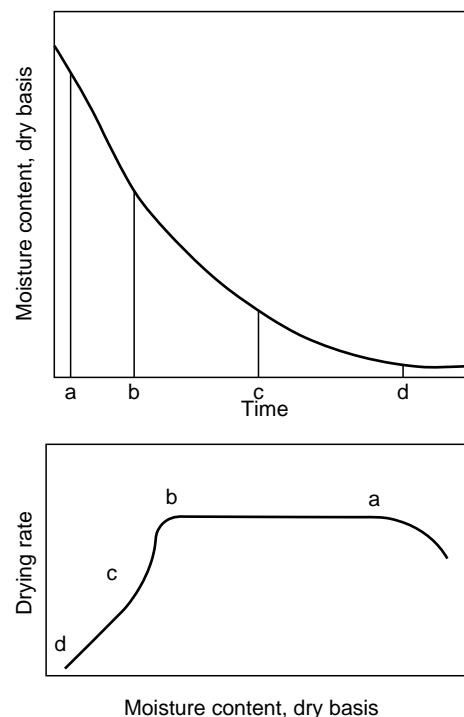


FIGURE 35.2 Typical drying curves.

molecules of water, which is not important for paper-makers as the drying ends at the oven-dry stage for almost all paper products.

35.2.2 TYPES OF DRYERS

The types of dryers used in the paper industry can be classified by the basic means of transferring heat to the paper. For conventional steam-heated cylinders, the predominant mechanism of heat transfer is conduction; convection is for air dryers, such as impingement and through-dryers; radiation is used for infrared dryers; and dielectric heating is used for microwave and radio frequency (RF) dryers. On the other hand, some of the dryers use more than one means of heat transfer, for example, Yankee dryers are a combination of cylinder and impingement dryers.

An estimate of the distribution of dryer types used in the paper industry is shown in Table 35.1. This table shows that the conventional steam-heated cylinder dryers are still dominant in the industry. The others have application to only one type of product, as shown in Table 35.2.

35.2.2.1 Cylinder Dryers

The multicylinder dryer section of a paper machine consists of a number (up to 70) of large, hollow, cast iron or steel cylinders over which the web passes. These cylinders are used to alternately heat the two sides of the sheet. The major mode of heat transfer is the conduction through the steam-heated cylinder shells.

Part of the multicylinder drying section of a newsprint machine is illustrated in Figure 35.3. The sheet is tightly pressed against the cylinders by a dryer felt to enhance heat transfer. This figure also shows some of the important elements of a dryer section such as felt rolls, felt dryers, felt stretchers, and felt guides. The other auxiliary equipments, lead dryer, breaker stack,

TABLE 35.1
Paper Industry Dryer Distribution

Dryer Types	Industry Share (%) ^a
Cylinder dryer	85–90
Impingement dryer	2–3
Yankee dryer	4–5
Infrared dryer	3–4
Through-dryer	1–2

^aPulp dryers excluded.

Source: From Kuhasalo, A., Niskanen, J., Paltakari, J., and Karlsson, M., in *Papermaking, Part 2, Drying*, M. Karlsson, Ed., Fapet Oy, Helsinki, Finland, 2000.

TABLE 35.2
Paper Industry Dryer Distribution by Application
(Pulp Dryers Excluded)

Dryer	Distribution (%)			
	Tissue	Paper	Board	Coating
Cylinder	5	95	95	35
Impingement	—	4	—	50
Yankee	84	—	3	—
Infrared	—	1	1	15
Through	11	—	—	—
Condebelt ^a	—	—	1	—

^aCondebelt is a trademark of Metso Paper Oy, Finland.

Source: From Kuhasalo, A., Niskanen, J., Paltakari, J., and Karlsson, M., in *Papermaking, Part 2, Drying*, M. Karlsson, Ed., Fapet Oy, Helsinki, Finland, 2000.

sweat dryer, size press, spring roll, and dryer doctors, are not shown in the figure.

The number of dryers and the auxiliary equipment used in a dryer section depend on the particular grade to be produced and the speed of the machine. Up to 70 cylinders may be used in a Kraft dryer section; 50 to 55 would be adequate for a typical newsprint or fine paper dryer section. On the other hand, although the breaker stack and size press are necessary for Kraft bag and the hot press, size press, and intercalender dryers for linerboard, none of these is used for corrugating medium [4].

Typically, the dryer cylinders are made of cast iron and have diameters of 0.91 to 1.83 m (mostly 1.52 to 1.83 m in modern machines). The length of the cylinders ranges up to 9.1 m for the largest paper machines. The shell thickness varies with diameter but is generally around 25.4 mm or more. The design and manufacture of these cylinders have important effects on the quality of the finished product. The outer cylinder surface must be highly finished and free of any imperfections to avoid marking the paper; the wall thickness must be uniform throughout the periphery to provide uniform heat transfer. As the machine runs at very high speeds (up to 3000 fpm; 15 m/s), the head, journal bearings, and other parts must also be carefully designed for smooth operation. Cast iron is a widely accepted material for dryer cylinders because of its low cost, corrosion-resistance, and ability to take a fine finish. However, its relatively low resistance to thermal shock may cause some problems [4].

35.2.2.1.1 Mechanism of Heat and Mass Transfer

Drying involves simultaneous heat and mass transfer. In conventional machines using steam-heated cylinders, the temperature of the paper entering the dryer

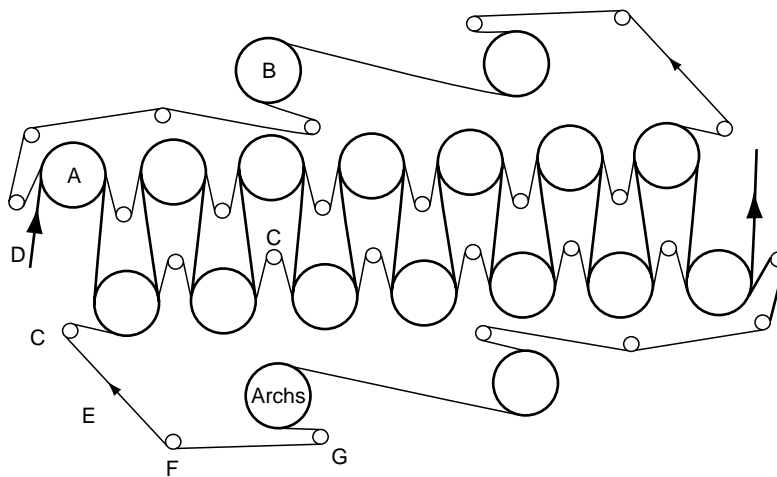


FIGURE 35.3 A typical newsprint dryer section (A, dryers; B, felt dryers; C, felt rolls; D, paper; E, felt; F, felt guides; G, felt stretchers).

section generally varies between 5 and 30°C. The web must be heated to a temperature at which significant evaporation can take place, which is normally between 77 and 93°C. The first two or three cylinders are generally used for this warming-up period, and beyond these the temperature of the sheet is assumed to be the evaporation temperature [5].

The drying process on felt-covered cylinders can be divided into four phases, as illustrated in Figure 35.4 [6]. Nissan et al. described the mechanism of water removal in those phases in a series of papers published between 1954 and 1962. They summed up their conclusions as follows [7]. There are three primary mechanisms for water removal on a cylinder dryer:

1. Direct evaporation in phases 1, 3, and 4
2. Partition of water between the sheet and the cylinder as the sheet leaves the cylinder
3. Extraction by the felt, as both vapor and liquid

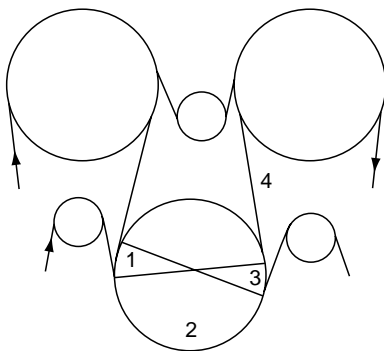


FIGURE 35.4 Phases of felt-covered cylinder drying.

Nissan et al. [7] also proposed a mechanism of moisture transfer from the sheet to the felt:

1. Liquid transfer due to capillary suction
2. Liquid transfer due to the force exerted by expanding gases
3. Evaporation from the sheet followed by condensation in the relatively cooler felt

However, the first hypothesis of Nissan et al. that the dryer felt removed liquid water from paper by capillary suction was disproved by Kirk [8]. A more detailed discussion on this subject can be found in the literature [9]. A more detailed summary of various mathematical models proposed to describe multicylinder drying is given in a recent review [10].

The heat flux for steady-state conduction can be written as

$$q = \frac{kA}{L} \Delta T \quad (35.1)$$

The individual resistances to the heat transfer on a steam-heated cylinder are shown in Figure 35.5. In practice, the overall heat-transfer coefficient, which is the combination of all conductive and convective resistances in the system, is used for calculation of the heat flux. The overall heat-transfer coefficient for the system illustrated in Figure 35.5 is

$$\frac{1}{h_o A_o} = \frac{1}{h_s A_s} + \sum_{i=1}^6 \frac{L_f}{k_i A_i} \frac{1}{h_a A_a} \quad (35.2)$$

The overall transfer area A_o , which is the dryer surface area, must be clearly defined, and all other areas

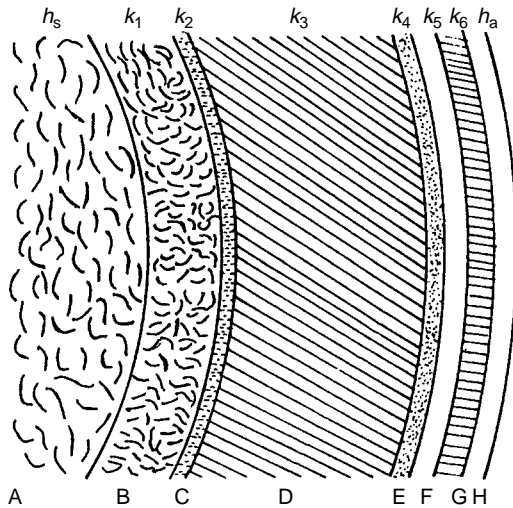


FIGURE 35.5 Resistances to heat transfer on a cylinder dryer (A, steam; B, condensate; C, scale; D, dryer shell; E, dirt and air; F, paper; G, dryer felt or fabric; H, air boundary layer).

should be adjusted to reflect the dryer diameter. Some suggested overall heat-transfer coefficients for various grades of paper are shown in Table 35.3.

In actual practice, paper drying is a transient process. For more realistic results, the following transient heat-transfer equation must be used:

$$\frac{\partial^2 T}{\partial x^2} + \frac{1}{u^2} \frac{\partial^2 T}{\partial t^2} = \frac{1}{\alpha} \frac{\partial T}{\partial t} = 0 \quad (35.3)$$

which assumes very thin and isotropic material. Further, curvature is neglected, as is conduction through the edges. For the fast speeds of commercial machines, the second term of Equation 35.3 is negligible; so, with little error one may write [11]

$$\frac{\partial T}{\partial t} = \alpha \frac{\partial^2 T}{\partial x^2} \quad (35.4)$$

TABLE 35.3
Overall Heat-Transfer Coefficients for Various Grades of Paper

Grade of Paper	h_o (W/m ² K)
Felt paper	45–85
Corrugating medium	140–230
Linerboard	170–230
Kraft sack paper	230–255
Fine paper	255–285
Newsprint	285–315

Nissan and Hansen [12] solved this equation numerically for the conditions of hot-cylinder drying with the following initial and boundary conditions. Initial conditions:

$$t = f(x) \quad \text{for } t = 0 \quad (35.5)$$

Boundary conditions, phases 1 and 3:

$$\begin{aligned} -\frac{\partial T}{\partial x} \Big|_{x=0} &= \frac{h_{cs}}{k} (T_{cyl} - T_{x=0}) \\ -\frac{\partial T}{\partial x} \Big|_{x=X} &= \frac{h_{sa}}{k} (T_{x=X} - T_a) + \frac{\lambda}{k} \left(\frac{\partial w}{\partial t} \right)_{x=X} \end{aligned}$$

Phase 2:

$$\begin{aligned} -\frac{\partial T}{\partial x} \Big|_{x=0} &= \frac{h_{cs}}{k} (T_{cyl} - T_{x=0}) \\ -\frac{\partial T}{\partial x} \Big|_{x=X} &= \frac{h_{sf}}{k} (T_{x=X} - T_f) + \frac{\lambda}{k} \left(\frac{\partial w}{\partial t} \right)_{x=X} \end{aligned}$$

Phase 4:

$$\begin{aligned} -\frac{\partial T}{\partial x} \Big|_{x=0} &= \frac{h_{sf}}{k} (T_a - T_{x=0}) + \frac{\lambda}{k} \left(\frac{\partial w}{\partial t} \right)_{x=0} \\ -\frac{\partial T}{\partial x} \Big|_{x=X} &= \frac{h_{sa}}{k} (T_{x=X} - T_a) + \frac{\lambda}{k} \left(\frac{\partial w}{\partial t} \right)_{x=X} \end{aligned}$$

The calculated temperatures agreed well with their own experimental results for a three-layer muslin sheet on an unfelted cylinder.

Water evaporates at the hot-cylinder interface, and the resulting vapor diffuses through the paper at a rate modeled by Fick's law:

$$w = -D \frac{dc}{dx} \quad (35.6)$$

At the air–paper interface, the mass-transfer equation may be expressed in the following form for simplicity:

$$w = \beta \Delta c \quad (35.7)$$

where β is a mass-transfer coefficient, which embodies both the flow conditions and the fluid properties, and Δc is the concentration difference. The partial pressure of water vapor is the most commonly used concentration term in the paper industry; therefore, Δc becomes the difference between the partial pressures of water vapor at the interface and in the bulk air.

There are two main factors that affect the rate of mass transfer: (1) the sheet temperature, which determines the vapor pressure of the evaporating water at the surface; and (2) the partial pressure of water

vapor in the air near the sheet, which is kept at a low value by replacing the air, i.e., pocket ventilation.

As there are so many different types of drying media in use today to supplement cylinder dryers, TAPPI has outlined the following procedure to calculate the drying rate. The calculation taken from the TAPPI technical information sheet [13] is

$$w = (0.318)S_m \frac{BM}{DN} \quad (35.8)$$

In this method, the drying rate is expressed in kilograms of water evaporated per second per square meter of the drying surface, where drying surface is the circumferential area of the dryer cans multiplied by the sheet width at the reel.

35.2.2.1.2 Steam Supply and Condensate Removal

All cylinder dryers are heated by the steam condensing inside the sheet because of its efficiency (heat-transfer coefficient for film condensation of steam ranges from 5,700 to 17,000 W/m² K (1000 to 3000 Btu/h ft² °F). In the early days of papermaking, steam pressures as low as 14 to 21 kPa (2 to 3 psi) were used; after World War II, the new machines for Kraft papers were made for 586 kPa (85 psi). Today, after the introduction of fabricated steel dryers, mills use 827 to 1016 kPa (120 to 150 psi), the Yankee machines have operated at pressures up to 1206 kPa (175 psi) [5].

In general, the dryer steam can be produced by central steam electric stations, heating plants, or any industrial steam systems. The rotating joints to introduce steam into the dryers and the condensate-removing devices must be specially designed, however.

In a paper machine, the dryers are arranged in groups that have a common steam-supply header. Each group has to operate in such a condition that neither the condensate nor the noncondensables accumulate in the dryer. There are basically two approaches to design the dryer steam and the condensate system. One is the recirculation system in which most of the blow-through is recirculated and the other is the cascade system in which the blow-through steam from higher-pressure sections is used in lower-pressure sections. The recirculation system is more flexible whereas the cascade system is simpler and cheaper [4].

Dryer steam should be free of superheat. Although the condensate inside the dryer desuperheats the steam, a separate desuperheater is used in some cases. The effect of superheating above 100°C is controversial.

The incidence of noncondensable gases (usually air or dissolved gases in feed water) in the steam supply is one of the problems that must be considered when designing the steam system. In most cases, only trace quantities are involved; but if they are allowed

to accumulate inside the cylinder, they cause trouble. The accumulated noncondensable gases affect the vapor pressure of water inside the cylinder, especially near the condensing surface, which in turn affects the condensation temperature at a given total pressure. As a result, the drying capacity and uniformity are reduced. For example, a 5.5°C drop in steam temperature on dryer-limited linerboard grades will result in an approximate 8% production loss [14]. In order to prevent this effect, special steam-supply arrangements are suggested; but the common way to deal with the problem is to simply bleed off sufficient steam with the condensate so that the accumulation of noncondensable gases is prevented [15]. However, recent studies [14] have showed that this practice (bleeds) is very poor and unnecessary.

Another problem that must be considered when designing the steam-heated dryers is the removal of the condensate. If the condensate is not removed from the cylinders continuously, the heat-transfer rate drops and the power load on the drive increases because of the buildup of a heavy load.

At slow machine speeds, the condensate forms a puddle in the dryer. As the speed increases, the centrifugal and viscous forces cause the condensate to climb the side of the cylinder; then the cascading from the top starts and a rim of condensate forms suddenly at higher speeds, as illustrated in Figure 35.6. At speeds higher than 5 m/s (1000 fpm), if the condensate layer is thin, there is a definite centrifugal action and the condensate will form a rim around the inner circumference. It is much better to maintain the

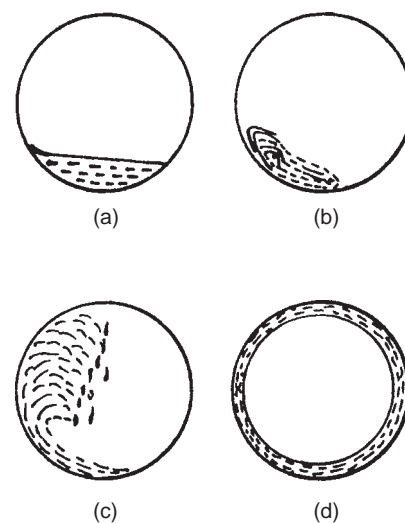


FIGURE 35.6 Condensate behavior at different machine speeds: (a) puddle; (b) climbing puddle; (c) cascading; (d) rimming.

rimming condition because of its relatively lower power requirement, more stable heat transfer, and steady machine speed [5].

Buckets (dippers) are used to remove the condensate in low-speed dryers. No pressure differential is required to lift the condensate above the dryer center-line, where it flows through a pipe to the outside. Buckets can be used in dryers up to the rimming speed [15]. Modern higher speed machines provide a pressure differential to remove the condensate. Syphons are used to pick up the condensate by applying a pressure differential. These can remove the condensate from both a puddle and a rim. A syphon is basically an open-ended pipe leading from the bottom of the dryer through the steam supply, opening to a condensate line. A simplified diagram of a syphon is illustrated in Figure 35.7. The extracted condensate flow depends on the velocity of blow-through steam, the pressure differential, and the distance between the condensate and the syphon inlet. The flow of the extracted condensate can be calculated by using the simplified equation

$$Q_e = Q_s R(P_A - P_B)^{1/2} \quad (35.9)$$

There are fixed and rotary syphons. Different types of tips have been designed to increase the efficiency of fixed syphons. The vibration of the syphon, the length of the cantilever, and the position of syphon for optimum heat-transfer uniformity are the basic problems that must be considered when using fixed syphons.

Rotary syphons may have one or more tips that move with the dryer shell. The basic problems here are the need for a rotary seal at the junction of the rotary pipe and the fixed condensate line and the extra force (centrifugal force) that must be overcome inside the pipe.

A more detailed discussion of steam-supply systems, condensate behavior, and condensate removal can be found in the literature [5,15].

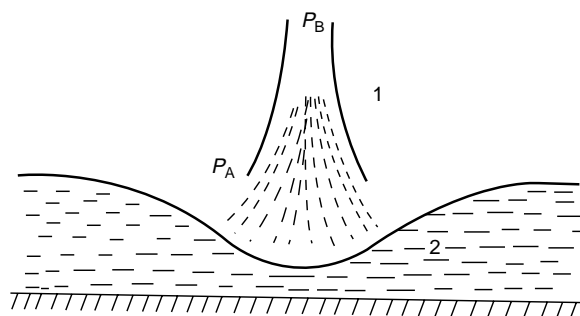


FIGURE 35.7 A simplified diagram of a syphon (1, syphon; 2, condensate).

35.2.2.1.3 Air Systems

The dryer air systems, namely, the dryer hoods and ventilation systems, are designed to pick up and carry the water vapor evaporated from the paper web. These systems affect not only the uniformity and rate of drying and ultimately the quality of paper but also the working conditions and the operating and capital costs.

To remove the large quantities of water vapor involved in paper drying and to keep the dew point of the exhaust air below the wall temperature of the hood, large volume of air is required. As the handling and heating of air is expensive, there are a variety of commercially available hood designs and heat-recovery systems.

There are basically two types of hoods: open and closed vapor hoods. In the open type, only a roof panel is placed on the top of the dryer bank. This panel can be either insulated or uninsulated. Open hoods are only suitable for limited production capacity paper machines. Modern paper machines are usually built with a fully closed vapor hood. The supply air can be heated to elevated temperatures so that the dew point, and thus the vapor-intake capacity of air, increases. The air can be heated to 180°C indirectly in the heat exchanger by saturated steam or directly to 450°C with gas or oil burners. Modern air hoods allow faster drying rates and a more uniform drying profile owing to the high dew point and efficient means of pocket ventilation. The most common dew point for these hoods is about 55 to 60°C. However, these high temperatures require additional insulation of the hood walls. A well-designed heat-recovery unit, however, permits an amortization in less than 2 y. Closed hoods operate on the principle of two-stage recirculation. The function of a heat-recovery system with a closed hood is shown in Figure 35.8. The fresh air supplied to the machine room is exhausted through stage 1 of the heat-exchanger system, where it is heated for the ventilation of the hood. If required, as during winter, the room-supply air can be preheated in stage 2. A comparison between the heat requirements of an open hood ventilated by convection and a totally enclosed unit with heat-recovery system is shown in Figure 35.9. With the open hood ventilated by convection (Figure 35.9a), the loss of heat in the exhaust air is denoted as 100%. With the totally enclosed hood (Figure 35.9b), the heat requirements of the drying steam [1] are substantially lower from the outset. The fluctuation of figures for unused waste heat [3] is attributable to the difference in the heat required for machine-room heating between winter and summer.

The space-bounded dryer cylinder, felt roll, and the sheet approaching and leaving the cylinder, as illustrated in Figure 35.10, are called the pocket.

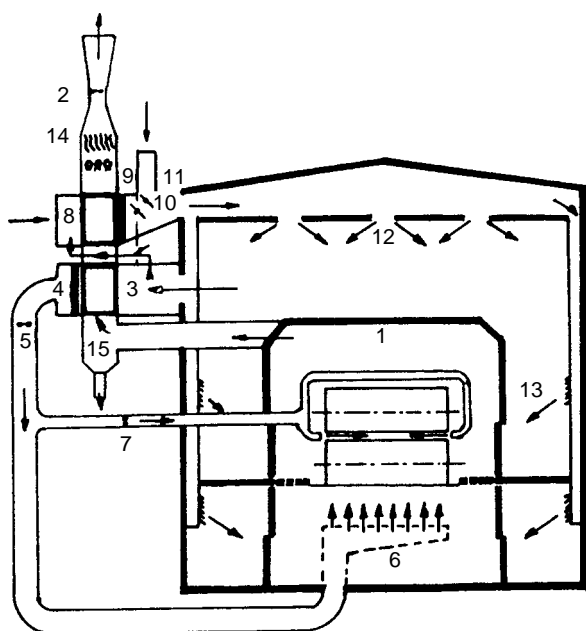


FIGURE 35.8 Function of a heat-recovery unit with closed vapor hood (1, vapor hood; 2, exhaust fan; 3, supply air heat exchanger; 4, supply air booster heater; 5, fan for supply air; 6, air distributor; 7, pocket ventilation; 8, room air heat exchanger; 9, room air booster heater; 10, temperature-control flaps; 11, room-supply air fan; 12, ceiling air distributor; 13, air outlet louvers; 14, warm-water unit; 15, warm-water discharge). (Courtesy of J.M. Voith GmbH.)

The moist air trapped in these pockets retards the drying of the sheet in the draws and causes nonuniform moisture distribution across the width of the sheet. However, conditions in the pockets can be

improved by introducing hot, dry air evenly across the width. Pocket ventilation systems can increase the drying efficiency up to 20% depending upon the previous conditions [16].

There are different designs to achieve ventilation in the pockets. Low-pressure crossblow pipes, arranged alternately on tender and drive sides, induce crossflow of air through the pockets. Low-pressure blow pipes, extending across the full width, have perforated nozzles that introduce hot air into the pockets at low velocities. These pipes are commonly used in low-speed machines. The Grewin system, which is a high-pressure blowing unit, is positioned similarly to blow pipes. The Grewin system uses a comparatively small quantity of hot air. However, because of its injection principle—alternately arranged nozzles—this system provides effective crossventilation and is applicable for heavier-basis weight products. Hot-air blowing rolls are also used for pocket ventilation purposes. Hot air is introduced into the pocket through the perforated shell of the felt roll. This system requires sufficiently permeable felts and is the most effective equipment for this application. However, these rolls are also the most expensive equipment for ventilation.

The application of very high-permeability dryer fabrics in the paper industry made it possible to introduce air through the fabrics. Therefore, the above-mentioned blowing rolls and the hot air ducts below (or above) the felt rolls between the cylinders became very popular. These systems provide better crossflow ventilation, which in turn results in more uniform and higher mass-transfer rates.

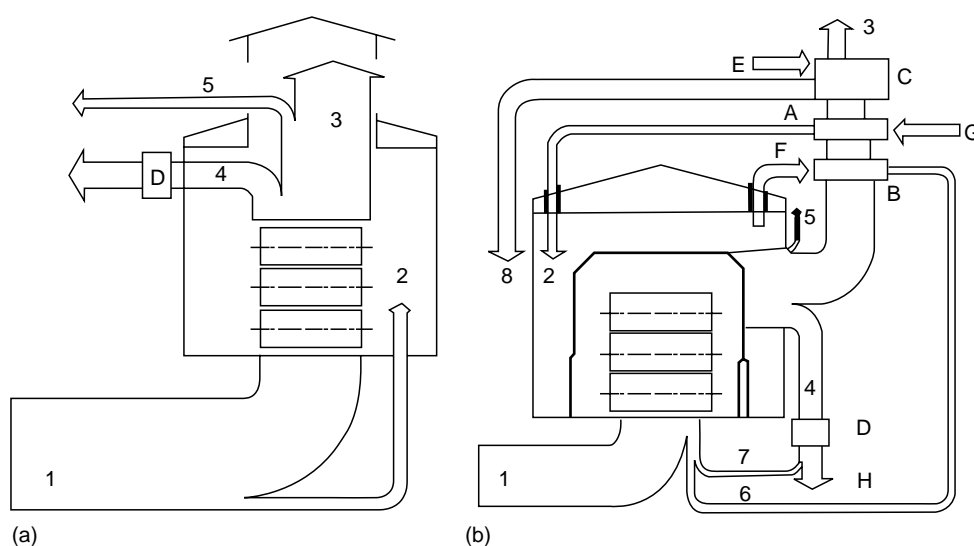


FIGURE 35.9 Heat-flow diagram of (a) an open hood ventilated by convection; (b) a unit with totally enclosed vapor hood. (Courtesy of J.M. Voith GmbH.)

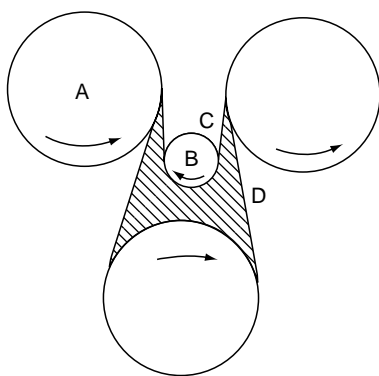


FIGURE 35.10 Dryer section pocket (A, paper dryer; B, felt roll; C, felt; D, paper).

A more detailed discussion on air- and heat-recovery systems can be found in the literature [4,5,17–20].

35.2.2.1.4 Dryer Felts and Fabrics

Dryer clothings are normally used for cylinder dryers only. The term fabric represents the highly permeable structure of synthetic materials, whereas by felt is understood a woven, comparatively low-permeability structure of both natural and synthetic fibers.

The basic function of dryer felts or fabrics is to improve the contact between the cylinder and the web by reducing the air gap between them. Other advantages of using drying felts can be summarized as: (1) prevent cockling, (2) control shrinkage of the web, and (3) support and guide the sheet [21]. The conventional dryer felts are very costly, however, and hinder mass transfer and pocket ventilation. On the other hand, felt tension variations, wet streaks, felt seam, and other problems also affect product quality. After high-permeability synthetic fabrics were introduced to the industry, most of these drawbacks were overcome. However, the use of high-permeability felts caused sheet-flutter problems on high-speed machines. This problem is solved by repositioning the lead rolls and by using less-permeable fabrics. The recent introduction of serpentine (or slalom) wire, which is an endless fabric and eliminates the bottom fabric, to the industry made it possible to run the newsprint machines at speeds higher than 20 m/s without flutter in the sheet and with no change in the drying capacity [22].

The advantages of dryer fabrics can be summarized [23] as: (1) increased drying capacity, (2) uniform moisture profile, (3) improved runnability and easier cleaning, (4) increased running life, (5) no fiber shedding, and (6) elimination of felt dryers.

35.2.2.1.5 Problems in Cylinder Dryers

At the beginning of the drying process, the fibers are free to slide over one another; but as moisture is removed they tend to come together, and at the end fiber–fiber bonding takes place. Then the sheet tends to shrink. The restraint owing to dryer felt and the tension due to draw tend to prevent shrinkage. This restraint results in a high tensile strength and very little ability to stretch. Otherwise, the net effect of those two forces is to reduce the bursting strength and basis weight and to increase compactness [24]. A very good discussion of the control of shrinkage has been presented by Nuttall [25].

The most common quality problem in papermaking is the nonuniform cross-machine (CM) moisture profiles. The traditional approach taken for wet streaks is to overdry the sheet. However, a 1982 study [26] showed that overdrying from 6 to 4% reel moisture for a typical 800 t/d capacity linerboard machine results in \$189,000 additional steam cost and 11,200 t production loss per year because of the reduction in machine speed. Uniform CM moisture profile is obtained by improving dryer pocket ventilation systems and using supplemental drying, such as air impingement or radiant drying.

Cockling of paper results mainly from imperfect sheet formation. The difference in felt tension may also cause cockling and uneven drying. Uniformity of moisture content and basis weight in CM direction prevents cockling.

Another common dryer-related defect is curl, which results mainly from the differences in stress–strain characteristics through the thickness of the sheet. However, moisture distribution through the sheet and the drying process may also cause the sheet to curl.

Low sheet tension, excessive sheet flutter, loose draws, and misaligned, cold, or undersize (due to wear resulting from dryer doctors) dryers caused dryer wrinkles. Other than these problems there are various defects related to dryer section, such as blistering due to loose dryer felt or a hot or dirty cylinder; linting (accumulation of solid particles); and dryer felt marks, especially felt seam marks.

35.2.2.2 Air Drying

In the early days of papermaking, the only means of paper drying was natural convection—air—and this practice continued until the advent of cylinder dryers. Today, the old loft, festoon, barber, and tunnel dryers are obsolete (or used in a few cases in which certain special low-production paper is made) because of their very low drying rates. However, it is very difficult

to operate conventional cylinder dryer sections at speeds much higher than 1000 m/min. Therefore, an alternative method of drying must be found to keep pace with increasing machine speeds. Currently, one viable alternative dryer is an air dryer. By combining impingement and through-drying methods, a dryer can be built to provide, at acceptable cost, not only high-speed operation and high drying rates at all sheet moisture contents but also safe sheet handling, improved paper quality, flexibility, convenient process control and profiling, and comfortable working conditions [27]. Of course, through-drying can be used for permeable grades only.

35.2.2.2.1 Impingement Drying

Early types of air dryers were all designed to operate at low air velocities and relatively low air temperatures. The drying rates were quite low for those dryers. The application of high-temperature, high-velocity air jets impinging on the wet web results in very high drying rates. The average heat and mass transfer under single or multiple jets can be represented functionally by

$$\overline{Nu}(\overline{Sh}) = C_1 Re^m Pr^n (Sc^n) H^r D_p^s (X_p^s) \quad (35.10)$$

for single jets and by

$$\overline{Nu}(\overline{Sh}) = C_2 Re^a Pr^b (Sc^b) H^c f^d \quad (35.11)$$

for jet arrays. An extensive summary of the correlations proposed for the prediction of transfer rates

under various turbulent impinging jet configurations can be found in Obot et al. [28] and in the relevant parts of this handbook.

Typical high-temperature, high-velocity air hoods used in industry are shown in Figure 35.11. The air-flow pattern is also indicated. A variety of heat sources, e.g., indirect heating by steam or oil, natural gas, or direct firing, can be applied to those hoods depending upon the temperatures sought. The air velocities for the modern high-velocity impingement hoods range between 60 and 130 m/s. In practice, jet velocities around 100 m/s are commonly employed. Air temperatures for those hoods range from 150 to 540°C, but 300°C seems to be the most commonly used value.

A schematic showing the drying capacities attainable at different temperatures and jet velocities is shown in Figure 35.12. The graphs denoting the specific energy requirements (referred to the effective dryer area) show that the energy demand rises rapidly with increasing air velocity and decreasing temperature. The diagram applies to an initial web dryness of 35%; when the web is simultaneously dried by contact drying, the specific drying capacity of the contact dryer must be added. An example for a desired specific drying capacity of 89 kg H₂O/m² h at 420°C air temperature is shown in this figure. The diagram gives approximately 90 m/s air velocity and 6.5 kW/m² as the specific power requirement.

The heat-transfer coefficient as a function of jet velocity, the pressure drop at the nozzle, and the power requirement are the key parameters in the design of impingement dryers. These parameters are

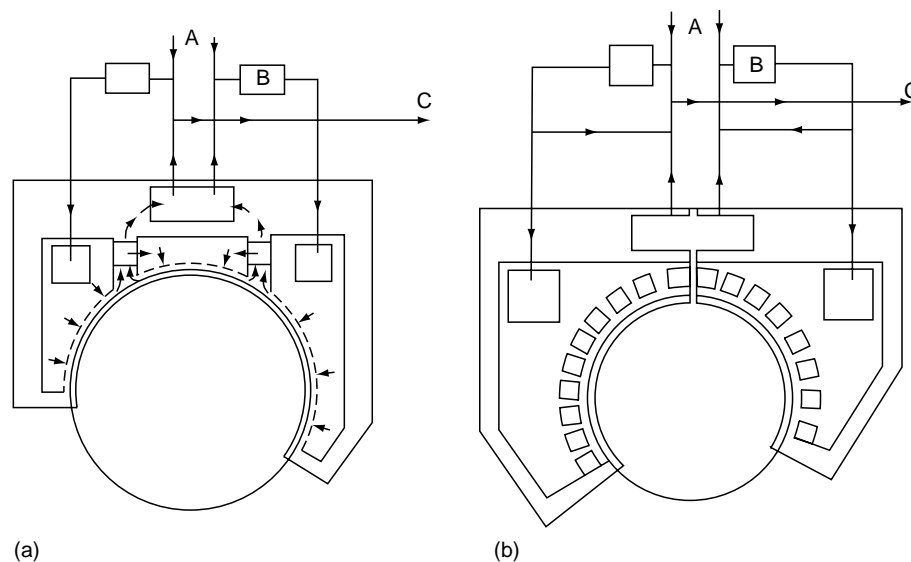


FIGURE 35.11 High-capacity hot air dryer hoods: (a) MG hood; (b) tissue hood (A, fresh air; B, heater; C, to heat recovery).

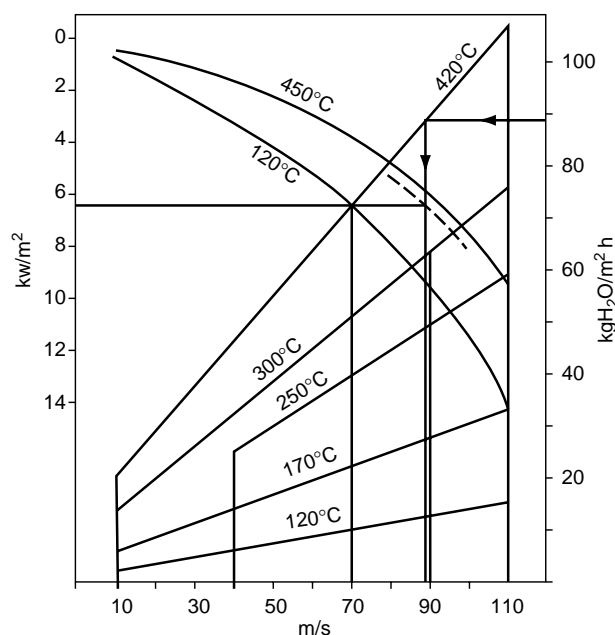


FIGURE 35.12 The drying capacities and specific energy requirements of high-velocity hoods. (Courtesy of J.M. Voith GmbH.)

influenced by various geometric parameters (e.g., arrangements of nozzles and exhaust ports, fractional open areas, nozzle shape and dimensions, nozzle-surface spacing, and others). For the calculation of the heat-transfer coefficients under various turbulent impinging jet configurations, the reader may refer to available reviews [28,29]. In a recent review of impingement drying, Polat has given a procedure for optimum design of such hoods [30]. There is also a simple and useful procedure for the design of convective dryers (impingement, through-, and floater dryers) for best efficiency and productivity [31].

Impingement air drying has found only limited application, either to increase the capacity or to control the moisture profile, in conventional cylinder drying systems. In such applications, the particular dryer cylinder is left unfelted. However, impingement dryers are widely accepted in two important fields, drying of coatings and drying of tissue and machine-glazed papers in conjunction with Yankee and MG cylinders.

A typical multizone impingement dryer for coatings is shown in Figure 35.13. In this particular dryer, supporting air decks are used to convey coated paper webs, which are dried by high-velocity impingement hoods. Different arrangements can be used for on- or off-machine drying of coatings. In some cases, a radiant dryer, with an infrared radiation (IR) zone, is located to accomplish the initial drying, and a couple of impingement units, usually two to three zones with different air velocities, follow to carry out the drying process. The impingement systems offer not only considerable improvement in drying rate but also undisturbed drying for coatings.

35.2.2.2.2 Yankee and MG Drying

Yankee and MG dryers are large, 3 to 6.1 m (10 to 20 ft) in diameter, steam-heated cylinders that are used on tissue and toweling grades and on machine-glazed papers, respectively. The surface of a Yankee or MG cylinder has to be very smooth in order to have a shiny finish on MG papers and a uniform crepe for tissue products. The application of high-velocity, high-temperature air impingement on conventional Yankee cylinders doubles their drying capacity. Therefore, it is a common practice to use air impingement on Yankee dryers for modern tissue machines.

Typical tissue and MG paper machines with high-velocity impingement hoods are shown in Figure 35.14.

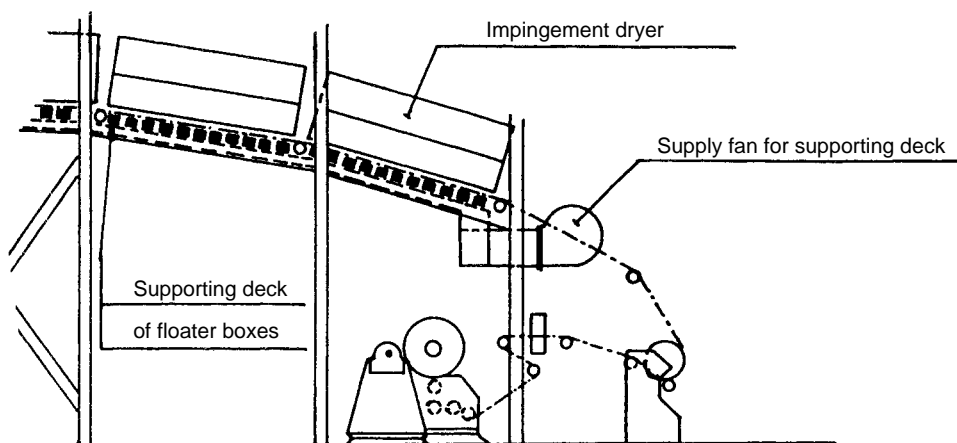


FIGURE 35.13 A multizone impingement dryer for coated board. (Courtesy of Flakt Ind. AB.)

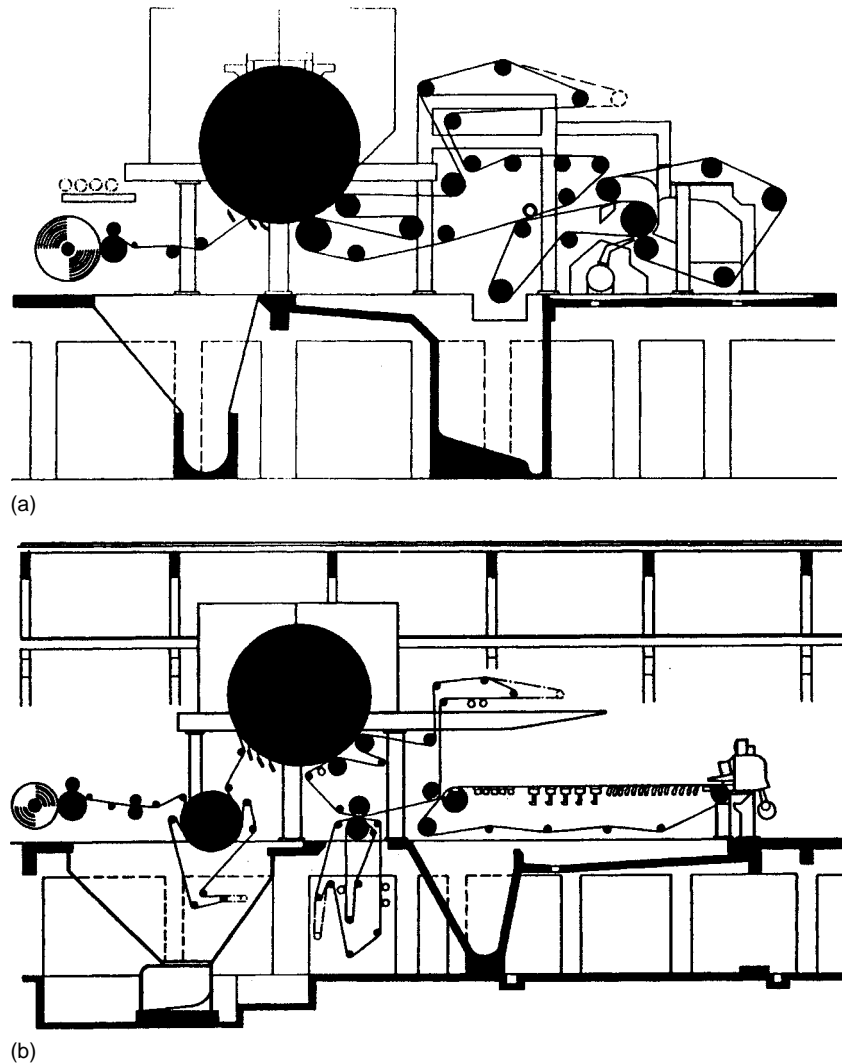


FIGURE 35.14 (a) Tissue machine with a creping dryer; (b) Yankee dryer in machine for MG papers. (Courtesy of J.M. Voith GmbH.)

35.2.2.2.3 Through-Drying

Through-drying is a process of drying permeable webs by the percolation of hot air through its mass. As the heat- and mass-transfer areas are increased owing to the intimate contact of flowing air with the fiber surface, much higher drying rates than those achieved by conventional methods can be obtained for sufficiently permeable products (including nonwovens).

The use of the through-drying technique for porous grades (e.g., tissue, toweling, filter, blotting, and nonwovens) has increased owing to (1) the production of soft, high-bulk products; (2) the extremely high forming speeds achieved by today's twin-wire formers, outstripping the capacity of the largest Yankee dryers; (3) manufacturing advantages of single-ply products; (4) little or no press dewatering needed

in through-drying; and (5) better, increased productivity coupled with high thermal performance of through-dryers [32].

Although through-drying has become a very effective method of drying permeable grades in recent years, there is no generally applicable approach for the prediction of the heat- and mass-transfer rates. It is known that there is a minimum pressure difference (threshold pressure, which can be related to the surface tension forces) below which no flow occurs [33]. Once this pressure is reached, the through-drying process commences. Some authors suggest that the air exiting the sheet is at the saturation temperature, which is valid for sufficiently thick sheets or at low through-airflow rates [27,34,35]. Otherwise, it overpredicts the actual drying rates. Recent works on the transport phenomena analysis [36], on modeling [37],

and reviews [38] of both science and technology of through-drying of paper have brought more understanding of this relatively new technique, but the present knowledge of the basic transport processes is still far from complete.

As it is still a relatively new technology, there is no well-established through-drying system design procedure. The existing applications can be classified into two basic configurations: rotary through-dryers and flat-bed dryers. The rotary type has an open (perforated) roll to allow the passage of air and a hood arrangement to provide the pressure differential required. Two methods of forcing air through the web have been used for cylinder-type through-dryers. The first method is to use a pressurized roll and apply the hot air through the roll and the web and out to a chamber. In the second case, the roll is under vacuum and the hot drying air is supplied from a pressure hood that covers the drum. The air is then pulled through the web and roll by vacuum. There is no need to use dryer felt for the second method, as the sheet is held tightly by vacuum. However, for the pressurized-roll case a very open fabric is necessary to hold the sheet onto the cylinder.

In flatbed design, there are a fabric support and conveying structure and top and bottom air chambers. The supply air can be applied through either chamber depending upon the design and process conditions. This flatbed arrangement is particularly suited to highly permeable products in which the pressure differential requirements are relatively low and do not create significant mechanical problems between the support members [39].

Through-drying is used in tissue drying because of its very high permeability, and very high drying rates (which exceed those on Yankee cylinders with high-velocity hoods) are achieved even at fairly low temperatures. It is not recommended for newsprint, especially at the wet end, where the permeability is relatively low [27]. However, a recent application of a high-vacuum through-drying unit in a roofing-felt (which has low permeability) mill is claimed to have resulted in a 40% production increase plus a 20% reduction in energy use per ton when compared with the previously used natural-gas-fired steam cylinders [40].

The high through-airflow rates for permeable sheets make possible high drying rates, at the same time virtually eliminating “one-sidedness” and CM variations in the product. The thermal efficiencies of through-dryers were increased up to 80% with the use of several drying stages and air recycling or cascade systems. Another potential advantage is the possibility of operating through-dryers at low temperatures by using low-grade waste energy [41]. Honeycomb vacuum cylinders of diameters up to 6.71 m (22 ft)

and widths up to 7.6 m (25 ft) operate at speeds up to 30 m/s and at air temperatures 370 to 430°C (700 to 800°F) with high air-recirculation ratios (e.g., 80%).

Recently published experimental data [42,43] and calculations [44] suggest that strategically placed through-dryers may even be used for higher basis weight, lower air permeability grades to improve drying rates. This process basically consists of a cylinder dryer section followed by a through-dryer. Use of through-air drying at lower sheet moistures not only improves the drying rates but also reduces the fan power requirement due to increased sheet permeability.

35.2.2.2.4 Airborne Drying

Use of air floater dryers is recommended especially for the drying of paper products for which unrestrained drying is desired.

In air floater dryers, the web floats freely while drying by the introduction of a contact-free guide plane. In practice, the guide plane is designed as a blowing deck, where air is supplied through blow boxes with specially designed perforations. Various dryer designs are shown in Figure 35.15. The one-sided dryer design is particularly useful for paper and board coated on one side. The blow boxes for perpendicular impingement are positioned facing the coated side and for slanting impingement to the untreated side. The uncoated side of the web can be treated with either warm or cold air, depending upon the process conditions. In the two-sided dryer, horizontal and vertical air impingement alternate on both sides of the web. This design affords a higher drying capacity and is suitable for booster drying of different kinds of paper and board, including drying of sack Kraft paper, liner, and board. The printing dryer is a special version of the one-sided dryer. The lower blow boxes use cold air; the upper ones use hot air. The noncontact dryer uses both vertical and angular air impingement in the same blow boxes and is suitable for drying two-sided coated webs, impregnated web materials, and so on.

The drying decks are stacked on top of each other, and the web is guided through the dryer in a multiple-pass fashion as shown in Figure 35.16. Each turning roll is driven individually, and the speed is controlled to conform to the machine-direction shrinkage of the web.

The internal ducting of a floater dryer is shown in Figure 35.17. The drying air is fed through horizontal ducts to the flow boxes and blown onto the sheet through the perforations in the boxes, then discharged through slots between the blow boxes. The air is recirculated inside the dryer by a series of fans along both sides of the dryer; only a small portion of the recycled air is withdrawn and replaced by fresh

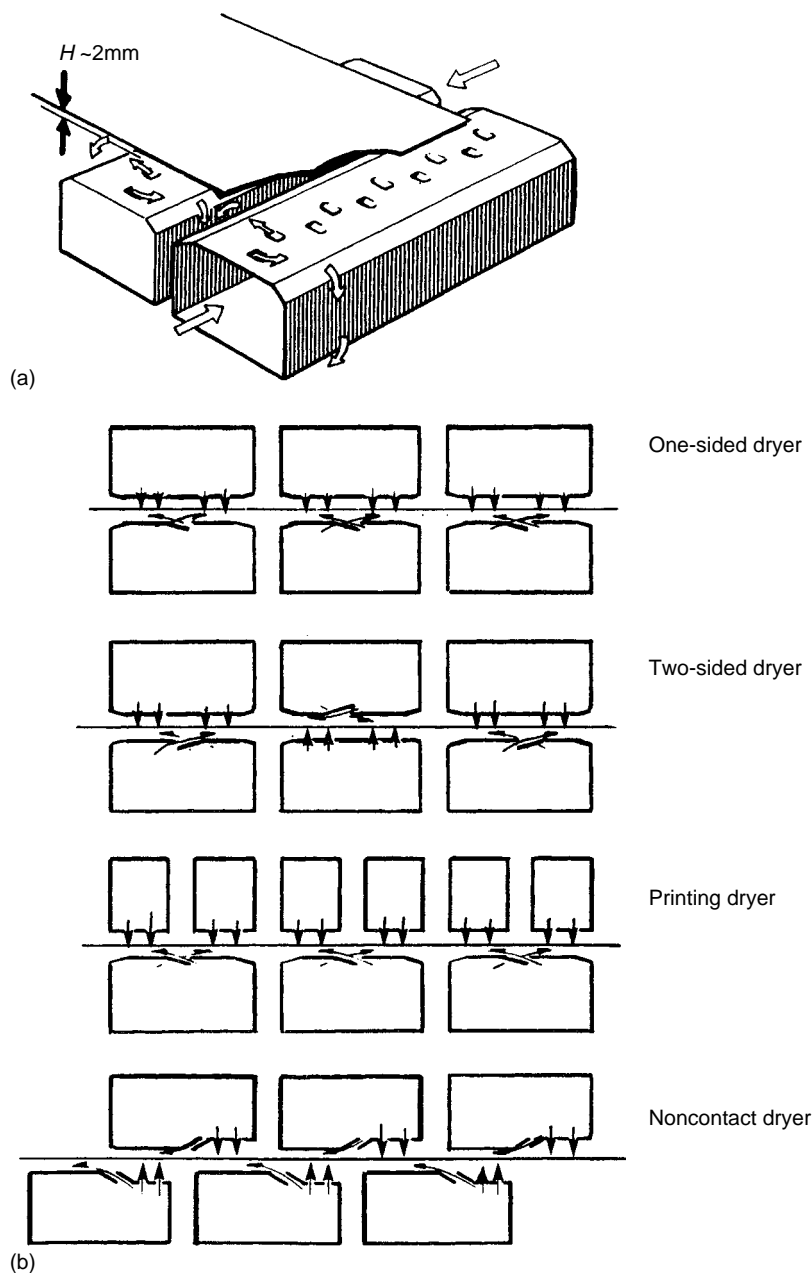


FIGURE 35.15 Air floater dryers: (a) blow-box arrangement; (b) various blow-box designs. (Courtesy of Flakt Ind. AB.)

air. The air velocity range normally used is 25 to 50 m/s. The air can be heated either by direct firing or by steam coils.

The application of floater dryers allows the sheet to be dried virtually tension-free in the CM direction with only slight tension in the machine direction. It is well known that drying under zero or moderate tension in the shrinkage range gives higher stretch and tensile energy absorption (see Figure 35.18) to the product whereas the modulus of elasticity and stiffness decreases.

A typical application of a floater dryer on a sack paper unit is shown in Figure 35.19. As natural shrinkage occurs principally between 50 and 85% dryness, the floater dryer is positioned between the predryer and the after-dryer in order to produce high-stretch paper. The after-dryer is used to remove the cockles and to give a smooth surface to the paper. If desired, a smoother surface can be obtained by calendering the paper.

Air floater (or airfoil) dryers have become well established for drying coated papers and sack paper,

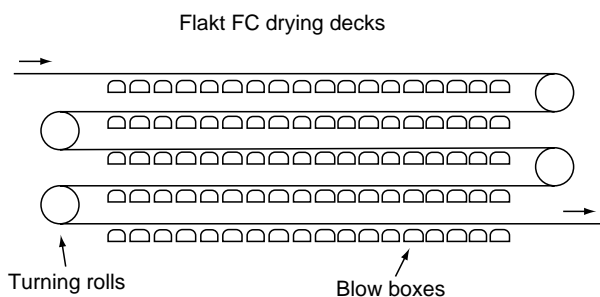


FIGURE 35.16 Stacked drying decks. (Courtesy of Flakt Ind. AB.)

for example, and operating at speeds greater than 10 m/s (2000 fpm).

The advantages claimed, other than those already mentioned, are high drying rates, uniformity of moisture profile across width, and favorable energy efficiency [45].

35.2.2.3 Radiant Drying

Radiant drying has not been used in the papermaking industry very extensively. Relatively recent applications of infrared dryers have been limited to special applications, such as drying of coatings, in which contact-free drying is a valuable asset. However, the applications of high-frequency dielectric heating, i.e., RF and microwave dryers, are still at the preliminary stage. The main drawback of radiant dryers is their relatively high operating costs.

35.2.2.3.1 Principles of Radiant Heating

The transfer of heat by radiation is expressed by

$$q_r = \sigma A_1 F_{1-2} (T_1^4 - T_2^4) \quad (35.12)$$

F_{1-2} in this equation is called the overall interchange factor and is a function of both the geometric parameters, such as view factors, and the emissivities of the radiating and receiving surfaces. As the temperature driving force is the difference between the fourth power of the absolute temperatures of the radiating and receiving surfaces, very high heat-transfer rates can be obtained by radiant dryers, given higher emissivities and optimal design of the system for better view factors.

The frequency and wavelength bands of commercially available radiant dryers are shown in Table 35.4. This table gives broad ranges for the dryers; but in practice almost all IR dryers are in near-infrared range (1 to 10 μm), and the specific frequencies for RF and microwave dryers are allocated by international agreement, such as 13.56, 27.12, and 40.68 MHz for RF and 433.92, 896, 915, and 2450 MHz in the microwave region.

35.2.2.3.2 Infrared Dryers

Infrared dryers consist simply of a bank of infrared heaters combined side by side or end to end. These heaters are arranged to provide high-density heating of the sheet by the geometry of the oven and the optically designed reflectors. There are basically two types of infrared heaters: electric infrared and gas

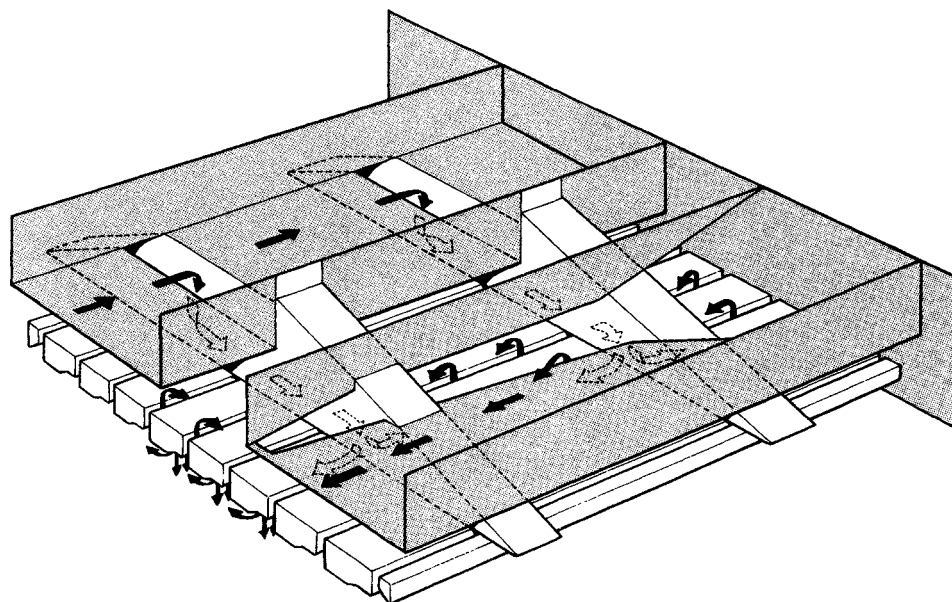


FIGURE 35.17 Internal ducting of a floater dryer. (Courtesy of Flakt Ind. AB.)

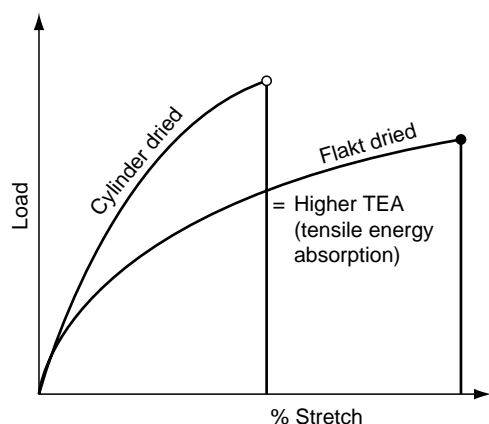


FIGURE 35.18 Stretch of paper depending on drying conditions. (Courtesy of Flakt Ind. AB.)

infrared. The electrical infrared heater is primarily a metal filament in a sealed enclosure. The spectral energy distribution of various infrared heat sources is shown in Figure 35.20. In addition to high radiant efficiency, the higher source temperature generates shorter wavelength radiation. The shorter wavelengths are more penetrating and more sensitive to color differential absorption. There will be a greater difference between black-and-white products in a high temperature-source oven. However, the greater radiant efficiency in the high-temperature ovens compensates for this drawback, as shown in Figure 35.21. The characteristics of commercially used electric infrared heat sources are summarized in Table 35.5, which shows that in some applications convection is a very important factor for heat transfer. Therefore, the overall oven design that allows better circulation of air, of which the primary function is to carry out the evaporated moisture, becomes very

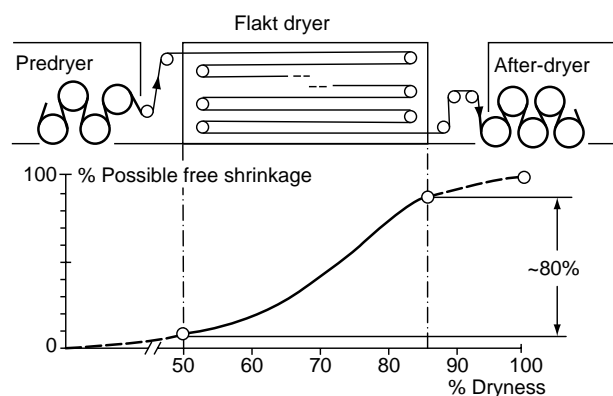


FIGURE 35.19 The position of a floater dryer in a sack paper dryer unit. (Courtesy of Flakt Ind. AB.)

TABLE 35.4
Frequency and Wavelength Ranges of
Radioactive Dryers

Dryer Type	Wavelength Range		Frequency Range (MHz)	
	Overall	Typical ^a	Overall	Typical ^a
Infrared	1–1000 μm	1–10 μm		10^7
Microwave	1–100 cm	10–70 cm	400–5000	900–2450
Radio frequency	1–100 m	7–22 m	1–100	10–40

^aCommonly used ranges in dryer designs.

important. The various types of oven design are shown in Figure 35.22.

Gas infrared burners for industrial processing are usually in two general styles: surface (Schwank) burners and impingement burners. A surface burner (Figure 35.23) is a gas generator consisting of a perforated special ceramic tile, set in a rugged cast iron or ceramic housing, with a special alloy screen grid that protects the tile and also serves as a reradiator and air deflector. These burners have an operating temperature range from 760 to 890°C (1400 to 1650°F) when the proper mixture and volume of gas and air are supplied. The fuel mixture is ignited on the surface of the ceramic tile. Higher temperatures (~1550°C) can be achieved by using refractory IR burners.

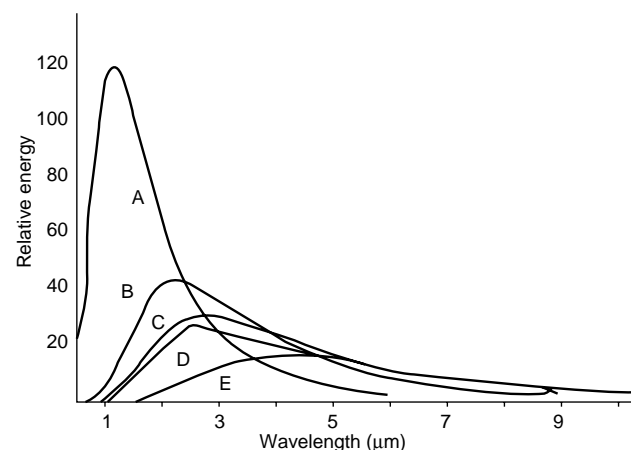


FIGURE 35.20 Spectral energy distribution of infrared heat sources (A, quartz lamp coiled tungsten, 2200°C; B, quartz tube coiled nickel chrome alloy, 1980°C; C, metal sheath heater surface, 760°C; D, Schwank gas infrared burner, 900°C; E, electric panel heater, 430°C). (Courtesy of Fostoria Ind. Inc.)

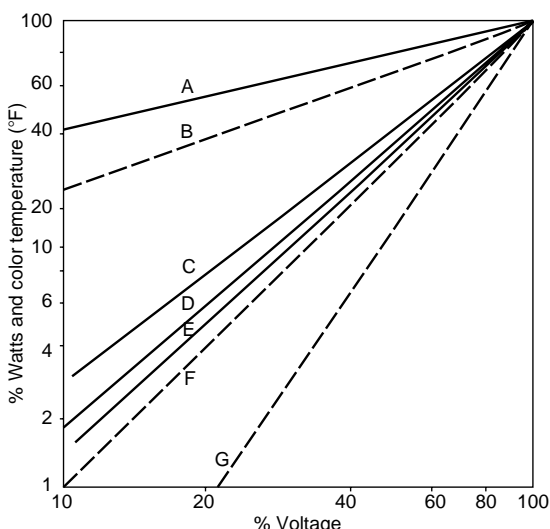


FIGURE 35.21 Voltage characteristics of infrared sources (A, tungsten color temperature; B, nickel chrome color temperature; C, 2200°C tungsten wattage; D, quartz tungsten radiation output; E, G-30 glass tungsten radiation output; F, nickel chromium wattage; G, nickel chromium radiation output). (Courtesy of Fostoria Ind. Inc.)

A newly developed, ceramic, fiber matrix, gas-fired infrared generator [46] is claimed to have a high gas-infrared conversion efficiency (50 to 70%) and high power density (30 to 100 kW/m² depending on temperature); it can be turned off from 850°C to touch in seconds; it can be rapidly modulated to meet heat load or machine speed variation; and it has no naked flame. The ceramic fiber pad of this generator is robust, and the supporting chamber and frame are air cooled.

The advantages of IR dryers can be summarized as: (1) low capital cost, (2) compactness, (3) contact-free drying, (4) instantaneous startup and shutdown of electric infrared, and (5) better product quality because of the possibility for zoning the electric heaters across the width of the web and modulating the heat in each zone separately. However, application of the IR dryers is still limited to the drying of coatings, primarily owing to the relatively high operating costs, i.e., the cost of electrical energy or gas. There are also safety problems related to IR dryers, especially with gas burners, because of the very high operating temperatures.

35.2.2.4 Recent Developments in Paper Drying

In this section, the more significant developments made to increase the efficiency of multicylinder dryers

in recent years, and the new or novel concepts not used extensively in the papermaking industry yet, will be summarized. Some of these methods are still at a laboratory stage but show the potential for a real breakthrough in the paper-drying technology.

35.2.2.4.1 Innovations in Conventional Drying of Paper

Cylinder-end insulation: The heat loss from a single cylinder (1.5 m in diameter) over a period of a year is found to be equivalent to 86 t of steam. Further calculation shows that 90% of the heat loss can be saved by the application of an insulator with an overall heat-transfer coefficient of 0.7 W/m² K to the cylinder ends. However, a practical insulator should be fire, oil, water, and chemical resistant, with the characteristics such as easy to install, cost-effective, have a long life, and should also provide access to manholes in order to solve this problem.

35.2.2.4.2 Ribbed Dryers, Spoiler Bars

Augmentation of heat transfer through the condensate layer becomes more important with increasing machine speed. In order to obtain higher heat transfer between the condensing steam and the wet web, the thickness of the condensate layer should be reduced by more efficient condensate-removal techniques or by the use of fins so that heat can be conducted around the condensate. This has led to such concepts as ribbed dryers, spoiler bars, and grooved dryers.

The ribbed dryer has a series of ribs or fins machined on the inside cylinder wall. The condensate forms in grooves and is removed by a series of small syphon pipes. The application of this type of dryer is limited (only on Yankee cylinders) due to the complexity of the condensate-removal system.

There is an oscillatory motion in the rimming condensate layer and, if some restrictions (e.g., bars) are spaced at the resonant frequency of the condensate, a higher intensity of turbulence can be achieved. This is the idea behind the dryers with spoiler bars. There are two ways to attach the bars inside the cylinder: (1) magnetic bars that utilize 460- to 920-mm bar magnets laid end to end and (2) spring-loaded hoop rings that use circumferential rings attached to the longitudinal bars.

It has been reported [47] that the overall heat-transfer coefficient can be increased up to 40 to 50% by ribbed dryers or spoiler bars at machine speeds of 1400 m/min, as compared with plain shell dryers.

35.2.2.4.3 Profilers

Profiling steam boxes and infrared and magnetic profilers are some of the devices utilized in paper machines.

TABLE 35.5
Characteristics of Commercially Used Infrared Heat Sources

Characteristics	Tungsten Filament Wire		Nickel Chrome Alloy Spiral Winding		Low-Temperature Panel Heater, Buried Nickel Chrome Alloy, Metallic Salt
	Glass Bulb	T3 Quartz Lamp	Quartz Tube	Metal Sheath	
Source temperature (°C)					
Normal maximum	2200	2200	870	650	315–430
Usual range	1650–2200	1650–2200	760–980	540–760	205–590
Brightness		Bright white heat	Cherry red	Dull red	Nonvisible light
Usual size	G-30 lamp	3/8-in. diameter tube	3/8- or 5/8-in. diameter tube		Various flat panels
Wavelength at energy peak (μm)					
Normal maximum	1.15	1.15	2.6	3.1	~4–5
Usual range	1.15–1.5	1.15–1.5	2.6–2.8	2.8–3.6	3.2–6
Relative energy distribution (%)					
Normal maximum					
Radiation	80	86	55	50 ^a	40–30 ^a
Convection and conduction	20	14	45	50 ^a	60–70 ^a
Usual range					
Radiation	65–80	72–86	55–45	53–45 ^a	50–20 ^a
Convection and conduction	35–20	28–14	45–55	47–55 ^a	50–80 ^a
Degree of heat penetration	Depth of penetration varies with the characteristics of the product; as a general rule, energy of shorter wavelengths penetrates deeper than energy of longer wavelengths				
Relative response to heating up	Seconds	Seconds	Minutes	Minutes	Scores of minutes
Cool down	Seconds	Seconds	Seconds	Minutes	Scores of minutes
Color sensitivity	Bodies of different colors can be heated at more nearly the same rate by IR with long wavelengths than they can be short-wavelength IR				
Ruggedness					
Mechanical shock	Poor	Good	Good	Excellent	Varies with panel
Thermal shock	Poor	Excellent	Excellent	Excellent	Design could be quite good

^aRelative energy distribution will vary with the amount of convective cooling, which can vary with the position of heater and volume of air moving by, among other factors.

Source: Courtesy of Fostoria Ind. Inc.

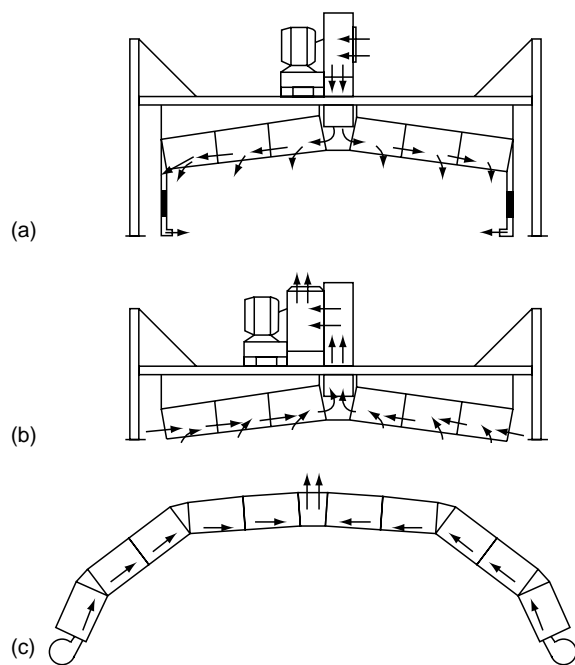


FIGURE 35.22 Infrared designs: (a) oven with pressurized sections to introduce heated air; (b) oven with pressurized sections for cooling and ventilation; (c) different design of an oven with pressurized sections for cooling and ventilation. (Courtesy of Fostoria Ind. Inc.)

Profiling steam boxes, which selectively heat the web, are placed just prior to the pressure-roll nip. The increased temperature of the web improves water removal, thus providing a means for moisture profile control. However, the width of profile control is typically 310 to 460 mm and the range of the moisture control is limited between 2 and 3% [48].

The infrared profiler uses segmented infrared heaters that are individually controlled in 150-mm zones in the CM direction.

The magnetic profiler uses independently excited electromagnets installed across the width of a dryer cylinder. As the cylinder, which is a conductor, passes through the magnetic flux lines, eddy currents (which produce heat) are induced in the shell. This heat provides very precise and discrete control of the mois-

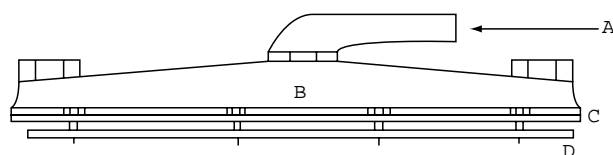


FIGURE 35.23 Surface-type gas infrared burner (A, premix gas and air; B, chamber; C, perforated ceramic tile; D, special alloy screen grid). (Courtesy of Fostoria Ind. Inc.)

ture profile. The eddy current heating produced can be 100 times higher than the input power to the inductor in the 150-mm width band. Commercial tests show that these profilers control the moisture within $\pm 0.3\%$ and increase the machine speed up to 5 to 10% [48].

35.2.2.4.4 Papridryer

The Papridryer consists of two principle components: a high-velocity hood and a vacuum cylinder. High velocity (60 to 100 m/s), hot (250 to 450°C) air jets impinge against the web supported on the vacuum roll, resulting in higher drying rates at the surface of the web. The vapor formed within the sheet is removed by suction. The application of suction provides not only through-drying but also enhanced impingement heat-transfer rates. Laboratory [49] and mill [50] trials of the Papridryer show very high drying rates (almost 10 times the average rate of modern conventional dryer section for newsprint) without significant change in quality. The preliminary design calculations showed that only six Papridryers, each having 1.5- to 2-m suction roll, could provide all the necessary drying for a 15-m/s newsprint machine with a dryer train consisting of about 60 steam-heated cylinders. A better CM moisture profile can be obtained by varying the jet velocities. Fast response to control action and better energy efficiency are among the other advantages of the Papridryer. Recent experimental studies published on the effect of surface motion on slot jet impingement heat transfer [51] and the effect of high-temperature difference (up to 300°C) on impingement heat transfer [52] will make it possible to improve the Papridryer design.

The need for higher drying rates and controlled paper properties has revived the Papridryer, i.e., combined impingement and through-air drying, idea in the last decade. Recent experimental studies [53,54] reaffirmed the drying benefits, but still there is no commercial application.

35.2.2.4.5 High-Intensity Impingement Drying

High-intensity drying is a concept using high-velocity impingement drying in combination with cylinder drying. It is similar to Yankee dryers on tissue machines, i.e., impingement hoods on a steam-heated cylinder. Very smooth paper surface is achieved due to the contact with the drying cylinder. Two dryers can be used for printing papers to ensure that both sides are treated in the same way [55].

The impingement-drying concept developed by Valmet comprises a combination of conventional cylinder dryers and high-intensity drying units. Each high-intensity drying unit has a large diameter roll and two retractable air-impingement hoods. Three

of such units can be installed in a dryer section for printing- and writing-paper grades. It is claimed that the specific energy consumption per ton of paper is almost exactly the same as the corresponding conventional dryer case [56,57]. It is expected that this type of dryer will play a significant role in developing compact, high-speed machines of the future [57,58].

35.2.2.4.5 *Radio Frequency and Microwave Drying*

Another alternative in maintaining higher drying rates is the application of high-frequency dielectric heating, covering both RF and microwave ranges. As shown in Table 35.4, microwaves have higher frequencies and shorter wavelengths whereas the RF radiations have lower frequencies and longer wavelengths.

An industrial microwave heating system consists of a dc power supply, a microwave generator (magnetrons are available in 915- and 2450-MHz bands and klystrons at the higher-frequency bands) and an applicator. A microwave heater has only one electrode. On the other hand, an RF device requires two electrodes. The power can only be drawn from an RF generator when there is material present in the applicator; therefore, the material is an essential electrical component of the circuit and affects the electrical characteristics [59].

In the early designs of the RF dryers, flat-plate electrodes are used and the product is placed between these electrodes. More recent designs use rod electrodes either in a staggered through-field arrangement in which the sheet is between the electrodes or in a stray-field arrangement in which the electrodes are on the same side of the sheet. These configurations have been applied successfully to paper drying.

As the heat is absorbed only by water and as the power-conversion efficiency is low, this mechanism is suitable only for profile correction and for the drying of laminated sheets.

A recently proposed system combines RF heating with hot-air impingement. RF energy is applied to the sheet through the air-impingement nozzles. Initial tests have shown that the same drying rate can be achieved up to one tenth of the time as that of the conventional dryer, for some products, by using only 10 to 20% of the energy required in the form of RF energy.

Jones [60] states that as the machine width increases to become a significant proportion of a quarter of a wavelength (i.e., 5 m at 13 to 56 Hz), the field uniformity becomes increasingly difficult to achieve. As paper machines are typically wider, an account must be taken of standing waves.

35.2.2.4.6 *Innovative Press-Drying Techniques*

In the last 20 y, the main research effort in paper drying has been concentrated on what we may call

the variations of wet pressing or a combination of pressing and drying. Press, impulse, and Condebelt drying techniques have made significant progress in recent years. Each of these processes uses mechanical, thermal, and interactive effects in a unique way; because of that they are sometimes called thermomechanical web consolidation processes [61]. All these techniques claim very high dewatering or drying rates that are applicable even to sheets made from high-yield pulps and offer opportunity for greater energy efficiency. Detailed, comparative reviews can be found elsewhere [61–64].

35.2.2.4.7 *Press Drying*

The FPL (Forest Product Laboratory, U.S. Department of Agriculture) press-drying process combines mechanical and thermal means of water removal for drying stiff pulp fibers (e.g., for linerboard production under compressive force that improves interfiber bonding). Press drying utilizes less energy because the sheet has lower moisture entering the drying section. Wet web is sandwiched between two felts and pressed between two hot surfaces.

Laboratory tests have shown that it is possible to achieve average drying rates about 10-fold at 177°C and 20-fold at 288°C compared with conventional drying of linerboard [65]. The estimated dryer size needed for paper-machine speeds ranging from 0.25 to 10 m/s (50 to 2000 fpm) based on the tests performed at FPL can be found in the literature [66].

The techniques developed for press drying and the effects of press drying, particularly changes in drying variables such as initial moisture content, temperature, pressure, and time, on sheet properties have recently been summarized by Mitchell [67].

Back and Anderson [68] have found that at a surface temperature of 150°C, both the tensile index and the modulus of elasticity remain at high levels in the range of 39 to 70% initial solids content, but decreases sharply at higher solids content. They used press-dried paper made from 60%-yield softwood Kraft pulp.

Both the studies of Back and Anderson [68] on the strength properties of press-dried sheets at the surface temperature of 150 to 300°C and Yang et al. [69] on the effect of density on modulus of elasticity of 107°C, 149°C, and 232°C showed no appreciable effect of change in temperature. Setterholm and Benson [70] showed that increased pressure increases the consolidation of the sheet, which is accompanied by increased strength properties such as breaking length and elastic modulus.

Some of the reported advantages of the press dryers are listed below [71]:

- Effective utilization of high-yield fibers
- Better use of hardwood fibers
- Improvement in the characteristics of paper products made from refiner and thermomechanical pulps
- Improvements in paper-containing waste paper that leads to a promising future for paper recycling
- Reduction in the amount of refining required to obtain the given web characteristics
- Improved dimensional stability and smooth-surface production
- Increase in the output of dryer-capacity-limited machines

Press drying offers a very promising method for paper and paperboard drying. However, the method is still in the development stage and many problems (i.e., the venting of water vapor and the need of extended nip residence time) remain to be solved. Pilot-scale testing, at PAPRICAN, PIRA, and so on, of this process is still underway.

35.2.2.4.8 Condebelt (or Convac) Dryer

The Convac process is an entirely novel concept in paper drying. In this process, the wet sheet is pressed on a steam-heated metal surface with a permeable felt or mat and an impermeable metal sheet that is water-cooled [72] (Figure 35.24). Before drying starts, air is removed from the web and felt by vacuum. Once drying begins, water vapor evaporates from the web, passes through the felt and condenses on the cold metal sheet, and this continues throughout the drying

process. Therefore, by using this mechanism, vacuum is maintained in the dryer. The static laboratory tests resulted in very high drying rates, exceeding $145 \text{ kg/m}^2 \text{ h}$ at a metal temperature of 170°C . Convac-dried paper and board products show higher stiffness and tensile strength, and the web surface in contact with the hot metal becomes smooth.

In the pilot-scale application of this process, the paper web is supported by a fine metal wire and a thick plastic wire and is fed between two metal bands (Condebelt process). The top or bottom band is heated by steam and the other band is cooled by water. Two proposed Condebelt arrangements are shown in Figure 35.25. Pilot-scale testing of this process at Valmet-Tampella Research Center gave promising results [73,74]. Drying rates close to $500 \text{ kg/m}^2 \text{ h}$ are reported for low-grade paper in these tests [73]. The advantages are claimed to be much higher drying rates, improvement in sheet properties (e.g., smoothness, enhanced strength properties, and no shrinkage), and the possibility of reduced space requirements.

There are two commercial installations of Condebelt dryers on board machines. According to recent reports [75,76], the minor operational issues have been resolved with modified design and expected improvements in product properties are realized.

35.2.2.4.9 Impulse (High-Intensity) Drying

High-intensity contact drying denotes the drying under sufficiently intensive heating conditions such that, following a brief warm-up period, the moist paper web operates at internal temperatures in excess

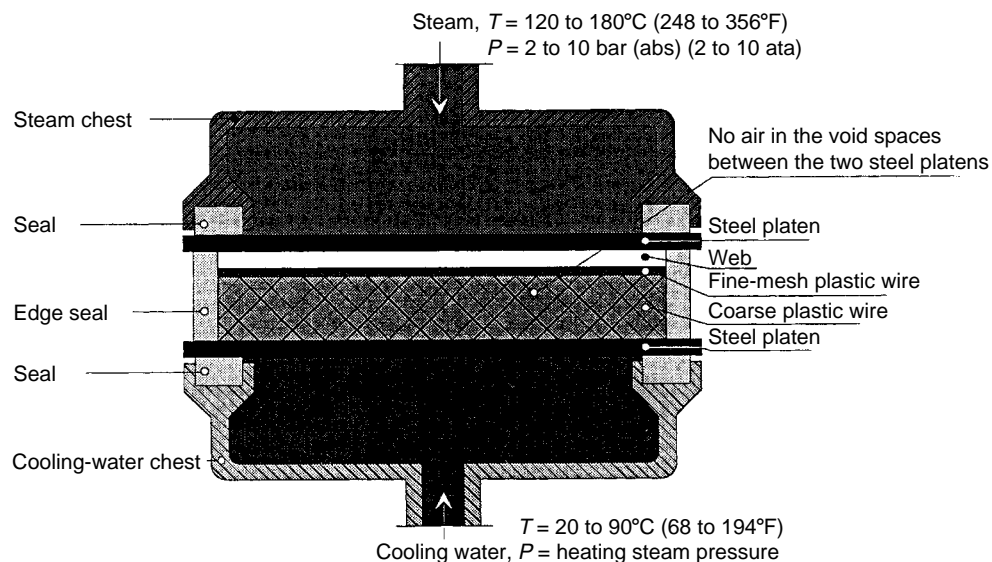


FIGURE 35.24 Basic scheme of a Condebelt device. (Courtesy of Valmet-Tampella.)

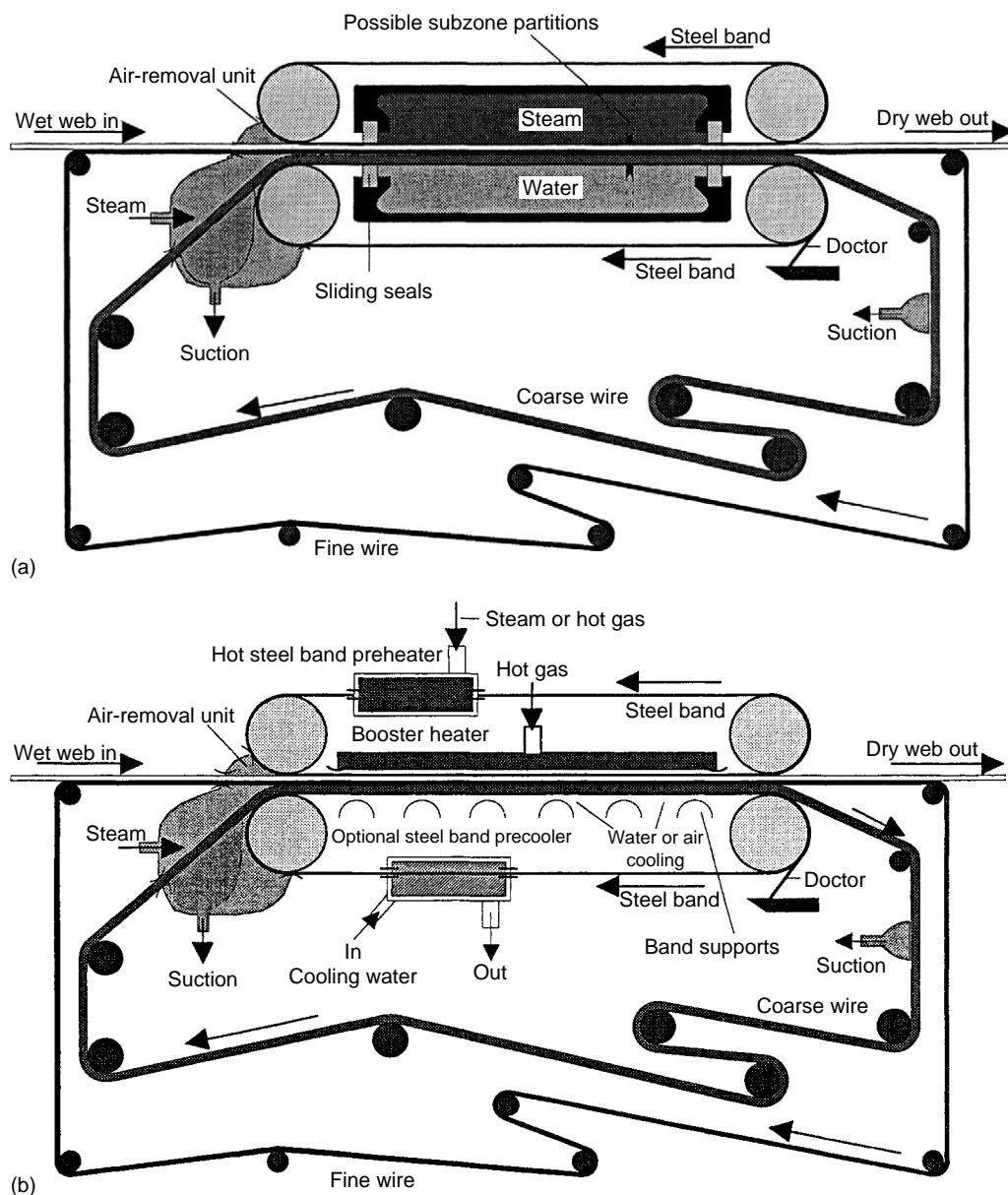


FIGURE 35.25 Condebelt dryer arrangements: (a) high z-pressure arrangement; (b) low z-pressure arrangement with pre- and booster-heated steel band. (Courtesy of Valmet-Tampella.)

of the ambient boiling point [77]. Dryer surface temperature might be elevated to 200°C or higher and sheet contact pressure in the range of 7 to 35 kPa or higher. These can be compared with the typical dryer surface temperatures, which range from 125 to 175°C and contact pressures that are less than 2 kPa. Drying rates may be 10 to 20 times those obtained in conventional drying [77].

Poirier and Sparkes [78] have successfully run a two-roll impulse drying unit at PAPRICAN pilot paper machine at speeds up to 800 m/min. Solids contents of 60% have been achieved with newsprint using this single-nip arrangement.

The main problems associated with this technique are delamination of the sheet, brightness and opacity losses, sidedness, and sticking of the sheet to the surfaces. There is a great investment in experimental and theoretical development work in order to gain a better understanding of the mechanisms involved in this highly promising process. Various theoretical views and debate published in the last 5 y show that a widely accepted mechanism has not been proposed yet [61–64,79–83]. Drying rates and process conditions for typical impulse drying process as compared with press and Condebelt drying are shown in [Table 35.6](#).

TABLE 35.6
Comparison of Press, Condebelt, and Impulse Drying

	Press Drying	Condebelt Drying	Impulse Drying
Temperature (°C)	100–250	120–180	150–500
Pressure (MPa)	0.1–0.4	0.02–0.5	1–5
Residence time (ms)	200–300	250–10,000	15–100
Drying rate (kg/m ² h)	25–120	100–400	500–8,000
Energy usage (kJ/kg H ₂ O)	2,500–3,500	2,200–3,000	550–1,400

Source: From Paulapuro, H., *Developments in Wet Pressing*, PIRA Information Services, Leatherhead, Surrey, U.K., 1993.

35.2.2.4.10 Superheated Steam Drying

Loo and Mujumdar [84] made a transient analysis to evaluate the technical feasibility of using superheated steam as a drying medium replacing hot air. The main attraction of the steam CIT (combined impingement and through-drying) process they propose is the possibility of extremely high thermal efficiencies attainable through reuse of the exhaust steam by reheating, compression, or use in other process-heating applications. Subsequently, Cui and Mujumdar proposed an alternate configuration for steam drying of paper and developed a simple mathematical model to estimate the drying rate and energy consumption [85]. They showed that although the drying rates for tissue products could be increased up to 25 to 30% compared with those of a Yankee dryer, the net heat consumption was extremely low. Their model was verified in a static drying test apparatus operated in the constant-rate period [86].

The effects of steam drying on paper quality need to be examined closely. However, as steam drying of pulp has been used successfully, it is not very likely that it will have adverse effects on sheet properties. Preliminary studies indicate that steam drying may actually enhance the strength properties of paper.

In a recent review of this process, Mujumdar [87] reported the drying rates of 100 to 200 kg/m² h and compared the limitations and advantages of both air and steam drying. He concluded that the process appears to be a viable concept due to its higher energy efficiency, enhancement of certain quality indices for at least some types of pulps, elimination of fire hazard, and reduced space requirements. The technical issues to be resolved are startup and shutdown, condensation of steam on web, air infiltration (sealing at high speed), materials of construction due to corrosion and erosion, steam cleaning, recirculation, compression, heating, and sheet-quality aspects.

35.2.2.4.11 Drying in the Presence of an Electrostatic Field

Recently, Rounsley [88] reported a 5 to 18% increase in the drying rates of paper and coatings in the presence of a nonuniform, static, electric field. This technique has been tested for both felted and unfelted drum drying as well as air impingement and radiant drying. If commercially successful, this concept has the advantage of fast response and may be used for moisture profile control.

In summary, it may be noted that most of the paper drying carried out industrially is accomplished by conventional multicylinder dryers. With the advent of more energy-efficient and economic dryers based on novel concepts, however, it is likely that new drying technology will find industrial acceptance within a decade.

35.3 DRYING OF PULP

If pulp is produced for use in an integrated paper-making machine, there is no need for drying. However, for market pulp, drying up to 10% moisture (90% fiber, 10% water) is necessary.

The pulp web was dried exclusively by contact with steam-heated cans until the mid-1950s. However, air drying (air floater dryers) of pulp is predominant today (e.g., 70% of the U.S. paper industry). Although the application of flash drying in the industry is relatively new, it has found an appreciable market (e.g., 15% of the U.S. paper industry). As the steam-dryer system was introduced to the industry only recently, it is not yet an established drying technique for pulp.

35.3.1 CONVENTIONAL PULP DRYING

In the conventional method, the pulp web is produced on either a fourdrinier wire or a revolving cylinder in a vat in which the level is kept constant by continuous pulp supply. A modern fourdrinier system has a closed head box working at constant level, wet suction boxes to allow the drainage on the wire, and hot water or steam boxes for preheating the web. Pressing is accomplished either with feltless press rolls or doublefelted press rolls. After the press rolls, the dryer section begins. Cylinder dryers or air floater dryers can be used in this section. The described modern fourdrinier system achieves high dryness ahead of the drying section, which affects the economy of the system and allows higher capacities and machine speeds up to 100 t/d and 200 m/min, for example. This unit may have a machine width up to 6.5 m. The revolving cylinder system is less expensive, but

it has lower capacity and presents some problems related to the quality of the sheet [89].

Although the conventional steam-heated cylinder dryers are still predominant in the paper industry, their share in pulp drying has diminished rapidly after the application of air floater dryers. The dryers are 1.2- to 1.5-m diameter case iron cylinders and the steam systems and air systems are similar to the paper dryer system described earlier.

The air floater dryers are also similar to those described in the paper-drying section. Hot air is impinged to the web from blow boxes above and below, and the web floats supported by the airflow. The low sheet tension ensures a greater ability to tolerate disturbances or sheet defects, and the quality of pulp is less affected. Pulp dryers also consist of stacked drying decks, as shown in Figure 35.16. Pulp dryers are larger than paper dryers as pulp is air dried from the press section to the cutter.

35.3.2 FLASH DRYING

Flash drying is a process in which wet pulp is introduced into a stream of hot gases and its moisture is vaporized.

Dewatering of wet pulp is accomplished by various types of presses. After dewatering, moist pulp is introduced into a hot gas stream. The pulp-hot gas mixture passes through a number of flash-drying towers (number depends on the design of the system, but usually two double-towered systems with a cyclone separator between them are used as shown in Figure 35.26) and the dried pulp particles and the moist gases are separated in cyclone separators.

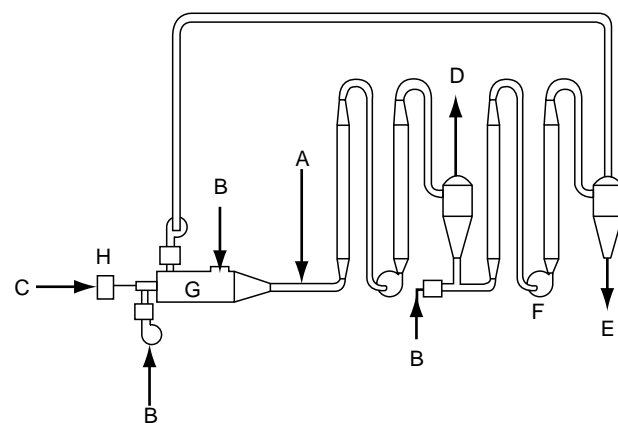


FIGURE 35.26 A flash-dryer unit for pulp (A, moist pulp; B, air; C, oil; D, moist air; E, dried pulp; F, circulation fans; G, air heater; H, steam-heated heat exchangers).

There are also single, rotating horizontal units available for pulp drying.

Flash dryers for pulp are in operation with capacities of up to 500 BDMT/d (bone-dry metric tons). One such plant dries pulp from approximately 60 to 12% water in a two-stage dryer. The lower moisture bound is critical, as overheating below it can cause thermal cross-linking that makes reconstitution of the original fiber difficult. Pulp temperature is maintained below 70°C in both stages, with air inlet temperatures of 400 and 170°C, respectively. Exhaust for the second stage is mixed with inlet air to the first stage. Surface moisture is removed in the first stage (about 30% moisture), and the more delicate second-stage drying is carried out at a lower temperature. Dried pulp is cooled before baling. The first-stage air heater may burn oil or natural gas; the second-stage dryer uses steam-heated air.

Although there are still some questions about the high gas temperatures involved in flash dryers and the absence of wide market acceptance of the products, flash drying is very promising because of its very low operating and capital costs compared with conventional pulp dryers [89].

35.3.3 STEAM DRYING

The steam drying of pulp is a very recent application and not yet widely accepted. Pilot-plant and full-scale tests have shown very attractive results.

The principle of the steam dryer is outlined in the flow diagram shown in Figure 35.27. Wet pulp is fed into the dryer by means of a plug feeder, then disintegrated and fluffed in a steam atmosphere and blown through the drying stages by means of fans. Each stage is a shell and tube heat exchanger, with steam of a higher temperature condensing outside the pulp-transport pipes. The dried pulp and carrier steam are then separated in a cyclone, and the pulp is fed out of the pressurized system by a specially designed discharge screw and blown with air to a cooling cyclone. The generated steam from the pulp moisture is withdrawn to keep the pressure constant and the rest is reused as carrier steam.

The first commercial installation of a steam dryer in Sweden for a chemithermomechanical pulp (CTMP) line with a capacity of 150 t/day showed a 30% reduction in overall drying costs per ton of market pulp compared with an equivalent-size flash dryer [90].

Lower power consumption, very short drying times, easy control, no risk for fire in steam atmosphere, and minor or no effect on pulp quality are among the advantages claimed by the manufacturer.

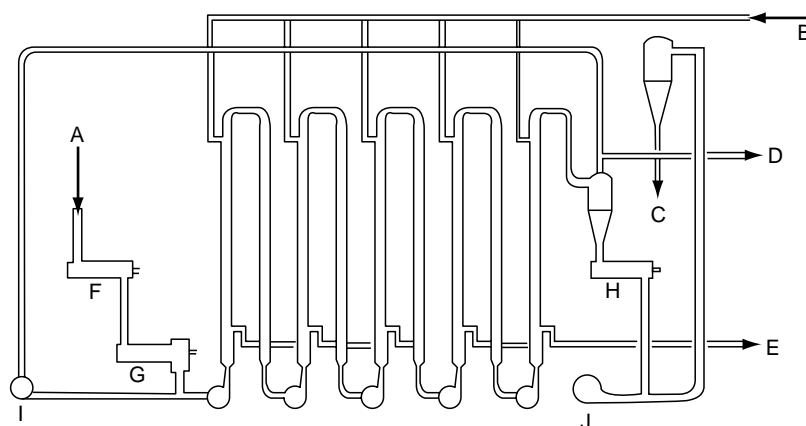


FIGURE 35.27 Flow diagram of a steam dryer for pulp (A, wet pulp; B, heating steam at 6 to 15 bar; C, dried pulp; D, generated steam at 2 to 5 bar; E, condensate; F, plug feeder; G, fluffer; H, discharge screw; I, circulation fans; J, cooling-air fan). (Courtesy of MoDo-Chemetics.)

35.4 CONCLUSION

This chapter summarized the current technology for drying of paper and pulp. Recent developments and trends are also indicated. The reader is referred to the literature cited and other relevant sections or chapters of this handbook for further details and additional information.

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NOMENCLATURE

A	cross-sectional area
B	basis weight of the sheet out of the dryer section as dried (wet basis), g/m^2
c	concentration
C_1, C_2	constants
D	diameter of the dryer cans
D_p	nozzle-plate diameter ratio
D	diffusivity
f	open area
F_{1-2}	overall interchange factor
h	heat-transfer coefficient
H	dimensionless jet-plate spacing
k	thermal conductivity
L	length across which ΔT is measured
M	kilograms of water evaporated per kilogram of paper dried (wet basis)
N	number of dryer cans in contact with the sheet
\overline{Nu}	average Nusselt number

P	pressure
Pr	Prandtl number
q	heat flux
Q_e	condensate flow rate
Q_s	blow-through steam rate
R	gas constant
Re	Reynolds number
S_m	speed of the machine, m/s
Sc	Schmidt number
\overline{Sh}	average Sherwood number
t	time
T	temperature
u	sheet velocity
w	weight of water removed per unit area per unit time, or drying rate
x	coordinate axes
X	sheet thickness
X_p	nozzle-plate width ratio

Greek Symbols

α	thermal diffusivity
β	mass-transfer coefficient
σ	Stefan-Boltzmann constant, $5.67 \times 10^{-8} \text{ W/m}^2 \text{ K}^4$
λ	latent heat of vaporization

Subscripts

a	air
cyl	cylinder
cs	cylinder to sheet
f	felt
o	overall
r	radiant
s	steam
sa	sheet to air
sf	sheet to felt

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36 Drying of Wood: Principles and Practices

Patrick Perré and Roger B. Keey

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36.1 STRUCTURE OF WOOD

36.1.1 FORMATION OF WOOD IN TREES

Wood is produced in the hard stems and branches of trees and shrubs. In these plants, the primary growth is responsible for the stem and branch elongation, whereas the secondary growth, achieved through the cambium activity, is responsible for the thickening of elements (Figure 36.1). Wood has evolved to fulfill the basic needs of these plants during their life: water transport of waterborne nutrients; mechanical strength to support the photoactive canopy of leaves; and resistance to biological attacks. Appearance and properties of this material depend strongly not only on the species but also on the biological diversity and growth conditions of site and climate. Indeed, even for the same species, wood properties depend on the tree and on the position within the tree. One part of this variability is genetically controlled, whereas the other part comes from the varying growth conditions (stand characteristics and silviculture practices). Consequently, wood is a variable material that is extremely difficult to characterize precisely. Hence wood processing, including drying, is very difficult to optimize.

A cross section of a tree (Figure 36.2), from the core to the outer region, shows the following features:

- Pith, a small core of tissue located near the middle of a tree's stem or branches, which originates from the primary growth of the plant
- Woody material, the most important part of mature trees, which is differentiated into sapwood (outer region), where the sap migrates from roots to leaves and heartwood (inner region) that is no longer used for sap transport, which exists only when the stem, at that height, is old enough

- Bark, differentiated into an outer corky dead part (external part of the stem), whose thickness varies greatly with species and age of trees, and an inner thin living part (just near the cambium zone), which carries food from the leaves to the growing elements

36.1.1.1 Knots

As the tree grows in height (primary growth), branching is initiated by lateral bud development. Knots are the bases of branches, which have been covered as the tree grows laterally. After a branch dies, the trunk continues to increase in diameter and surrounds that portion of the branch while the dead branch is still present. This branch has to drop from the tree before clearwood can form. If the knot was alive when the trunk grew around it, the xylem of the trunk and the branch are continuous and the knot fits tightly into the wood. If the branch was dead when the trunk grew around it, no anatomical connection exists between the xylem of the knot and the trunk. The knot is nonadhesive; it may fall out of the wood, leaving a knothole (Figure 36.3).

36.1.1.2 Tissue and Cellular Structure of Wood

Growth in thickness of the bark and wood is caused by cell division in the cambium. New wood cells are formed on the inside of the cambium and new bark cells on the outside. In the cambium region, immature cells differentiate into various kinds of mature xylem (wood) and phloem (inner bark) cells characteristic of the species (Panshin and de Zeeuw, 1980). Then enlargement, elongation, and maturation allow the woody material to be developed (Figure 36.4). Most of the wood cells stay alive for not more than few

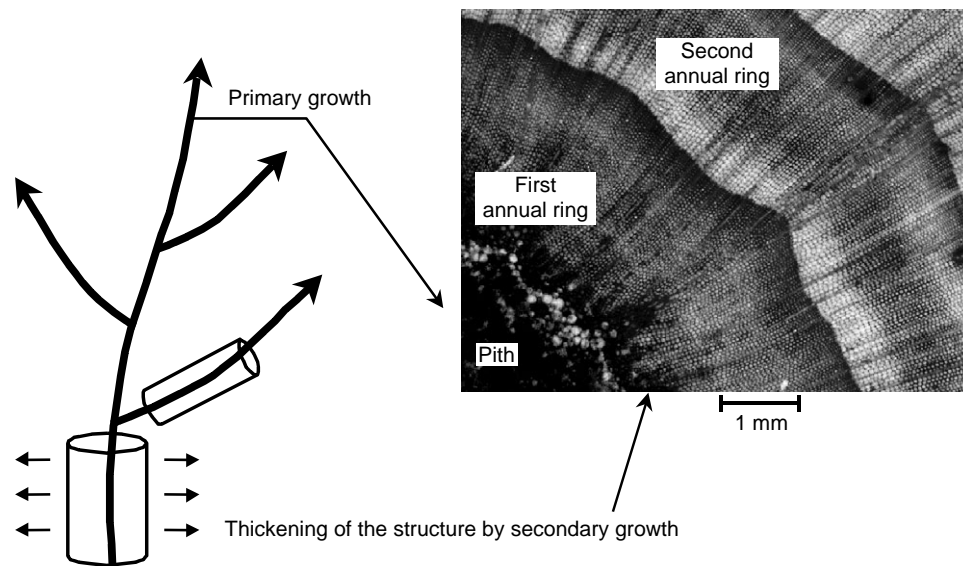


FIGURE 36.1 Formation of wood in trees: the pith originates from the primary growth whereas the wood material is added, along the years, by the secondary growth. (Microphotograph: polished disc of Douglas-fir (*Pseudotsuga menziesii*), LERMAB-ENGREF.)

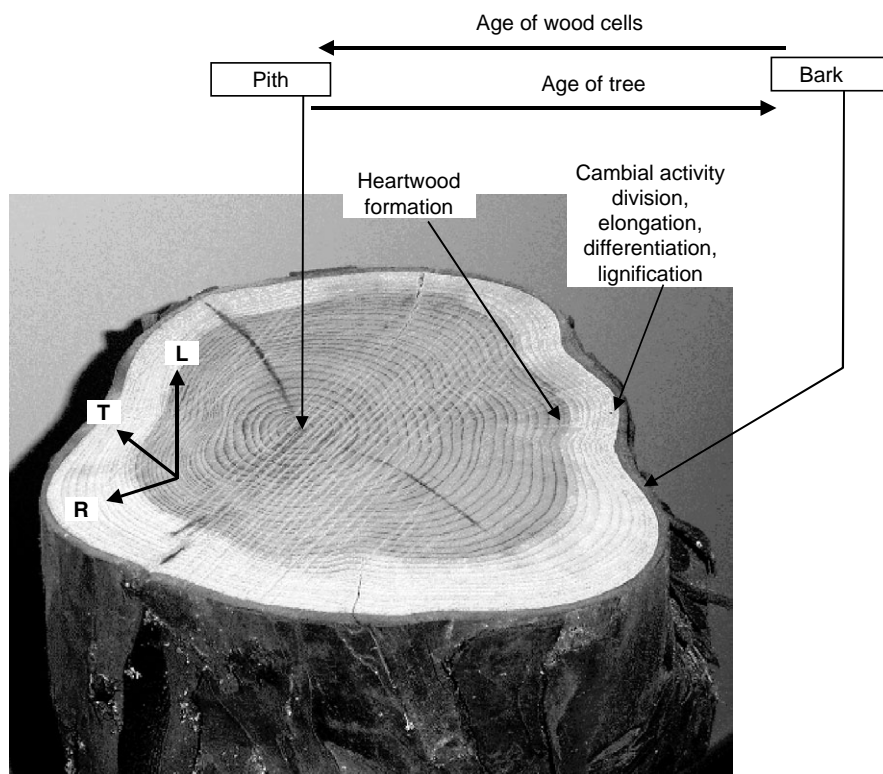


FIGURE 36.2 Cross section of a tree showing the internal structure of the stem. Growth rings can also be observed: light parts are earlywood and dark parts are latewood. Due to this stem geometry, three material directions: longitudinal (L), radial (R), and tangential (T), can be defined at each location. (Photograph: Yew (*Taxus baccata* L.), LERMAB-ENGREF.)

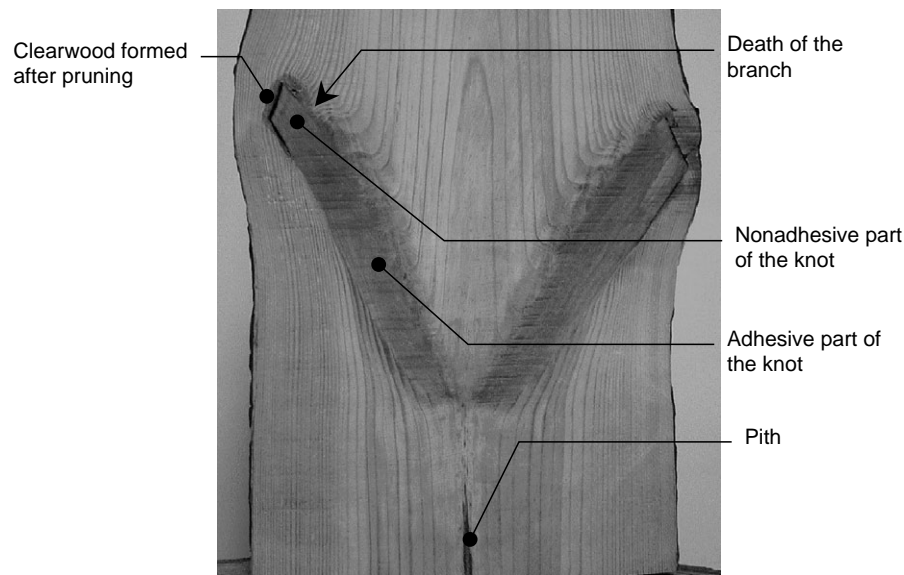


FIGURE 36.3 The existence of knots in wood comes from the interaction between primary growth, responsible for the branch formation, and secondary growth, responsible for thickening of stem and branches. (Photograph: Scots pine (*Pinus sylvestris*), LERMAB–ENGREF.)

weeks; the last development stage, namely lignification, induces the death of these cells. Only parenchyma cells can live for years and they are responsible for the development of heartwood.

In most species in temperate climates, the difference between woods that are formed early in a growing season (earlywood) and that formed later (latewood) is sufficient to produce concentric contours in a cross section (Figure 36.2). These rings are known as growth rings. Each increment in size in the branch or trunk diameter can be observed in these growth rings that remain unchanged once formed. Provided no false rings exist (due to an interruption of the growth in diameter by drought or defoliation by insects), the age at any cross section of the trunk may be determined by counting these rings. Obviously, this simple rule does not apply for tropical species for which growth may be practically continuous throughout the year and no well-defined growth rings are formed, or for which growth rings are the result of the individual rhythm of each tree.

Wood cells are of various sizes and shapes. They are cemented together to form the structural wood material. The majority of wood cells are elongated and pointed at the ends. The types and dimensions of wood cells depend strongly on the species.

In softwoods, woods formed by cone-bearing trees (e.g., fir, pine, and spruce) with naked seeds, the xylem contains mainly tracheids (90%). Tracheids are considerably elongated cells (around 40 μm in diameter and between 2 and 8 mm in length), which ensure both sap flow, by means of numerous bordered pits situated

on the radial cell walls, and mechanical strength. In softwoods, the earlywood is characterized by cells with large radial diameters and thin walls, and hence relatively large cavities. Latewood cells have a much smaller radial diameter and thicker walls, which result in much smaller cavities (Figure 36.4). In addition, some softwoods have resin canals. Parenchyma cells surround these canals and actively secrete resin into the canals, and ultimately into the heartwood.

Hardwood is the common name for the wood of species whose seeds are enclosed in ovaries. These species are more advanced than softwoods in terms of biological evolution; consequently, they produce a more sophisticated anatomical pattern, with cells much more adapted to meet specific requirements in relation to water transport, food storage, and mechanical support:

- Fibers are usually relatively thick-walled, sparsely pitted, and about 1 mm in length. Different fiber cells exist, but the tracheid fibers having bordered pits are generally the most abundant. Although fibers may have a certain role in sap conduction, they basically function as the mechanical support, making the wood usually stronger, denser, and more durable than softwoods.
- Vessels of relatively large diameter are also known as pores. These cells form the main conduits for sap flow. A vessel is built up by several vessel elements, with more or less open end plates, aligned in the longitudinal direction.

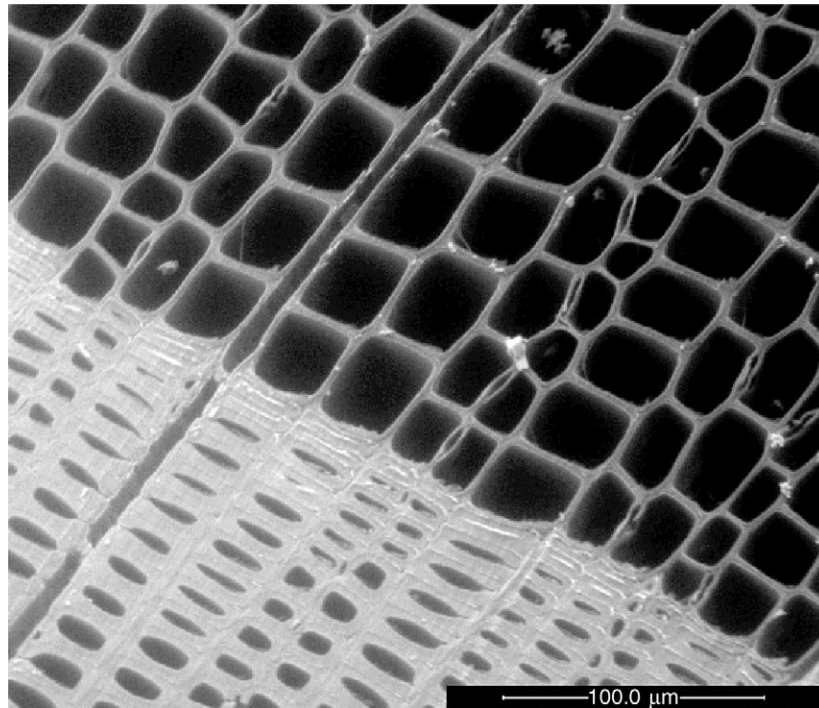


FIGURE 36.4 Wood is formed by cell division in the cambium zone, hence the radial cell lines appear clearly in this figure, in spite of the large variation of radial diameter between earlywood and latewood. Some bordered pits allowing sap flow from one tracheid to the other can also be observed on radial cell walls. (ESEM Photograph: Norway spruce (*Picea abies*), LERMAB-ENGREF.)

- Longitudinal or axial parenchyma cells function mainly in the storage of food.

Several thousands of hardwood species exist, and each one has its own anatomical pattern. The density, for example, ranges from less than 100 kg m^{-3} (i.e., balsa wood) up to more than 1200 kg m^{-3} (i.e., ebony wood). They are usually divided into ring-porous and diffuse-porous types, though all intermediate types can be found:

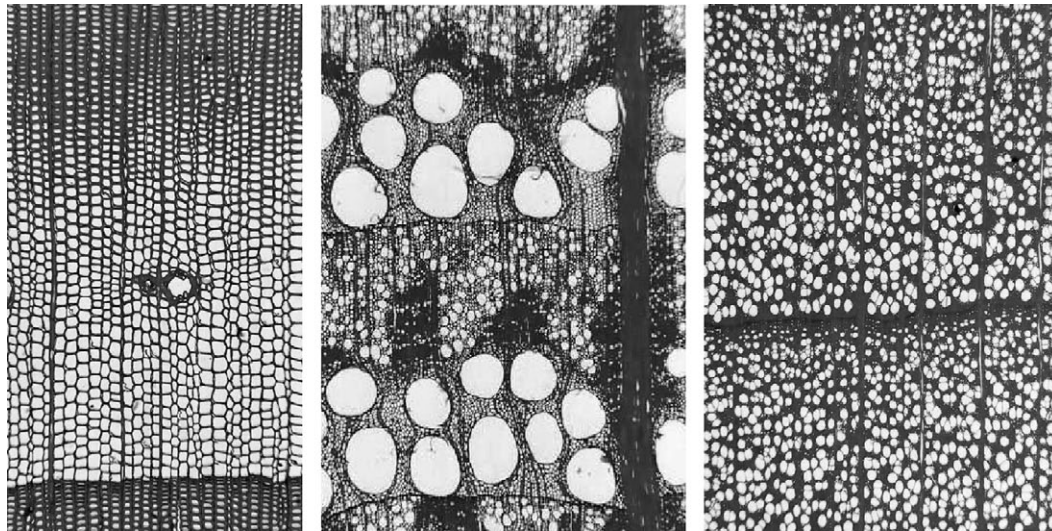
- A ring-porous species produces very large vessels (up to 500 μm in diameter) in earlywood. Simple perforation plates allow the vessel cells to communicate easily.
- Diffuse-porous species have smaller vessels (50 to 100 μm in diameter) of almost uniform size and distribution throughout the growth ring. In diffuse-porous species, vessel cells are usually connected by scalariform (“ladderlike”) perforation plates.

Figure 36.5 depicts the typical anatomical patterns encountered in temperate species.

Both hardwoods and softwoods have cells (usually grouped into structures or tissues) that are oriented horizontally in the radial direction and which are called rays. The rays, composed of parenchyma with lignified cell walls or sclerenchyma, connect various layers from pith to bark for storage and transfer of food. In softwoods, rays are one-cell thick. In hardwoods, they vary in size from one-cell wide and a few cells high to more than 15-cell wide and several centimeters high. Rays represent planes of weakness along which drying checks develop easily.

36.1.1.3 Heartwood Formation

Just few weeks after the formation of xylem, it contains mostly dead cells, but continues to play a role of utmost importance for the plant: the transport of sap. Hence the vascular system so produced is called sapwood. This active zone may vary in thickness and number of growth rings, commonly up to about 15 years, which represents several centimeters in radial thickness. As a rule, the more vigorously growing trees have more extensive sapwood. In sapwood, parenchyma cells stay alive and function primarily in the storage of food.



(a) Norway spruce (*Picea abies*) (b) Pedunculate oak (*Quercus rubra*) (c) European beech (*Fagus sylvatica*)

FIGURE 36.5 Typical anatomical patterns encountered in temperate species: (a) softwood, (b) ring-porous species, and (c) diffuse-porous species. The height of these images represents about 2 mm. (Microphotographs: J.C. Mosnier, LERMAB–ENGREF.)

After some years, an intense biological activity of these parenchyma cells gives rise to heartwood formation. Metabolites are deposited in the heartwood and the tree uses the heartwood as a place to store waste products that are collectively known as extractives. These include resins, gums, oils, and tannins that stop up the vessels (or the tracheids in softwood) and clog the wood. In heartwood, all the cells are dead and inactive; they do not function in either water conduction or food storage. The heartwood is often darker, slightly denser, and more durable (resistance to fungi or insect attack) than the sapwood and plays an important role in supporting the tree. However, numerous species do not have dark heartwood (e.g., spruce, fir, and beech) and no correlation exists between heartwood color and durability.

In some species, such as certain oaks, the vessels become plugged with the development of cellular membranes known as tyloses, which enter the vessels from adjacent parenchyma cells (Figure 36.6).

36.1.2 CHEMICAL COMPOSITION

Wood is a typical organic material made up of three main elements: carbon, oxygen, and hydrogen. Because very small variations between different wood species are observed, the numbers depicted in Table 36.1 are broadly general but can be higher for some tropical species. Nitrogen as well as some additional inorganic elements (sodium, potassium, calcium, magnesium, and silicon) are also essential compounds, which are mostly involved in the metabolism of living cells during wood

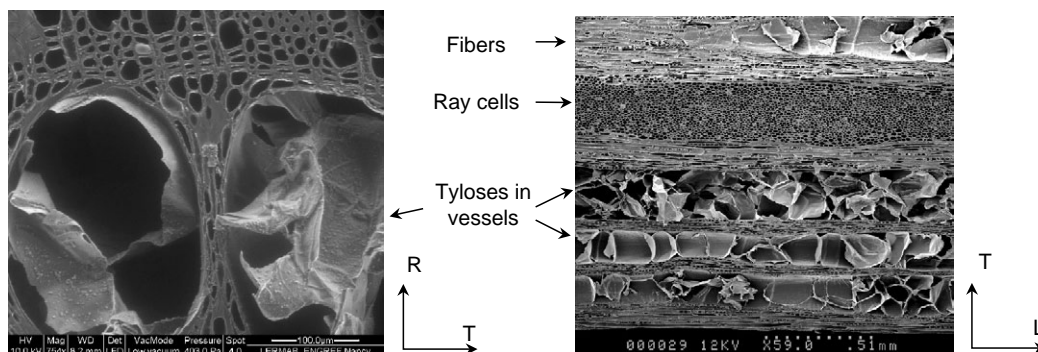


FIGURE 36.6 Example of tyloses development in the heartwood of Pedunculate oak (*Quercus rubra* L.). (ESEM photographs: Patrick Perré and Riad Bakour, LERMAB–ENGREF.)

TABLE 36.1
Elementary Composition of Wood

Element	Content (%)
Carbon	49–50
Hydrogen	6
Oxygen	43–44
Nitrogen	<1
Inorganic elements (Na, K, Ca, Mg, Si)	≤1*

Source: Bosshard, H.H., *Holzkunde, Band 2, Zur Biologie, Physik und Chemie des Holzes (Wood Science: Biology, Physics and Chemistry of Wood, Vol. 2)*, Birkhäuser Verlag, Basel, 1984.

*Exceptions exist in tropical species.

formation but are also important for the strength of the cell wall.

The cell wall has a complex structure, composed of a network of cellulose microfibrils interconnected by hemicellulose chains and incrustated by pectins and proteins (Mark, 1967). Cellulose is a linear polymer with an average degree of polymerization of 10,000. These polymer chains are organized in elementary fibrils that are partly crystalline to the extent of 50 to 70%. Elementary fibrils are bundled with hemicelluloses to form microfibrils. The latter are very strong in their longitudinal direction and their crystalline part is very difficult to penetrate with water. The hemicelluloses are branched, low-molecular-weight polymers composed of several different kinds of pentose- and hexose-sugar monomers. Cells that are dead at maturity (fibers, tracheids, etc.) may impregnate their walls with lignin, which makes these walls impervious to water. Lignin is often regarded as the cementing agent that binds individual cells together and is insoluble in most solvents. Softwood lignin is thought to incorporate about 500 phenylpropane units (Keey et al., 2000), but it is a variable and complex three-dimensional polymer.

Several layers have to be distinguished in the wall thickness, not only by the macromolecular composition but also by the microfibril orientation, i.e., the angle between the fibril orientation and the longitudinal axis of the cell. Between two adjacent cells lies a highly lignified region called middle lamella. Both middle lamella and adjoining primary walls are sometimes referred to as compound middle lamella. The secondary cell wall is laid down after cell extension is complete. In direction to the cell lumen, the secondary cell wall is subdivided into three zones: S₁, S₂, and S₃. The lignin content significantly decreases from about 90% in the middle lamella down to about 20% of the S₂ layer. The varying fibril orientation in the

particular layers (50° to 70° in S₁, 10° to 30° in S₂, and 60° to 90° in S₃) causes a mechanical locking effect, leading to a very high stiffness of the overall cell. Due to its thickness and low value of microfibril angle, the S₂ layer is responsible for the high strength and low shrinkage of wood in the longitudinal direction. The overall distribution of wood polymers does not vary greatly with common species (see Table 36.2), but a comprehensive study (Fengel and Wegener, 1984) reveals much wider variations.

Both organic and inorganic extraneous materials are found in wood. The organic component takes the form of extractives, which contribute to such wood properties as color, odor, taste, decay resistance, density, hygroscopicity, and flammability. Extractives include tannins and other polyphenolics, coloring matter, essential oils, fats, resins, waxes, gum starch, and simple metabolic intermediates. Extractives may constitute roughly 5 to 30% of the wood substance. The inorganic component of extraneous materials generally constitutes 0.2 to 1.0% of the wood substance. Calcium, potassium, and magnesium are the more abundant elemental constituents.

36.1.3 REACTION WOOD AND JUVENILE WOOD

As plants cannot move, they have developed several features to adapt to their environment such as the effects of wind and site slope. Reaction wood is part of this capacity for adaptation; it allows the plants to maintain or modify the orientation of different axes in space. In softwood trees, the reaction wood produces growth stresses smaller than in normal wood; the tensile level is smaller and can even become negative in this compression wood. In hardwood trees, reaction wood forms wood with higher growth-stress level than in normal wood and is known as tension wood. For example, branches are horizontal elements that must work against gravity. In the case of hardwoods, tension wood is present in the upper part of the cross sections of branches whereas, in the case of

TABLE 36.2
Typical Wood Composition

Compound	Content (%)	
	Softwoods	Hardwoods
Cellulose	42 ± 2	45 ± 2
Hemicelluloses	27 ± 2	30 ± 5
Lignin	28 ± 3	20 ± 4

Source: Walker, J.C.F., *Primary Wood Processing Principles and Practice*, Chapman & Hall, London, 1993.

softwoods, compression wood is found in the lower part. However, the compression wood can also be found in some primitive groups of hardwoods (Carlquist, 2001).

36.1.3.1 Compression Wood

Compared with normal wood, this tissue is characterized by shorter tracheids, higher lignin and hemicellulose content, and lower cellulose content. This reaction wood is easily identified on smooth surfaces, in particular in a transverse view. When compression wood is formed, the growth rings appear darker, reddish brown, and often wider than on the opposite side. Therefore, when compression wood develops in the same side for several years, the cross section of the stem tends to be oval with an eccentric pith in the core; this is typical of branches or stem of bent trees.

At the anatomical level the cells observed in cross section are more rounded than rectangular, showing large intercellular spaces (Figure 36.7). The cell wall consists only of ML, P, S₁, and S₂ layers. Once dry, the cell wall shows deep, helically arranged checks from the lumen. The latter is rather thick and its microfibril angle is much larger than in normal wood (about 45° from the axial direction). In consequence, the density is higher, the longitudinal shrinkage is increased to some percentage (compared with 0.1 to 0.2% for normal wood), and, in spite of the higher density, the longitudinal mechanical properties are less than in normal wood. Due to the large microfibril angle in the S₂ layer, the effect is opposite in the transverse plane: lower shrinkage and higher stiffness.

36.1.3.2 Tension Wood

Tension wood is characterized by increased cellulose content and increased density. Sawn surfaces appear

woolly and rough. Strength properties are reduced. Longitudinal shrinkage can be more than 1%. No significant coloration marks out tension-wood zones. In addition, the regulation of hardwood seems to be subtler than in softwoods, sometimes leading to very localized presence of reaction wood (thin layers in the radial direction and variable angular position from one year to the other). Therefore, the macroscopic features are not very reliable.

At the microscopic level, tension wood is much easier to identify when it is fully developed. Fiber cell walls are much thicker than normal, enclosing very small lumens. Secondary walls are loosely attached to the primary wall and thus are responsible for some of the differing mechanical properties. The thick secondary wall of tension-wood fibers is significantly lesser lignified; it consists of almost pure cellulose. Due to this consistency, this layer is termed gelatinous or G layer (Figure 36.8). The microfibrils are almost parallel to the grain. The reason for tension-wood shrinkage is not well understood. The loose contact between the G layer and the remaining cell wall, which does not restrain the outer cell region (i.e., primary layer) and from contraction during drying, is one possible explanation. However, a recent work tends to prove that the G layer itself has a high longitudinal shrinkage, which is not consistent with the small microfibril angle (Clair, 2001). It may be noted that intermediate indications of tension wood exist without the G layer.

36.1.3.3 Juvenile Wood

All trees during their growth produce juvenile wood, i.e., the inner core of xylem surrounding the pith. The time during which juvenile wood is formed varies among individuals, with species, and with environmental conditions. It is often recognized that juvenile

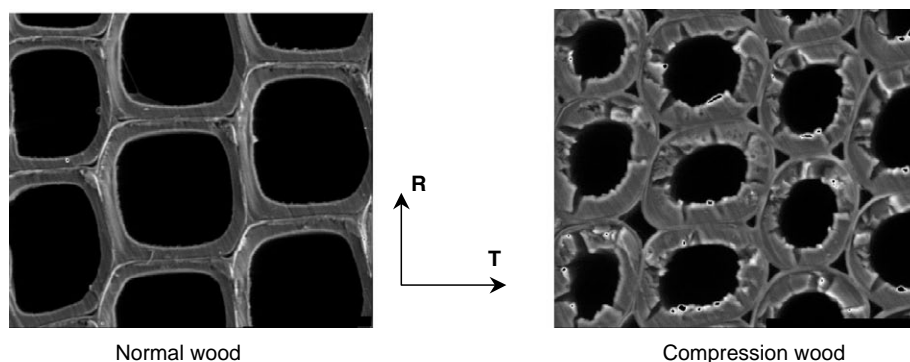


FIGURE 36.7 Compression wood compared with normal wood. (ESEM photographs: White fir (*Abies alba*), LERMAB-ENGREF.)

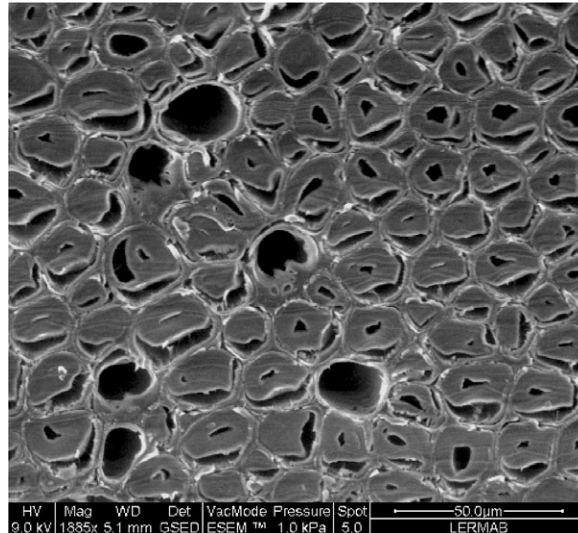


FIGURE 36.8 Tension wood in Pedunculate oak (*Quercus rubra*); note the existence of the G layer. (ESEM photograph: LERMAB–ENGREF.)

wood formation lasts as long as living branches exist at the corresponding height of the tree. The transition from juvenile to mature wood takes place gradually. No clear demarcation exists between juvenile and adult wood.

In juvenile wood the cells are smaller than those of the mature xylem. Particular differences exist in the length of the cells as well as in the structure of the layered cell wall. The microfibril angle in the S_2 layer is greater than in cells of the mature tissue. As for compression wood, this causes a higher value of longitudinal shrinkage and a reduced tensile strength. In addition, the spiral grain (angle between the stem axis and the fiber orientation) is often large in the juvenile wood. Together with the high longitudinal shrinkage, this explains why the warping of timber containing juvenile wood may be dramatic after drying. Juvenile wood is a major problem in processing wood from plantations of fast-growing species (eucalyptus, radiata pine, etc.) that produce logs made up mostly of juvenile wood.

36.1.4 IMPLICATIONS FOR THE DRYING PROCESS

To induce the ascent of sap in tree, the menisci present in the leaf stomata pull up water (Zimmerman 1983). Because most trees are more than 10 m high, one can deduce that the absolute liquid pressure in the sap column is negative. No gaseous phase can exist in such conditions. The vascular system developed in trees has many other implications for the drying process:

- Because the system is designed for longitudinal sap flow from the roots to the canopy, the wood material is strongly anisotropic.
- Because of negative pressure, the vascular system must be able to support a gas invasion due to injury or cavitation. This is the role of bordered pits or vessel-to-vessel pits. These anatomical features may dramatically inhibit the fluid migration in the wood.
- In heartwood, due to metabolite deposition, aspiration or closure of bordered pits, or tylose development, the permeability is often reduced by one or several orders of magnitude.
- The wood is fully saturated in the sapwood part of logs (an air-free sap column is required to obtain negative pressures), whereas the heartwood zone is generally only partly saturated.

Table 36.3 indicates some orders of magnitude generally observed for the moisture content of greenwood. Indeed, because the sapwood part is fully saturated, the maximum moisture content in this zone can be calculated by assuming that the entire pore volume is filled with water:

$$X = \frac{\phi \rho_\ell}{(1 - \phi)\rho_s} \text{ with } \phi = 1 - \frac{\rho_0}{\rho_s} \quad (36.1)$$

where ϕ is the porosity, ρ_0 is the basic density (oven-dry mass/green volume), ρ_s is the density of the cell

TABLE 36.3
Typical Values of Moisture Content Found in Greenwood

Wood Type	Moisture Content Dry Basis (%)	
	Sapwood	Heartwood
Softwoods	150–200	40–80
Hardwoods	80–120	60–100

wall substance ($\rho_s \cong 1530 \text{ kg m}^{-3}$), and ρ_ℓ is the sap density ($\rho_\ell = 1000 \text{ kg m}^{-3}$).

To highlight that the development of wood in trees leads to a very anisotropic material, Table 36.4 indicates some order of magnitude for dimensionless anisotropy ratios found in wood for the most important properties involved in drying. The ease of fluid migration in wood (i.e., the permeability) is by far the property that presents the highest anisotropy ratio. The reduction in wood permeability from sapwood to heartwood affects particularly the longitudinal direction in hardwoods (especially for ring-porous species developing tyloses) and all directions in softwoods.

36.2 BOARD SCALE

36.2.1 WATER IN WOOD: SORBED AND CAPILLARY WATER, AND SHRINKAGE

36.2.1.1 Moisture Content of Wood

The moisture content of wood (dry basis) is defined by the following mass ratio:

$$X = \frac{\text{mass of water}}{\text{oven-drymass}} \quad (36.2)$$

TABLE 36.4
Order of Magnitude of the Dimensionless Anisotropy Ratios Encountered in Wood Relative to Tangential Value

Property	Direction		
	T	R	L
Stiffness	1	2	20
Shrinkage	1	0.5	≈ 0
Thermal conductivity	1	1.5	2
Mass diffusivity	1	1–2	20
Permeability	1	1–10	100–10 ⁵

Moisture in wood is distinguished between the liquid sap present unattached in the cell lumens (free water) and the water molecules held in the cell walls (bound water), with an activity less than 1.

36.2.1.2 Free Water

The sap flows in wood through the vascular system produced by the cambium. This liquid, present in the cell cavities, is usually referred to as “free water” just because its properties are very close to those of liquid water: density, viscosity, saturated vapor pressure, etc.

However, one has to keep in mind that this water is tied to the solid matrix through capillary forces. Due to the surface energy of the interface between liquid and gas (the surface tension), together with the contact angle between this interface and the woody matrix, a pressure difference exists between liquid and gaseous phases. This pressure difference, which obeys Laplace’s law, increases with decreasing pore diameter. Because liquid is the wetting phase in the case of water and wood, the liquid pressure is less than the gaseous one.

In addition, due to the curvature of the interface, a deviation of the saturated vapor pressure exists. This deviation can be calculated from the definition and properties of the Gibbs free energy and leads to Kelvin equation (Dullien, 1992):

$$\ln \varphi = \ln \left(\frac{P_v}{P_{vs}} \right) = -\sigma \left(\frac{1}{r_1} + \frac{1}{r_2} \right) \frac{M_v}{\rho_\ell RT} \quad (36.3)$$

In Equation 36.3, P_{vs} is the saturated vapor pressure, P_v is the equilibrium vapor pressure at the curved interface, σ is the surface tension, r_1 and r_2 are the two principal radii of the surface, and M_v is the molar mass of water. The quantity φ is known as the relative humidity. However, a very small pore radius is required for the deviation to become significant (Table 36.5).

TABLE 36.5
Pressure Difference ΔP and Relative Humidity φ at the Surface for Different Radii Values (Values Calculated at 20°C, for a Perfectly Wetting Liquid and Cylindrical Tubes)

Radius of the Capillary	ΔP (Equivalent Water Column)	$\varphi = P_v/P_{vs}$
1 mm	0.146 kPa (14.9 mm)	0.999999
100 μm	1.46 kPa (0.149 m)	0.999989
10 μm	14.6 kPa (1.49 m)	0.99989
1 μm	0.146 MPa (14.9 m)	0.9989
0.1 μm	1.46 MPa (149 m)	0.989
0.01 μm	14.6 MPa (1.49 km)	0.898

36.2.1.3 Bound Water

Bound moisture is associated with the hygroscopic nature of the woody components. There are some uncertainties about the limits of hygroscopic behavior, particularly with woods of high extractives content; but it is useful to define a maximum sorptive moisture content, called the fiber saturation point (FSP). If the capillary condensation effects in pores greater than $0.1\ \mu\text{m}$ in equivalent cylindrical diameter are ignored, FSP of the wood may be defined as the equilibrium moisture content (EMC) in an environment of 99% relative humidity. This yields a value of 30 to 32% for most commercial species (Keey et al., 2000) at room temperature. FSP falls with increasing temperature. For a softwood such as Sitka spruce (*Picea sitchensis*), FSP falls from about 31% at 25°C to 23% at 100°C (Stamm, 1964).

As the relative proportions of the woody components vary only within narrow ranges for common commercial species, the EMCs at a given relative humidity and temperature are closely similar for these woods. However, at high relative humidities deviations from mean values can appear. Shubin's data (1990) show, for instance, that at 95% relative humidity the EMC at 42.4°C ranges from 22% for a pine to 33% for an oak. Hoadley (1980) notes that in species with a high extractives content, such as redwood (*Sequoia sempervirens*) and mahogany (*Swietenia mahoganii*), the fibers remain saturated at 22 to 24% moisture content, whereas birch (*Betula* spp.) may have a moisture content up to 35% at fiber saturation.

Because of sorption hysteresis, the EMC at a given relative humidity is higher in drying (desorption) from greenwood than in wetting (adsorption) from perfectly dried wood. The ratio of values normally ranges between 0.75 and 0.88 (Schniewind, 1989). However, in previously dried wood subject to environmental swings in relative humidity, the isotherm for adsorption and desorption becomes more similar. Thus, for practical purposes, the hygroscopicity of wood is sufficiently similar for most commercial species for generalized sorption data to be useful. The curves depicted in Figure 36.9 represent an average between desorption and adsorption. They were plotted using the mathematical correlation used in the numerical code "Transpore." The parameters of this expression have been fitted from various data available in the literature: those published by Rasmussen in 1961 (Siau, 1984) and those of Loughborough on Sitka spruce published by Hawley in 1931 and arranged by Keylwerth in 1949 (Kollmann and Côté, 1968; Joly and More-Chevalier, 1980).

36.2.1.4 Differential Heat of Sorption

Sorbed water in the cell wall has a lower enthalpy than liquid water. However, contrary to other forms of water, such as solid, the enthalpy of bound water increases with increasing moisture content up to FSP. Above this value, the enthalpy of water in wood is essentially the same as that of liquid water.

The value of the differential heat of sorption can be calculated from the sorption isotherms using the Clausius–Clapeyron equation (Skaar, 1988):

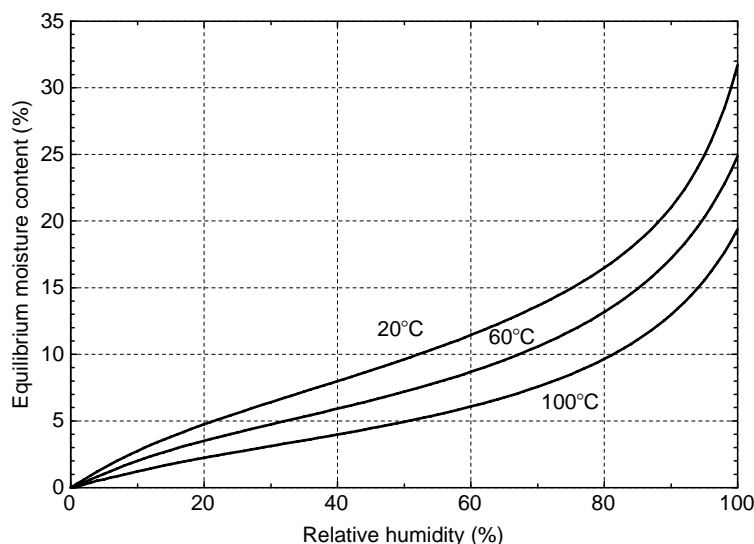


FIGURE 36.9 Sorption isotherms calculated by a mathematical expression fitted from published data.

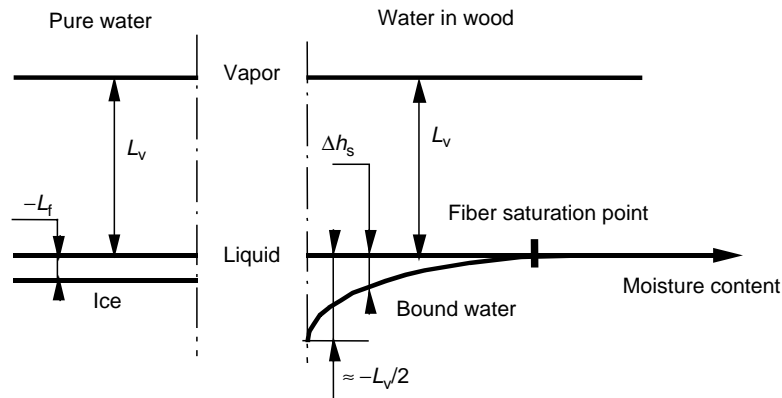


FIGURE 36.10 Differential heat of sorption versus the moisture content. (Adapted from Skaar, C., *Wood–Water Relations*, Springer, Berlin, 1988.)

$$\Delta h_s = -\frac{R}{M_v} \frac{d\left(\ln \frac{P_v}{P_{vs}}\right)}{d\left(\frac{1}{T}\right)} \quad (36.4)$$

This heat of sorption is slightly higher than 1000 kJ kg^{-1} for oven-dry wood and decreases rapidly with moisture content to reach zero, with a horizontal slope at the FSP (Figure 36.10).

36.2.1.5 Shrinkage

In trees, the cell walls of wood are in a fully swollen condition. However, when sorbed water is removed from the cell wall, new hydrogen bonds form between the hydroxyl groups of the molecular chains and reduce the distance between the chains. This variation affects the dimension in a direction normal to the microfibril direction. This fact explains why the longitudinal shrinkage is usually negligible. In the transverse directions, the dimensional variations plotted against bound water content are very close to a straight line (Figure 36.11). Shrinkage can then be defined by the total shrinkage (variation of dimension between the green state and the oven-dry condition) or in terms of a shrinkage coefficient (variation of dimension divided by the variation in moisture content). For most species, the tangential shrinkage is about twice the radial shrinkage. Several features can explain this observation: the cell arrangement of tissues, the difference between earlywood and latewood, and the presence of ray cells. The extrapolation of length values against moisture content to the original unshrunk length yields the so-called shrinkage intersection point (SIP) (Figure 36.11). Normally, this moisture content is a little higher than the FSP, but often the two moisture content values are assumed to be same.

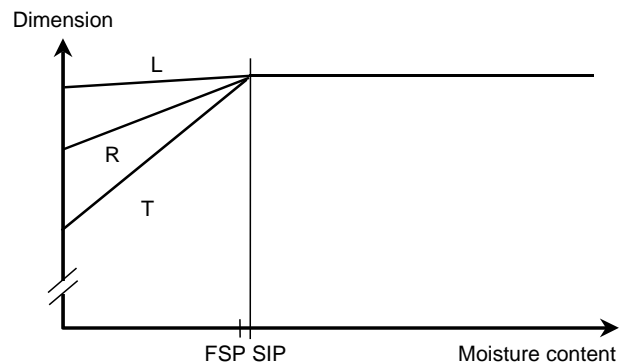


FIGURE 36.11 Dimensional variations of a typical wood sample vs. moisture content.

The shrinkage values tend to increase with basic density, but values vary strongly between and within species. Table 36.6 indicates the order of magnitude values encountered in some species.

In the case of compression wood, the microfibril angle in the S_2 layer is large, which results in a lower transverse shrinkage and a significant longitudinal shrinkage. In spite of the very low value of the microfibril angle in the G layer, tension wood of hardwood has also an abnormally high longitudinal shrinkage.

As logs are sawn in green state, shrinkage occurs after sawing. The size of each section is reduced by shrinkage and also the shape of the section changes caused by the tangential or radial anisotropy in shrinkage (Figure 36.12):

- Only the quartersawn sections without the pith keep their rectangular shape
- Flatsawn sections cup into a trough-like shape

TABLE 36.6

Order of Magnitude of Shrinkage Values in the Transverse Plane for Some Species. FSP is Assumed to be Equal to 30% for All Species

Species	Radial Shrinkage		Tangential Shrinkage	
	Total (%)	Coefficient (%/%)	Total (%)	Coefficient (%/%)
Spruce, pine	5.0	0.17	9.0	0.30
Oak	6.0	0.20	11.0	0.37
Beech	6.5	0.22	12.0	0.40
Balsa	3.0	0.10	6.0	0.20
Teak	3.0	0.10	5.0	0.17
Okoumé	4.0	0.13	6.5	0.22
Azobé	8.0	0.27	11.0	0.37

Source: Kollmann, F.P. and Côté, W.A., *Principles of Wood Science and Technology, Solid Wood*, Vol. 1, Springer, Berlin, 1968; Aléon, D., Chanrion, P., Négrié, G., and Perré, P., *FormaXylos 4—Le séchage (Training in Wood Science: Drying, Vol. 4)*, CD-Rom français/English, CTBA, Paris, 2003.

- c. Quartersawn sections, which include the pith, present a nonuniform thickness once dried; near the pith, the thickness shrinks along the radial direction whereas the tangential shrinkage is involved elsewhere
- d. A squared section cut along the grain direction becomes rectangular after drying
- e. A diamond shape is obtained if the grain direction is along the diagonal

36.2.2 HEAT AND MASS TRANSFER IN WOOD

36.2.2.1 Fluid Migration in Wood: Single-Phase Flow

The understanding of the fluid migration in wood is of utmost importance to understand the drying process. The differences in drying behavior between species and within the log primarily come from the permeability value and from the initial moisture saturation. When only one fluid phase is present, the classical Darcy's law applies:

$$\mathbf{v} = -\frac{K}{\mu} \nabla(P) \quad (36.5)$$

where \mathbf{v} is the apparent velocity of the fluid through the specimen (m s^{-1}), K is the permeability (m^2), and μ is the dynamic viscosity of the fluid (Pa s^{-1}).

Fluid migration in wood uses the vascular system developed by trees for their physiological requirements. For this reason, wood has several specific features concerning permeability among which the most important are the anisotropy ratios. Wood has dramatic anisotropy ratios: the longitudinal permeability can be 1000 times greater than the transverse permeability for softwoods and more than a million-fold for hardwoods (Banks, 1968).

Table 36.7 summarizes some values of directional permeability available in the literature for different species. Depending on the experimental apparatus and the protocol used by the authors, some data are missing in the papers. For example, it is not always

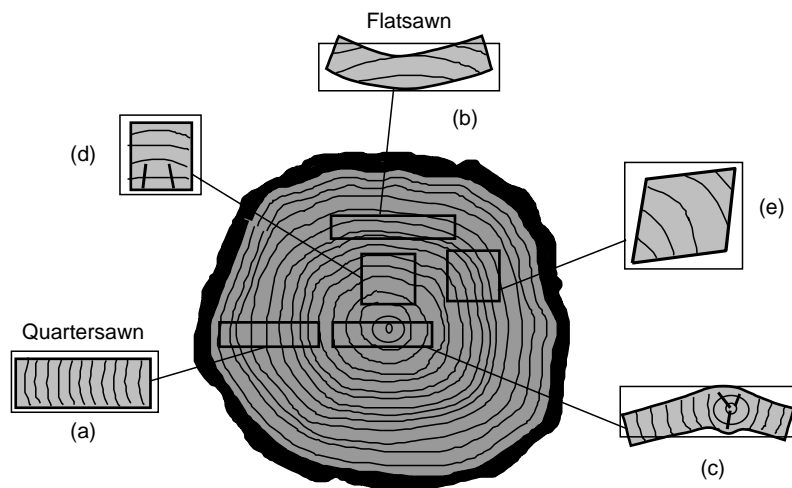


FIGURE 36.12 Section deformations depending on the sawing pattern. The shape after drying results from the anisotropy ratio between radial and tangential shrinkage. These deformations exist even when the equilibrium is achieved and with a uniform moisture content throughout the section. (Adapted from Aléon, D., Chanrion, P., Négrié, G., and Perré, P., *FormaXylos 4—Le séchage (Training in Wood Science: Drying, Vol. 4)*, CD-Rom français/English, CTBA, Paris, 2003.)

TABLE 36.7
Order of Magnitude of Permeability for Some Species along the Three Material Directions of Wood

References	Species	Notes	Permeability (m ²)			Anisotropy Ratio		
			L × 10 ¹²	R × 10 ¹⁶	T × 10 ¹⁶	K _L /K _R	K _L /K _T	K _R /K _T
Choong and Kimbler, 1971	<i>Populus</i> sp.	S. l. a	0.59	—	0.44	—	13,000	—
	<i>Populus</i> sp.	H. l. a	0.61	0.15	0.18	40,000	33,000	0.835
	<i>Alnus rubra</i>	S. l. a	1.9	0.15	0.12	125,000	166,000	1.327
	<i>Liquidambar</i> sp.	S. l. a	2.7	0.19	0.69	145,000	39,000	0.273
	<i>Liriodendron</i> sp.	H. l. a	5.4	—	0.80	—	67,000	—
	<i>Sequoia</i> sp.	S. l. a	4.9	11.2	19.4	4,000	2,000	0.580
	<i>Sequoia</i> sp.	H. l. a	0.25	0.112	0.88	23,000	3,000	0.127
	<i>Pseudotsuga</i> sp.	H. l. a	0.026	0.000	0.000	—	—	—
	<i>Pseudotsuga</i> sp.	S. l. a	0.049	—	0.37	—	1,300	—
	<i>Pinus</i> sp.	H. l. a	0.0026	—	0.17	—	153	—
Choong. et al., 1974	<i>Acer rubrum</i>	S. g. o	10.3	—	—	8,300	11,400	1.4
	<i>Acer rubrum</i>	H. g. o	7.4	—	—	—	—	—
	<i>Liriodendron</i> sp.	S. g. o	28.9	—	—	1,450	1,600	1.1
	<i>Liriodendron</i> sp.	H. g. o	1.87	—	—	—	—	—
	<i>Liquidambar</i> sp.	S. g. o	13.9	—	—	1,300	2,750	2.1
	<i>Liquidambar</i> sp.	H. g. o	15.3	—	—	—	—	—
	<i>Quercus rubra</i>	S g	62.0	—	—	13,000	18,000	1.4
	<i>Quercus rubra</i>	H g	56.0	—	—	—	—	—
	<i>Quercus falcata</i>	S g	69.0	—	—	110,000	143,000	1.3
	<i>Quercus falcata</i>	H g	13.0	—	—	—	—	—
Chen et al., 1998	<i>Liriodendron</i> sp.	S l	26.0	4.5	—	57,000	—	—
	<i>Liriodendron</i> sp.	H l	0.1	—	—	—	—	—
	<i>Juglans nigra</i>	S l	27.0	0.08	—	3 × 10 ⁶	—	—
	<i>Juglans nigra</i>	H l	0.0052	—	—	—	—	—
	<i>Quercus rubra</i>	S l	61.0	0.68	—	900,000	—	—
	<i>Quercus rubra</i>	H l	45.0	—	—	—	—	—
Perré, 1992 and 2002	<i>Picea</i> sp.	S. g. a	0.2	—	—	700	—	—
	<i>Fagus sylvatica</i>	S g. a	3.8	—	—	3,000	65,000	21.0
	<i>Fagus sylvatica</i>	H g. a	1.4	—	—	3,000	—	—
	<i>Populus</i> sp.	H g. s	0.03	—	—	10,000	—	—

L, longitudinal; R, radial; T, tangential; S, sapwood; H, heartwood; l, permeability to liquid; g, permeability to gas; a, air-dried sample; o, oven-dried sample.

easy to calculate the permeability ratios from permeability value, or vice versa. Choong et al. (1974), for example, have reported the permeability values for sapwood and heartwood for the longitudinal direction, but not for the transverse directions. Only the mean anisotropy ratio is available in this paper. Perré (1992) and Perré et al. (2002) have used an experimental procedure to determine the longitudinal permeability and the anisotropy ratio on the same sample. In these instances, they just obtained the ratios and decided not to calculate the transverse permeability accordingly.

The variability is impressive for both the permeability values and the anisotropy ratios. In spite of the scatter in the data, general trends are exhibited for the longitudinal permeability:

- The most permeable species in the longitudinal direction are among the ring-porous species (*Quercus* spp.) that have very large vessels (up to 500 µm in diameter).
- The diffuse-porous hardwoods are fairly permeable too; probably, the large number of vessels can offset their smaller diameter (around 50 µm).
- Softwood species are generally less permeable; these trees have no specific elements for sap flow, so the fluid has to pass through the small openings, the bordered pits, at the end of each tracheid along the path, which is 1 to 2 mm each. Certain softwood species, such as *Pinus radiata*, however, are very permeable.

The transverse permeability and the anisotropy ratios are very variable. The heartwood part of logs is usually much less permeable than the sapwood part (Comstock, 1967). This is due to tyloses development and extractives deposition (tannins, gums, etc.) in hardwoods and due to the aspiration of bordered pits in softwoods.

36.2.2.1.1 Pit Aspiration

In softwoods, the tracheids have considerable pitting in their radial walls. These bordered pits are specialized valves to seal and isolate tracheids if they become damaged or embolized, but remain open for sap flow. At the pit location, the double cell wall takes the shape of the external part of a torus. The external diameter of this torus is in the range 10 to 20 μm , depending on the position in the annual growth ring (the diameter is smaller in latewood). The torus is suspended by the margo, a net of radially oriented microfibrils, with openings up to some micrometers wide. In normal operation, the sap flows from one tracheid to the other simply by using these small openings (Figure 36.13a).

When gas invades one tracheid, the gas-liquid interface is blocked in the margo due to capillary forces. These forces press the torus against the opposite border of the pit (Figure 36.13b). Then, hydrogen bonds keep the torus in this position; the pit is now impermeable (Figure 36.13c). Such sealing of a pit is known as aspiration.

This subtle mechanism is vital for trees, but causes some difficulties in wood drying. Indeed, pit aspiration occurs as soon as water is removed from the wood, sometimes even during the heartwood formation. In particular, it is impossible to avoid pit aspiration when drying softwoods under normal conditions.

Only freeze-drying, or changing the liquid phase for a solvent with low surface tension before drying, allows air-dried samples without pit aspiration to be obtained (Comstock and Côté, 1968; Meyer, 1971; Bolton and Petty, 1978; Fumoto et al., 1984). The permeability values depend on the species and on the author, but all these data show that air-dried samples, with aspirated pits, have permeability values considerably smaller than unaspirated samples (typically ranging from 1 to 10%). Due to thicker cell walls, smaller pit radii, and more rigid structures, the percentage of aspirated pits is much less in the latewood part of samples (Siau, 1984).

36.2.2.2 Generalized Darcy's Law: Multiphase Flow

When two phases coexist, the generalized Darcy's law must be used. The volumetric flow rate of each phase is considered to be proportional to the pressure gradient of the corresponding phase. The phenomenological coefficient is the product of the permeability K by a function of saturation called "relative permeability" to the considered phase.

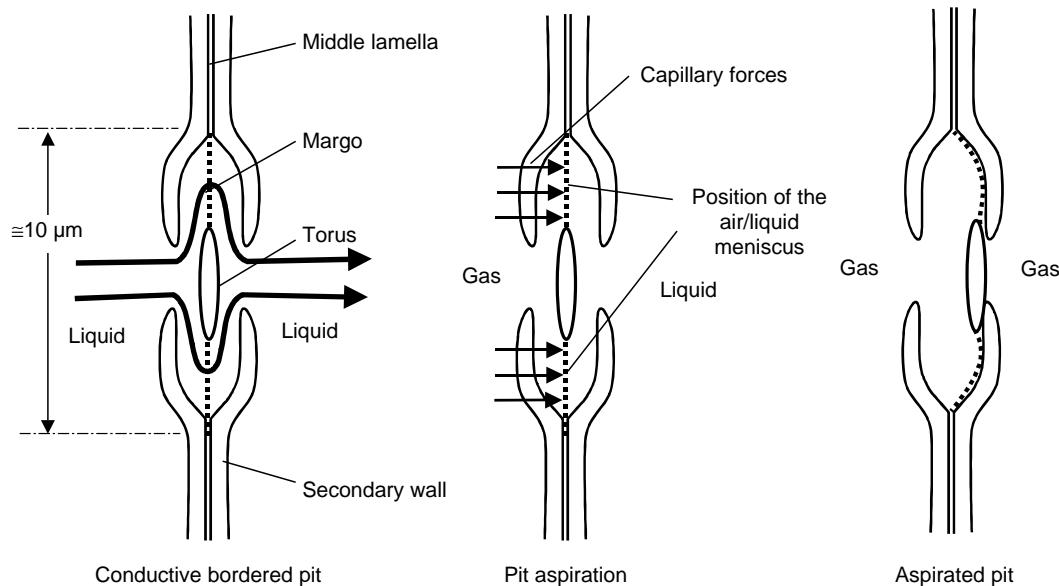


FIGURE 36.13 The mechanism of pit aspiration: a clever strategy to limit the damage caused by any gas invasion due to injury or cavitation of the sap column. (Adapted from Siau, J.F., *Transport Processes in Wood*, Springer, Berlin, 1984.)

For the gaseous phase

$$\mathbf{v}_g = - \frac{K k_{rg}(S)}{\mu_g} \nabla P_g \quad (36.6)$$

For the liquid phase

$$\mathbf{v}_\ell = - \frac{K k_{r\ell}(S)}{\mu_\ell} \nabla P_\ell \quad (36.7)$$

The liquid pressure is related to the gaseous pressure through the capillary pressure function:

$$P_\ell = P_g - P_c(S) \quad (36.8)$$

Equation 36.6 and Equation 36.7 must be consistent with Darcy's law (Equation 36.5) when one single-fluid phase occupies the porous medium. Consequently, the relative permeability functions fulfill the following conditions:

Gas only Liquid only

$$k_{r\ell}(0) = 0, \quad k_{r\ell}(1) = 1 \quad (36.9)$$

$$k_{rg}(0) = 1, \quad k_{rg}(1) = 0$$

Although depending on both initial and boundary conditions, the relative permeability values are usually supposed to be the function of saturation only. Even with this simple assumption, their experimental determination remains very challenging.

Very few results are available in the literature for wood. Some data have been published by Tesoro et al. (1972, 1974). A recent paper by Tremblay et al. (2000) describes an indirect method to measure this property.

Based on the geometrical model of tracheids proposed by Comstock (1970), Spolek and Plumb (1980) derived an expression for the relative permeability in softwoods. This model assumes that all tracheids are exactly similar and that the wetting-phase distribution is ideal. Their final expression involves an irreducible saturation, below which no liquid flux is possible.

This concept of irreducible saturation has been widely discussed, especially in the context of petroleum production (Dullien, 1992). Depending on the boundary conditions, one part of the fluid phase that occupied the medium at the beginning of the experiment remains in the medium, even though its flow rate has vanished. This part seems to be trapped in the medium in what is called a "pendular" state. The amount of the residual part increases with the imbibition or drainage velocity. In addition, the amount of trapped phase after the experiment reduces with time; surface spreading in the solid phase and the

edge-capillary pressure are the two mechanisms that can explain this observation (Dullien, 1992). Due to the microporosity that exists in the cell walls, such mechanisms are likely to exist in wood also. However, the concept of irreducible saturation has led to unrealistic computed moisture content profiles during drying (Perré, 1987), though later modeling of softwood drying incorporates a similar concept successfully as a mean to account for pit aspiration (Nijdam et al., 2000). In this case, "irreducible saturation" corresponds to the condition when pit aspiration is sufficiently advanced such that the remaining free moisture is immobilized in isolated pockets in the wood.

Based on the measurements in the longitudinal direction reported by Tesoro et al. (1972), on the curve computed from the tracheid model (Spolek and Plumb, 1980), and on those considerations regarding the concept of irreducible saturation, the following functions have been proposed for softwoods by Perré et al. (1993) (Figure 36.14):

In the transverse direction (radial or tangential)

$$k_{rg}^T = 1 + (2S - 3)S^2 \quad \text{and} \quad k_{r\ell}^T = S^3 \quad (36.10)$$

In the longitudinal direction

$$k_{rg}^L = 1 + (4S - 5)S^4 \quad \text{and} \quad k_{r\ell}^L = S^8 \quad (36.11)$$

36.2.2.3 Capillary Pressure

Some works can be found on the determination of capillary pressure functions in wood. Mercury porosimetry exhibits a dramatic effect of the sample thickness in the longitudinal direction (Trénard, 1980). For short samples, the cell lumens are directly accessible to the mercury; whereas for longer samples, the liquid has to pass through the small openings, the pits that exist between cells, clearly illustrating their bottlenecking effect. The centrifuge method has been used successfully by Spolek and Plumb (1981) on softwoods and by Choong and Tesoro (1989) on various species. To determine the moisture content–water potential relationship of wood, Cloutier and Fortin (1991) used a tension plate and a pressure plate. This is a drainage method and their results could easily be converted into a classic capillary pressure function.

Because certain anatomical features govern the morphology of wood pores, particular methods can be applied to wood:

- To compute a capillary pressure curve, Spolek and Plumb (1981) have developed the geometrical

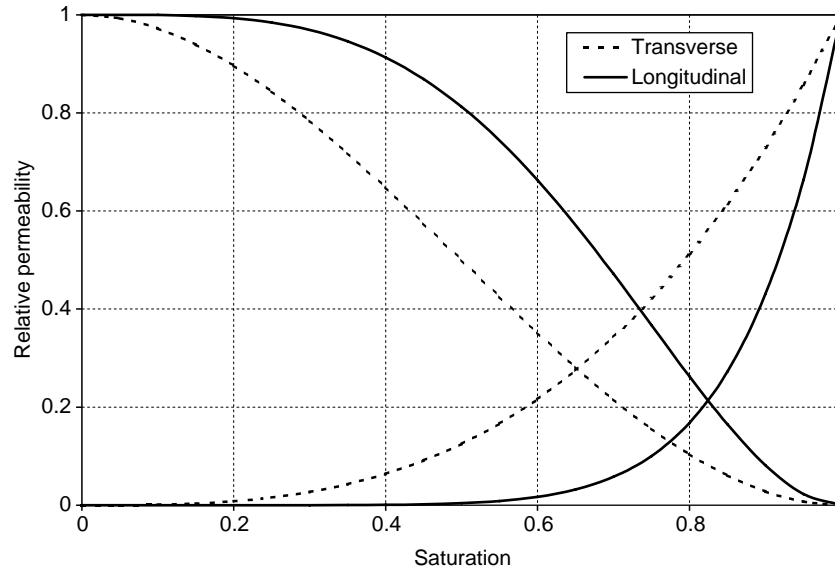


FIGURE 36.14 Relative permeability curves calculated using equations (36.10) and (36.11).

model of the tracheid shape proposed by Comstock (1970). Although it may be simplistic to assume that all tracheids have exactly the same shape, they obtained a good trend for the capillary pressure function by this means.

- Because the longitudinal direction of wood is very marked, it is quite simple to obtain the three-dimensional structure of the material from a cross section. Figure 36.15 depicts the examples of capillary pressure curves calculated from microscopic images of cross sections of wood. In this case, the pore-size distribution has been calculated using image processing (Perré, 1997; Perré and Turner, 2002).

diffusivities, respectively having units $\text{m}^2 \text{s}^{-1}$ and ω_v is the mass fraction of vapor in the gaseous phase.

By using the bound-liquid diffusivity data of Stamm (1963), it is possible to obtain the following least-squares, best-fit correlation for D_b :

$$D_b = \exp\left(-12.82 + 10.90 X_b - \frac{4300}{T}\right) \quad (36.14)$$

where T is the temperature in Kelvin.

On assuming isothermal conditions and constant total pressure, the microscopic vapor flux can be expressed with the gradient of the bound-water content as the driving force by Equation 36.13.

$$\mathbf{f}_v = -\left(\frac{M_v}{RT} D_v \frac{\partial P_v}{\partial X_b}\right) \nabla X_b \quad (36.15)$$

36.2.2.4 Bound-Water Diffusion

Macroscopic bound-water diffusion results from transport mechanisms that take place at the microscopic scale, i.e., diffusion of the bound water through the cell walls and vapor diffusion due to Fick's law. At the microscopic scale, these two fluxes can be expressed as follows:

$$\mathbf{f}_b = -\rho_s D_b \nabla X_b \quad (\text{Bound-water flux}) \quad (36.12)$$

$$\mathbf{f}_v = -\rho_g D_v \nabla \omega_v \quad (\text{Vapor flux}) \quad (36.13)$$

In Equation 36.12 and Equation 36.13, D_b and D_v represent the microscopic bound-liquid and vapor

Within the anatomical structure of wood, any combination in series or parallel of vapor diffusion (in lumen and pits) and bound-water diffusion (in the cell walls) is a possible pathway to drive water from high to low moisture content regions (Figure 36.16). Because Equation 36.12 and Equation 36.15 use the same driving force, the expressions for the macroscopic bound-water diffusivity in the radial and tangential directions can be calculated even using homogenization techniques according to these microscopic properties, together with the pore morphology (Perré and Turner, 2002). Equation 36.14, derived from specific experimental measurements, exhibits a dramatic increase of

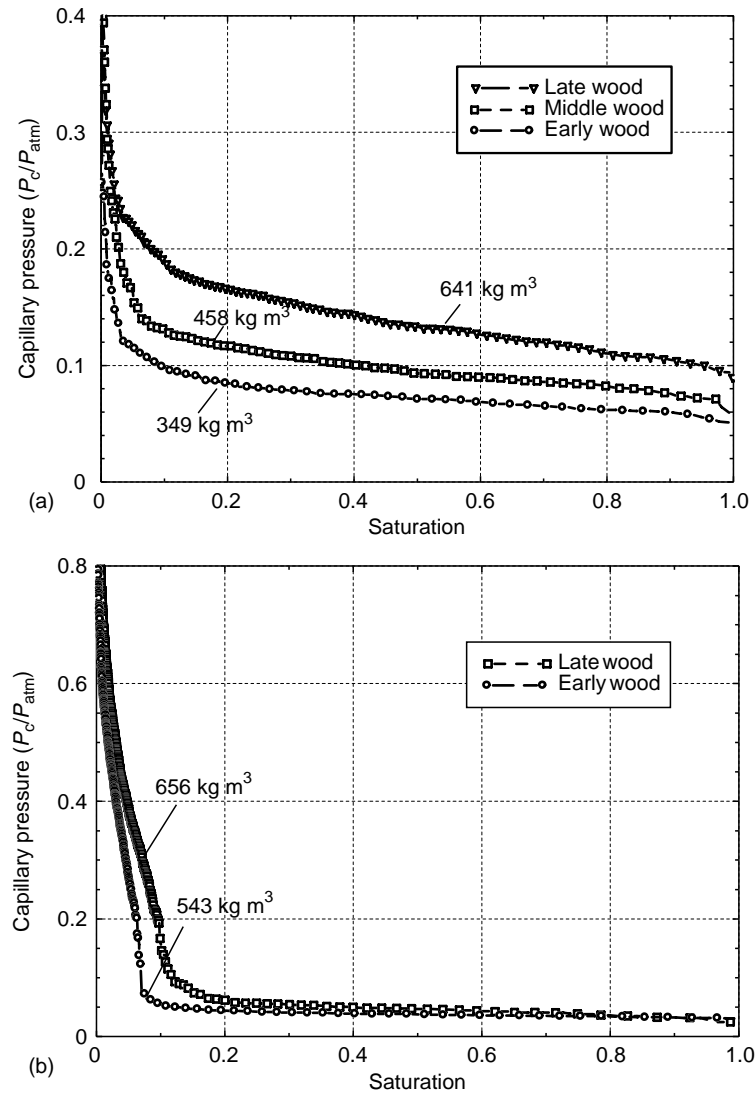


FIGURE 36.15 Capillary-function curves determined using image analysis. (a) spruce (*Picea abies*): the cells have thicker walls and smaller radial extension in latewood part, hence the highest value of the capillary-pressure curve. One also has to be aware that full saturation is obtained with a lower amount of water in latewood (the porosity of this part is very small); (b) beech (*Fagus sylvatica*): because beech is a pore diffuse-porous hardwood species, no significant difference is observed between these parts. The low capillary pressure obtained for saturation values above 0.2 corresponds to the meniscus radii located in the vessel elements. The dramatic increase for low saturation values is due to the small lumen diameters of the parenchyma and fiber cells.

the bound-water diffusivity as the bound moisture content increases. For this reason, the same trend is predicted from the calculations or measurements at the macroscopic scale; bound-water diffusion is always easier for higher bound-water contents.

When using the gradient of bound water as a driving force, the macroscopic flux reads

$$\mathbf{f}_b = -\rho_0 D_b \nabla X_b \quad (36.16)$$

This expression is consistent with the derivation of the second Fick's law by using the mass balance equation:

$$\frac{\partial X_b}{\partial t} = \nabla(D_b \nabla X_b) \quad (36.17)$$

For the sake of simplicity, the foregoing expressions have always assumed isothermal conditions. However, this assumption fails for certain drying processes

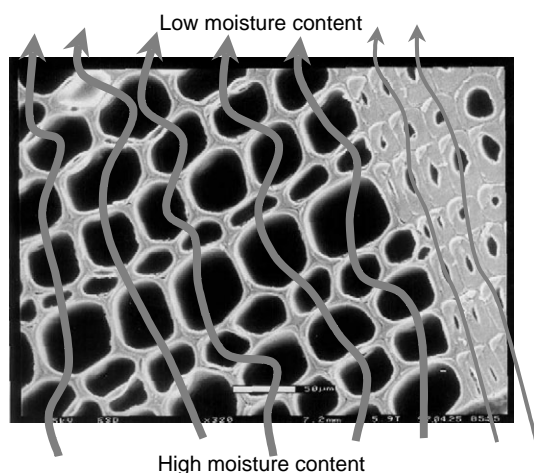


FIGURE 36.16 Moisture diffusion in the hygroscopic range: any combination in series or parallel of vapor diffusion (in lumen and pits) and bound-water diffusion (in the cell walls) is able to drive water from high to low moisture content regions.

(e.g., impingement drying, radio-frequency, or microwave heating). The problem of mass migration due to a thermal gradient is a matter of scientific debate. Siau (1984, 1995) gives a good review of possible formulations that can be used in nonisothermal conditions.

36.2.2.5 Physical Formulation

Several sets of macroscopic equations are proposed in the literature for the simulation of the drying process. The first fundamental difference between them lies in the number of state variables used to describe the process:

- One: moisture content (or an equivalent variable: saturation, water potential, etc.)
- Two: moisture content (or equivalent) and temperature T (or an equivalent variable: enthalpy etc.)
- Three: moisture content (or equivalent), T (or equivalent), and gaseous pressure P_g (or an equivalent variable: air density, intrinsic air density, etc.)

Perré (1999) gives a critical review of the possibilities and limitations given by these different sets of equations. The use of moisture content alone is the basis of many correlations of lumber-drying rates (see Keey et al., 2000). Doe et al. (1996a) have found that, at relatively low temperatures ($<40^{\circ}\text{C}$) used in drying eucalyptus hardwoods, both moisture content and temperature are needed in their evaluation of moisture movement. The importance of gaseous

pressure appears in the simulation of high-temperature seasoning, vacuum drying, or microwave drying (Perré and Degiovanni, 1990; Perré and Turner, 1996, 1999a).

As a general rule, a one-variable model (moisture content only) should be avoided (except, possibly, when considering kiln behavior), especially because it is not able to account for the very important coupling that exists during drying between heat and mass transfer. The two-variable model is appropriate for most of the drying conditions encountered in industry. Finally, the three-variable model, quite complex, should be reserved for processes during which the internal pressure has a significant impact on internal moisture transport (processes with internal vaporization: vacuum drying, high-temperature drying, radio-frequency drying, etc.).

The second fundamental difference between the drying models lies in the number of space variables used to describe the process. One-dimensional models (thickness) are fast running and allow the most important phenomena to be caught (transfers and drying stress), as the transfers occur mainly over the long faces of lumber boards in a kiln. For a more refined approach, a second direction can be useful:

- The width is needed for a better evaluation of drying rate and stress development, in particular to account for the grain direction and the presence of both sapwood and heartwood.
- The length is absolutely required in the case of processes with internal vaporization due to the huge anisotropy ratios of wood.

Finally, a three-dimensional model allows a comprehensive geometrical modeling (Perré and Turner, 1999c).

The most comprehensive macroscopic formulation is presented in this chapter; it accounts for all transfer mechanisms presented in this section, including the effect of internal gaseous pressure [Equation T1 through Equation T5]. The set of equations, as proposed below, derives for the most part from Whitaker's works (Whitaker, 1977, 1998) with extensions required to account for wood properties and drying with internal overpressure (Perré and Degiovanni, 1990). In the equations set out below, all variables are averaged over a REV (representative elementary volume), large enough for average local quantities to be defined but small enough to avoid variations due to macroscopic gradients. Although very powerful for numerous different configurations, these equations assume equilibrium conditions at the microscopic level and have some limitations (Perré, 1998).

Water Conservation

$$\frac{\partial}{\partial t} (\varepsilon_w \rho_w + \varepsilon_g \rho_v + \bar{\rho}_b) + \nabla (\rho_w \bar{v}_w + \rho_v \bar{v}_g + \overline{\rho_b v_b}) = \nabla (\rho_g \bar{D}_{\text{eff}} \nabla \omega_v) \quad (\text{T1})$$

Energy Conservation

$$\begin{aligned} \frac{\partial}{\partial t} (\varepsilon_w \rho_w h_w + \varepsilon_g (\rho_v h_v + \rho_a h_a) + \bar{\rho}_b \bar{h}_b + \rho_o h_s - \varepsilon_g P_g) + \nabla (\rho_w h_w \bar{v}_w + (\rho_v h_v + \rho_a h_a) \bar{v}_g + h_b \overline{\rho_b v_b}) \\ = \nabla (\rho_g \bar{D}_{\text{eff}} (h_v \nabla \omega_v + h_a \nabla \omega_a) + \bar{\lambda}_{\text{eff}} \nabla T) + \Phi \end{aligned} \quad (\text{T2})$$

Air Conservation

$$\frac{\partial}{\partial t} (\varepsilon_g \rho_a) + \nabla (\rho_a \bar{v}_g) = \nabla (\rho_g \bar{D}_{\text{eff}} \nabla \omega_a) \quad (\text{T3})$$

where the gas- and liquid-phase velocities are given by the generalized Darcy's law:

$$\bar{\mathbf{v}}_\ell = -\frac{\bar{K}_\ell \bar{k}_{r\ell}}{\mu_\ell} \nabla \varphi_\ell, \quad \nabla \varphi_\ell = \nabla P_\ell - \rho_\ell g \nabla \chi \quad (\text{T4})$$

where ℓ is w, g, the quantity φ is known as the phase potential, and χ is the depth scalar. All other symbols have their usual meaning.

Boundary Conditions

For the external drying surfaces of the sample, the boundary conditions are assumed to be of the following form:

$$\begin{aligned} \mathbf{J}_w|_{x=0^+} \cdot \hat{\mathbf{n}} &= h_m c M_v \ln \left(\frac{1 - x_\infty}{1 - x_v|_{x=0}} \right) \\ \mathbf{J}_e|_{x=0^+} \cdot \hat{\mathbf{n}} &= h(T|_{x=0} - T_\infty) \\ P_g|_{x=0^+} &= P_{\text{atm}} \end{aligned} \quad (\text{T5})$$

where \mathbf{J}_w and \mathbf{J}_e represent the fluxes of total moisture and total enthalpy at the boundary, respectively, and x denotes the normal position from the boundary in the external medium.

36.2.3 PROCESS OF DRYING

36.2.3.1 Low-Temperature Convective Drying

Low-temperature convective drying is the most widespread industrial process for seasoning wood in kilns. In this case, the role of internal gaseous pressure is almost negligible and transfer occurs mainly in the direction of the board thickness. Two periods of drying may be distinguished: (1) a constant drying-rate period and (2) a decreasing drying-rate period.

36.2.3.1.1 The Constant Drying-Rate Period

This stage is very common for certain porous media, but is rarely seen with wood. However, it exists almost always for fresh boards consisting of sapwood that are dried under moderate conditions (Perré et al.,

1993; Perré and Martin, 1994). During this period, the exposed surface of the board is still above the FSP. As a result, the vapor pressure at the surface is equal to the saturated vapor pressure, and is a function of the surface temperature only.

Coupled heat and vapor transfer occur across in the boundary layer (Figure 36.17). The heat flux supplied by the airflow is used solely for transforming the liquid water into vapor. During this stage, the drying rate is constant and depends only on the external conditions (temperature, relative humidity, velocity, and flow configuration). The temperature at the surface is equal to the wet-bulb temperature. Moreover, because no energy transfer occurs within the medium during this period, the whole temperature of the board remains at the wet-bulb temperature.

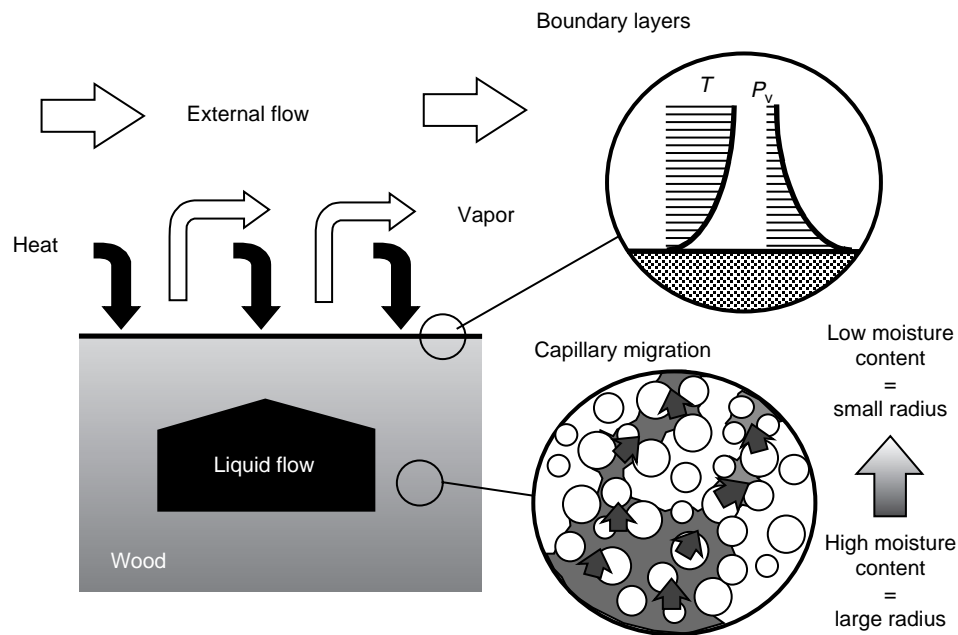


FIGURE 36.17 Constant drying-rate period: the moisture migrates inside the medium mostly by capillary forces; evaporation occurs at the exchange surface with a dynamical equilibrium within the boundary layers between the heat and the vapor flows. (Adapted from Perré, P., The numerical modeling of physical and mechanical phenomena involved in wood drying: an excellent tool for assisting with the study of new processes, Tutorial, *Proceedings of the Fifth International IUFRO Wood Drying Conference*, Québec, Canada, 1996, 9–38.)

The exposed surface is supplied with liquid water coming from the inside of the board by capillary action; the liquid migrates from regions with high moisture content (liquid–gas interfaces within large pores) toward regions with low moisture content (liquid–gas interfaces within small pores).

The constant drying-rate period lasts as long as the surface is supplied with liquid. Its duration depends strongly on the drying conditions (magnitude of the external flux) and on the medium properties. The liquid flow inside the medium is expressed by Darcy's law (permeability \times gradient of liquid pressure).

36.2.3.1.2 The Decreasing Drying-Rate Period

Once the surface attained the hygroscopic range, the vapor pressure becomes smaller than the saturated vapor pressure (Figure 36.18). Consequently, the external vapor flux is reduced and the heat flux supplied to the medium is temporarily greater than what is necessary for liquid evaporation. The energy in excess is used to heat the board, the surface at first and then the inner part by conduction. A new, more subtle, dynamic equilibrium takes place. The surface vapor pressure, hence the external vapor flow, depends on both temperature and moisture content. To maintain the energy balance, the surface temperature increases as the surface moisture content decreases. This

leads to a decreasing drying rate (the heat supplied by the airflow becomes smaller and smaller).

A two-zone process develops inside the wood: (1) an inner zone, where liquid migration prevails, and (2) a surface zone, where both bound-water and water-vapor diffusion take place. During this period, a conductive heat flux must exist inside the board to increase the temperature and to evaporate the liquid driven by gaseous diffusion. The region of liquid migration naturally reduces as the drying progresses and finally disappears. The process is finished when the temperature and the moisture content attain the outside air temperature and the EMC, respectively.

36.2.3.2 Drying at High Temperature: The Effect of Internal Pressure on Mass Transfer

To reduce the drying time without decreasing the quality of the dried product, the drying conditions must be such that the temperature of the product is above the boiling point of water. Such conditions ensure that an overpressure exists within the material, which implies that a pressure gradient drives the moisture (liquid or vapor) toward the exchange surfaces (Lowery, 1979; Kamke and Casey, 1988).

At normal atmospheric pressure, the boiling point of water equals 100°C. Consequently, in order to

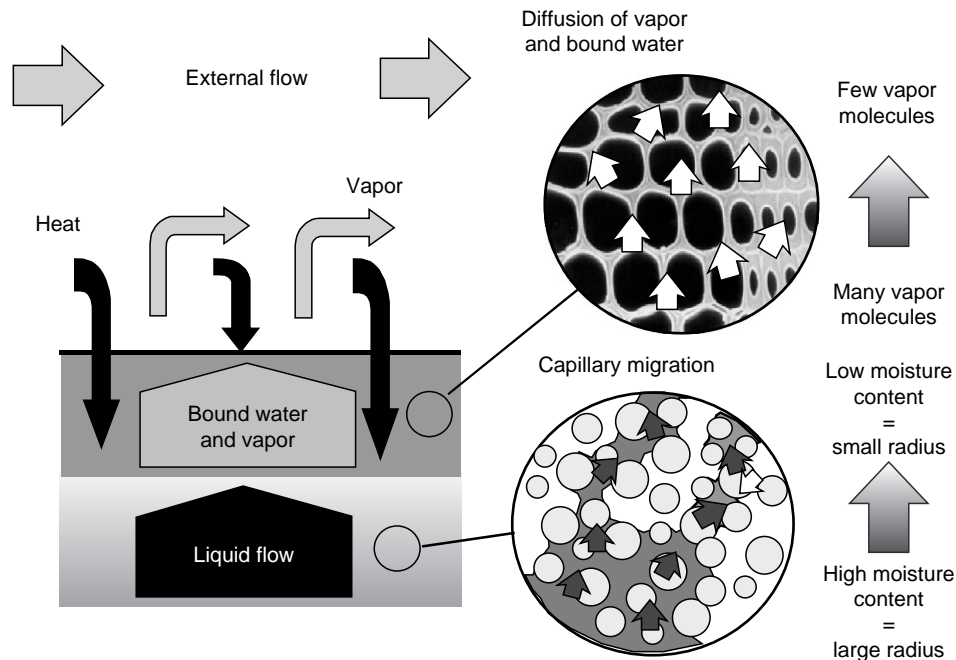


FIGURE 36.18 Second drying period: a region in the hygroscopic range develops from the exposed surface. In that region, both vapor diffusion and bound-water diffusion act. Evaporation takes place partly inside the medium. Consequently, a heat flux has to be driven toward the inner part of the board by conduction. (Adapted from Perré, P., The numerical modeling of physical and mechanical phenomena involved in wood drying: an excellent tool for assisting with the study of new processes, Tutorial, *Proceedings of the Fifth International IUFRO Wood Drying Conference*, Québec, Canada, 1996, 9–38.)

obtain an internal overpressure, the temperature of the porous medium must be above that level during at least one part of the process. This is the aim of convective drying at high temperature (moist air or superheated steam) and a possible aim of contact drying or drying with an electromagnetic field (microwave or radio frequency).

However, as shown in Figure 36.19, it is possible to reduce the boiling point of water by decreasing the external pressure and, consequently, to obtain a high-temperature effect with relatively moderate drying conditions. This is the principle of vacuum drying, particularly useful for lumber that would be damaged by high temperature levels.

Whenever an overpressure exists inside a board, the large anisotropy ratios imply intricate transfer mechanisms. Heat is often supplied in the thickness direction while, in spite of the length, the effect of the pressure gradient on gaseous (important for low moisture content) or liquid migration (important for high moisture content) takes place in the longitudinal direction (Figure 36.20). This is a result of the anatomical features of wood. In the case of very intensive internal transfer, the end piece can be fully saturated and, sometimes, moisture can leave the sample in the liquid state. (This is quite easy to observe during microwave heating.)

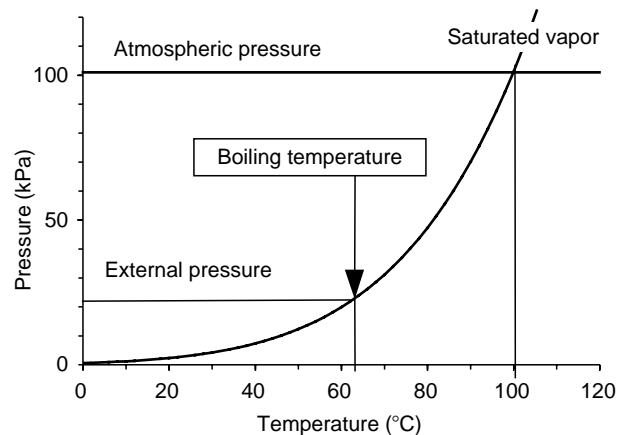


FIGURE 36.19 Vacuum drying seeks to reduce the boiling point of water in order to obtain a high-temperature effect with moderate drying conditions. (saturated vapor pressure values from Lide 1995.)

36.2.3.3 Typical Drying Behavior: Difference between Sapwood and Heartwood

In a tree, freshly cut down and sawn, it is easy to distinguish sapwood from heartwood (by touch or by sight). But a few days later, the loss of surface moisture content makes it impossible to do that. Nevertheless, in

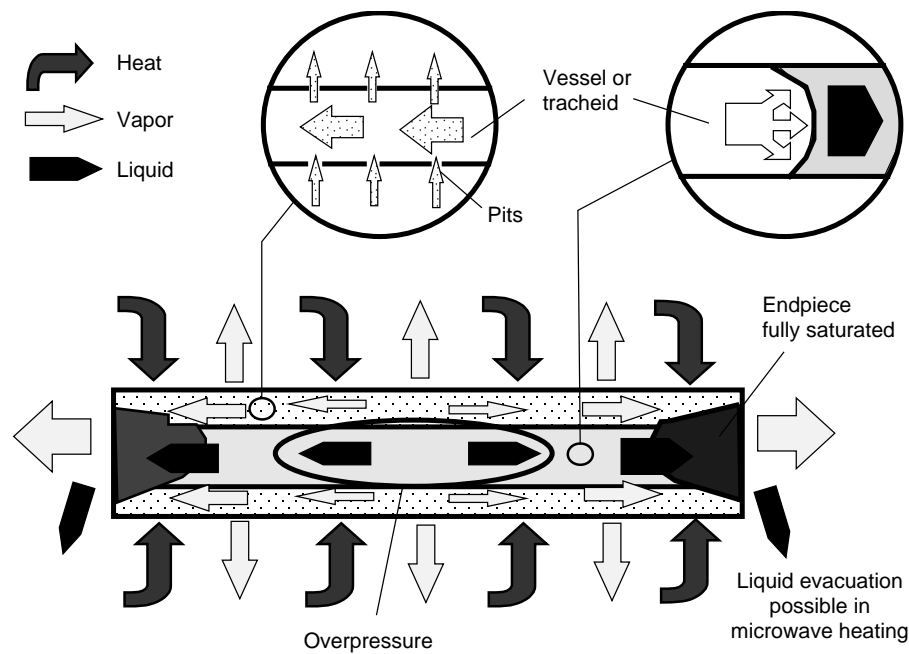


FIGURE 36.20 Drying at high temperature (second drying period): a high-temperature regime means that an overpressure develops inside the medium. Depending on the moisture content, this overpressure induces liquid or gaseous flow; in addition, as wood is strongly anisotropic, the most part of the flow occurs in the longitudinal direction (see the magnified views). (Adapted from Perré, P., The numerical modeling of physical and mechanical phenomena involved in wood drying: an excellent tool for assisting with the study of new processes, Tutorial, *Proceedings of the Fifth International IUFRO Wood Drying Conference*, Québec, Canada, 1996, 9–38.)

the case of high-temperature drying, the increase of internal pressure gives rise to longitudinal migration of liquid toward the end pieces, provided that the permeability and the moisture content are high enough. This is a good way to spot sapwood after a few hours of drying (Figure 36.21b). This phenomenon can be observed in industrial kilns (Figure 36.21c).

To illustrate the effect of these differences on the drying process, Figure 36.22 depicts drying experiments carried out with superheated steam at 150°C on both sapwood and heartwood of Norway spruce

(*Picea abies*). They are representative of the trends observed by different authors (Salin, 1989; Pang et al., 1994; Perré and Martin, 1994).

After the initial transient period, the constant drying-rate period takes place for the sapwood board. During this period, which lasts several hours, all temperatures are equal to the wet-bulb temperature and the overpressure remains very small. At the beginning of the second drying period (around 350 min), an important overpressure develops due to the temperature increase. It disappears only when the

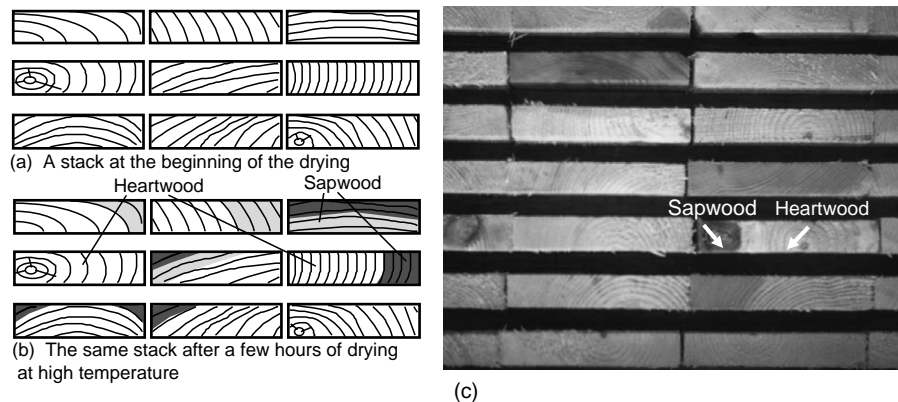


FIGURE 36.21 A stack of boards during high-temperature drying (shaded areas indicate wet zones).

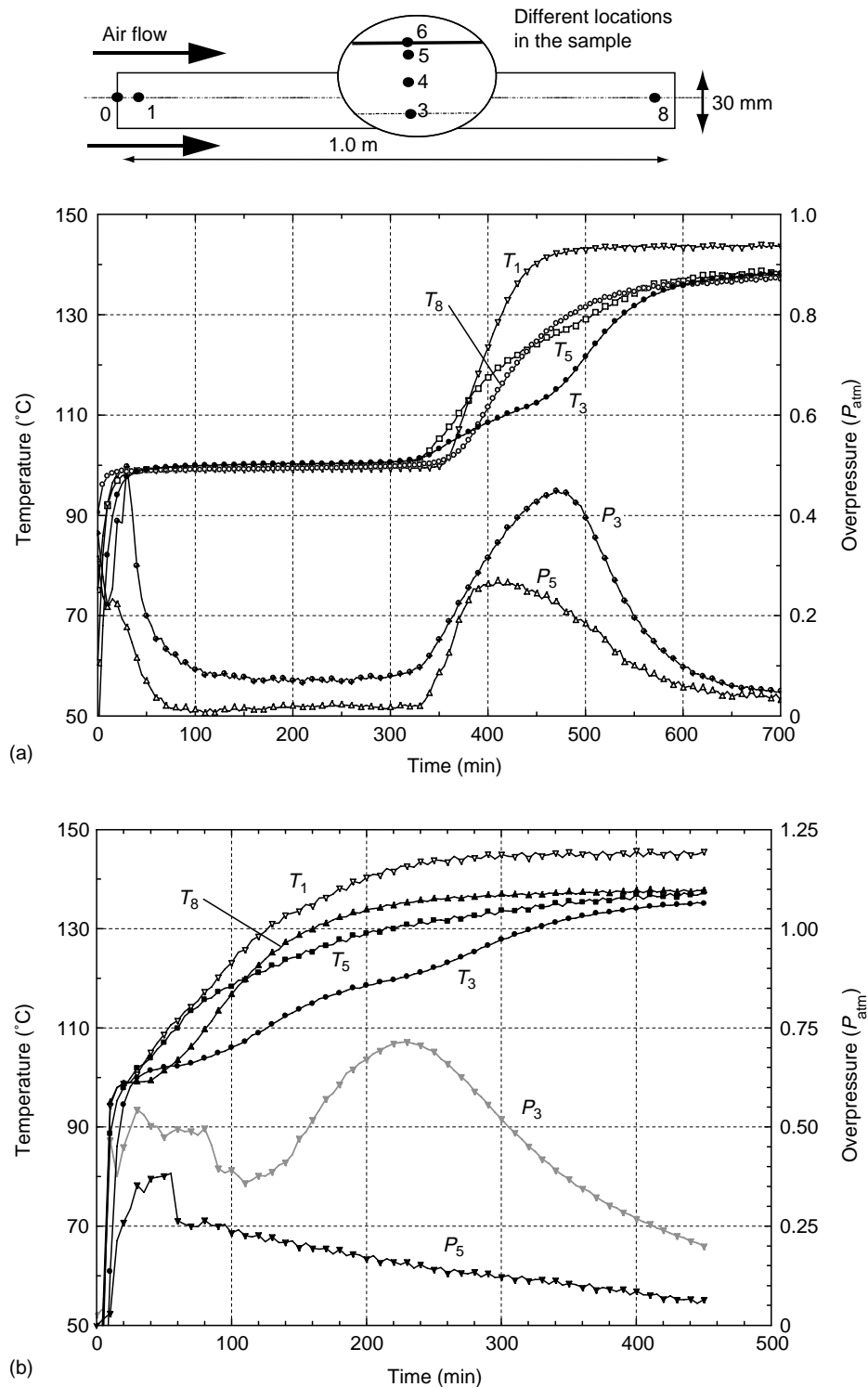


FIGURE 36.22 Experiment on spruce (*Picea abies*) dried with superheated steam at 150°C . Temperature and internal pressure at different locations. Note the difference between sapwood (a) and heartwood (b).

entire board enters the hygroscopic range. At this moment, all temperatures approach the dry-bulb temperature.

The results obtained for heartwood are quite different. No constant drying-rate period can be observed. Just a short plateau at the boiling point

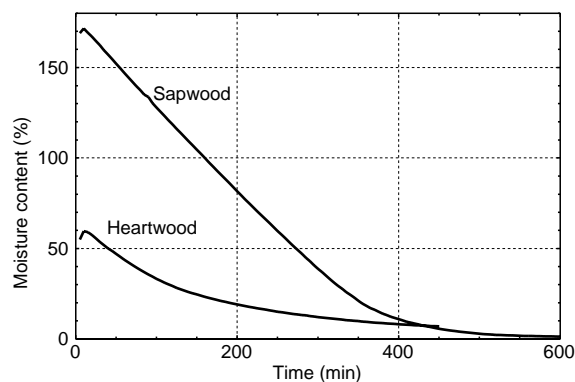


FIGURE 36.23 Moisture content loss obtained for sapwood and for heartwood (same experiments as in Figure 36.22).

is detectable at the rear end of the board (T_8). Consequently, the overpressure remains high (especially for the center pressure P_3) up to the end of the drying. The maximum pressure is higher for heartwood than for sapwood. The differences in drying kinetics are of great interest. In spite of the high initial moisture content of sapwood (170% against 60%), the permeability of heartwood is so low that the curves cross each other at 450 min of drying (Figure 36.23). This is consistent with the observations on entire stacks (Salin, 1989).

The strategy of simulating the differences between heartwood and sapwood lies in only two sets of parameters: the permeability and the initial moisture content (for these experiments, 180% for sapwood and 70% for heartwood). The values of permeability used to differentiate sapwood from heartwood (Table 36.8) are based on the considerations concerning pit aspiration.

By using only these differences, Perré and Turner (1996) found that all the trends observed for sapwood and heartwood were found in the simulated results. The most spectacular effect is the longitudinal flow due to the overpressure (Figure 36.24). In the case of high-temperature convective drying, the sapwood board delivers a large supply of water to the end piece after 5 h, while the heartwood end piece is already within the hygroscopic range. These carpet plots should be compared with Figure 36.21.

36.2.4 MECHANICAL ASPECTS OF WOOD DRYING

36.2.4.1 Mechanical Behavior of Wood

Industrial wood drying consists of not only removing moisture from greenwood but also ensuring that its quality (fitness for purpose) is adequate in end use. Because wood shrinks during drying, deformations and stresses develop that can lead to

TABLE 36.8
Intrinsic Permeabilities Used in the Drying Modeling

Direction	Heartwood	Sapwood
<i>Longitudinal</i>		
Gas	10^{-13} m^2	2.10^{-13} m^2
Liquid	10^{-13} m^2	10^{-12} m^2
<i>Transverse</i>		
Gas	10^{-16} m^2	2.10^{-16} m^2
Liquid	10^{-16} m^2	10^{-15} m^2

unusable products (Figure 36.25). The understanding of these aspects must account for the complex mechanical behavior of wood, including its memory effect.

Shrinkage is the “driving” force for drying stress, i.e., without shrinkage, no drying stresses would develop. Figure 36.26 exhibits the dimensional variation of an unladen sample with the moisture content (the latter is assumed to be uniform). Under normal conditions, the dimensions do not change until the moisture content attains a moisture content close to the acknowledged FSP. This condition is sometimes called SIP. Then, the dimension variations are almost proportional to the change in moisture content. This strain field is called free shrinkage.

A sample subjected to tensile or compressive stress (Figure 36.27) exhibits the instantaneous deformation (elastic part) at first, which then increases with time (viscoelastic creep). After cycling the moisture content to and from a higher moisture content, the creep has been significantly greater due to mechanosorptive action.

Thus, a sample subjected to a compressive stress (as shown in case 1, Figure 36.28) exhibits a smaller length at the end of drying than an unloaded specimen (case 2), which itself has a smaller length than the sample subjected to a tensile stress (case 3). In this experiment, the viscoelastic behavior acts because of time and the mechanosorptive behavior acts because of the removal of water molecules due to drying.

These ideas can now be applied to the drying of lumber boards, which is assumed to be stress-free at the beginning. At the beginning of drying (constant drying-rate period), sap throughout the entire board remains free. No shrinkage occurs; hence, stress buildup is absent.

At the beginning of the second drying period, shrinkage exists close to the exposed surfaces (Figure 36.29a). At this moment, if the section was cut into slices, the outer slices would have a shorter length than the inner ones (Figure 36.29b). This displacement field is not compatible and induces, in the

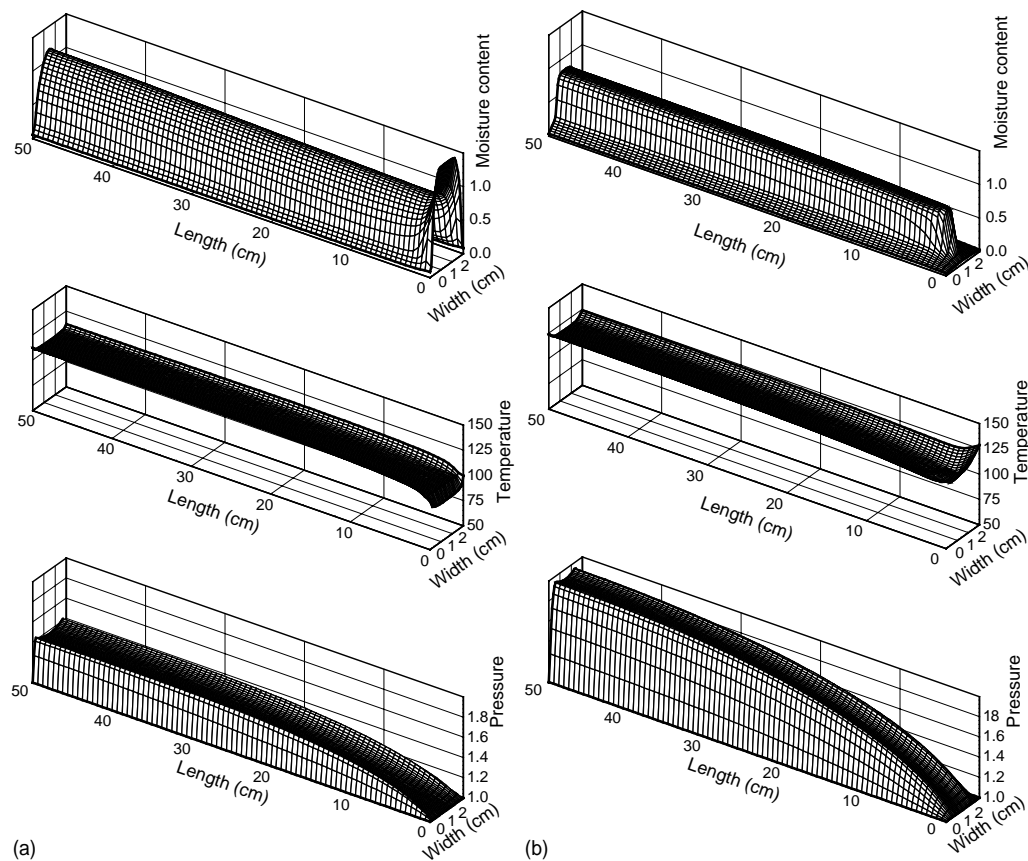


FIGURE 36.24 High-temperature drying (140/85°C). Carpet plot after 5 h of drying. Internal overpressure, resaturation of the end piece, thermal conduction along the thickness, and end piece close to the wet-bulb temperature are evident on these plots. Note the high value of internal pressure and the absence of end-piece resaturation obtained for heartwood (b).

actual section, a tensile stress in the surface layers and (because of equilibrium conditions) a counteracting compressive stress in the core layers (Figure 36.29c). During this period, surface checking is possible. From this point onward, the wood layers dry under load.

As the drying proceeds, viscoelastic creep develops, together with mechanosorptive creep. The outer slices appear similar in configuration to that exhibited for slice n°3 in Figure 36.28, while the internal slices resemble slice n°1. Consequently, in spite of the flat moisture content profile, slicing the section at the end of the drying would give picture Figure 36.30b; the core slices, dried under compression, are smaller than the outer ones, dried under tension. In the actual section, compressive stress exists in the inner part (Figure 36.30c). This phenomenon is known as stress reversal or case-hardening. The residual stress level depends on many parameters (growth history, sawing pattern, drying conditions, species, thickness, etc.), which provide most of the problems of drying optimization. In addition, one must keep in mind that gradients of moisture content, strain, and stress exist along the thickness. This explains the curvature of the slices observed in

prong test or cup method commonly used in industry to assess stress levels (Figure 36.31). When the inner tensile stress is too high, internal checking occurs (Figure 36.25). An interesting simulation of this test can be found in Dahlblom et al. (1994).

36.2.4.2 Drying Stress Formulation

During drying, shrinkage appears in all parts of the board for which the moisture content X is within the hygroscopic range. The shrinkage strain is proportional to the difference between the local moisture content and the local value of the moisture content at fiber saturation at the same temperature. A deformation field noted, ε^{sh} , is defined in the material's axes by Equation S1.

If this deformation field does not fulfill the geometrical compatibility, a strain tensor ε^{mec} related to stress is generated. The constitutive equation, which represents the mechanical behavior of the material, relates this strain tensor ε^{mec} and the stress tensor. Due to the memory effect of wood, this tensor ε^{mec} has to be divided into two parts: (1) an elastic strain, $\varepsilon^{\text{elas}}$, connected to the actual stress tensor and (2) a memory strain, ε^{mem} , which includes all the strain due

to the history of that point (ε^{mem} can deal with plasticity, creep, mechanosorption, etc.).

The geometrical compatibility applies to the total strain field ε^{tot} . When solving the mechanical problems in terms of displacement, the total strain tensor is deduced from the displacement field and this geometrical condition is automatically fulfilled within the domain. The stress field must satisfy the local mechanical equilibrium and the boundary conditions. Finally, the complete formulation of the stress problem is given by Equation S1 through Equation S4.

36.2.4.3 Memory Effect

While describing the strain field, ε^{mem} , lies the entire problem of developing a constitutive model for wood, which requires both theoretical and numerical work. Comprehensive formulations are also very difficult to characterize (Ranta-Maunus, 1975). The problem lies in the fact that the memory effect of wood depends not only on the temperature and moisture content values but also on their variations in time and on the history of their variations in time. This

$$\varepsilon^{\text{sh}} = H(\vec{x}) \begin{bmatrix} A & 0 & 0 \\ 0 & B & 0 \\ 0 & 0 & C \end{bmatrix} \quad (\text{S1})$$

with

$$H(\vec{x}) = \begin{cases} 0 & \text{if } X(\vec{x}) \geq X_{\text{fsp}} \\ X(\vec{x}) - X_{\text{fsp}} & \text{if } X(\vec{x}) \leq X_{\text{fsp}} \end{cases}$$

$$\varepsilon^{\text{mec}} = \varepsilon^{\text{elas}} + \varepsilon^{\text{mem}} \quad (\text{S2})$$

$$\begin{cases} \varepsilon_{ij}^{\text{tot}} = \frac{1}{2}(u_{i,j} + u_{j,i}) & \text{over } \Omega \\ \sigma_{ij,j} + \rho f_i = 0 & \text{over } \Omega \\ \sigma_{ij} = a_{ijkl}(\varepsilon_{kl}^{\text{tot}} - \varepsilon_{kl}^0) & \text{over } \Omega \text{ with } \varepsilon^0 = \varepsilon^{\text{sh}} + \varepsilon^{\text{mem}} \text{ and } \forall i, \Gamma_{D_i} \oplus \Gamma_{T_i} = \Gamma \\ \sigma_{ij} n_j = T_i & \text{on } \Gamma_{T_i} \\ u_i = D_i = 0 & \text{on } \Gamma_{D_i} \end{cases} \quad (\text{S3})$$

Remarks:

- This static formulation requires that boundary and volumetric forces satisfy the global equilibrium.
- Γ is the surface surrounding the domain Ω . Γ_{D_i} refers to the subdomain of Γ where the i component of the displacement is known and Γ_{T_i} to the subdomain of Γ where the i component of the traction force is known. In order to ensure the uniqueness of the solution, additional conditions are required on the boundary conditions: $\forall i, \text{mes}(\Gamma_{D_i}) > 0$. Otherwise, the solution is defined within a rigid body motion.
- As wood is orthotropic, each behavior law involves nine independent terms. In fact, it is more common to define the inverse of a_{ijkl} that, for the case of linear elasticity, leads to the generalized Hooke's law:

$$\begin{bmatrix} \varepsilon_{LL} \\ \varepsilon_{RR} \\ \varepsilon_{TT} \\ 2\varepsilon_{LR} = \gamma_{LR} \\ 2\varepsilon_{LT} = \gamma_{LT} \\ 2\varepsilon_{RT} = \gamma_{RT} \end{bmatrix} = \begin{bmatrix} \frac{1}{E_L} & -\frac{\nu_{RL}}{E_R} & -\frac{\nu_{TL}}{E_T} & 0 & 0 & 0 \\ -\frac{\nu_{LR}}{E_L} & \frac{1}{E_R} & -\frac{\nu_{TR}}{E_T} & 0 & 0 & 0 \\ -\frac{\nu_{LT}}{E_L} & -\frac{\nu_{RT}}{E_R} & \frac{1}{E_T} & 0 & 0 & 0 \\ 0 & 0 & 0 & \frac{1}{G_{LR}} & 0 & 0 \\ 0 & 0 & 0 & 0 & \frac{1}{G_{LT}} & 0 \\ 0 & 0 & 0 & 0 & 0 & \frac{1}{G_{RT}} \end{bmatrix} \begin{bmatrix} \sigma_{LL} \\ \sigma_{RR} \\ \sigma_{TT} \\ \sigma_{LR} \\ \sigma_{LT} \\ \sigma_{RT} \end{bmatrix} \quad (\text{S4})$$

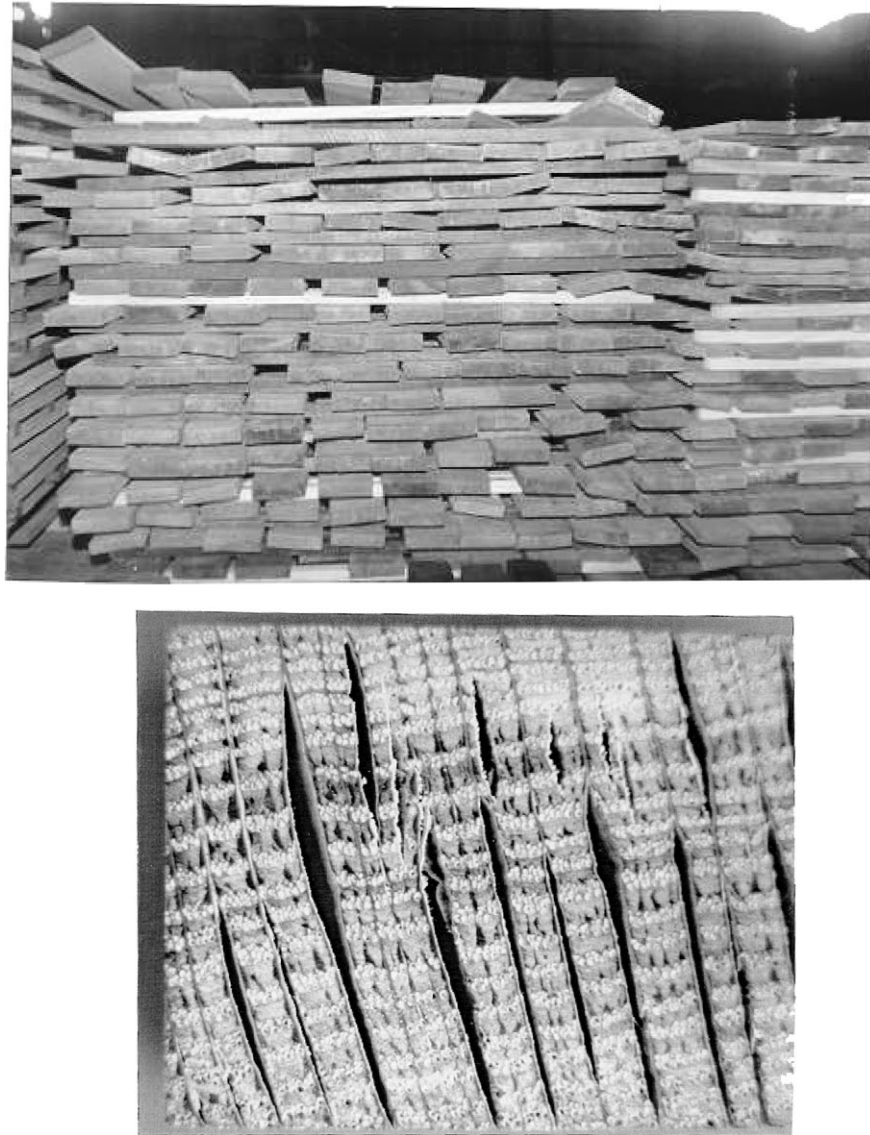


FIGURE 36.25 Two examples of mechanical degrade during wood drying: board deformation and internal checking.

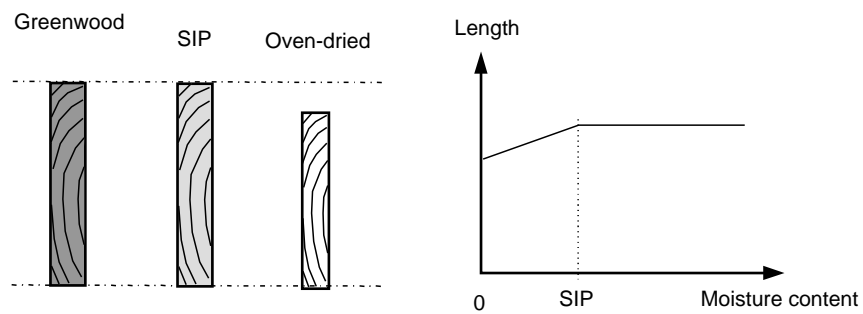


FIGURE 36.26 Wood shrinkage: the shrinkage intersection point (SIP), often close to 30%, depends on species and temperature.

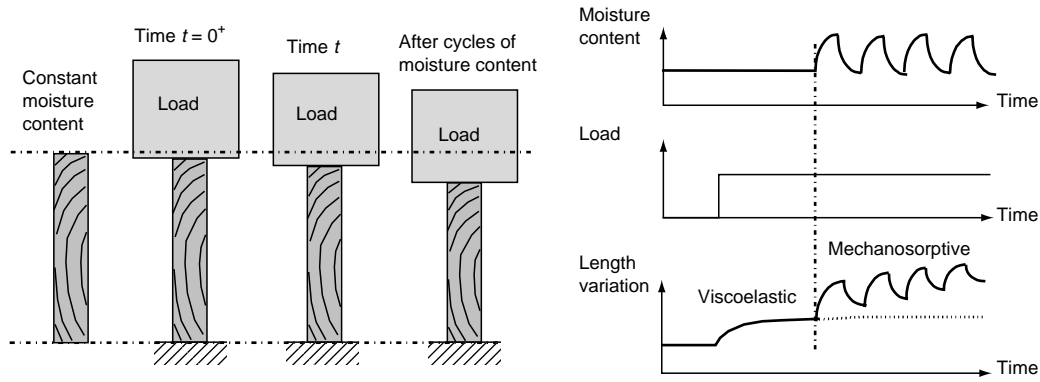


FIGURE 36.27 Viscoelastic and mechanosorptive behavior of wood. (Adapted from Perré, P., The numerical modeling of physical and mechanical phenomena involved in wood drying: an excellent tool for assisting with the study of new processes, Tutorial, *Proceedings of the Fifth International IUFRO Wood Drying Conference*, Québec, Canada, 1996, 9–38.)

subject remains a matter of some scientific debate (Keey et al., 2000).

Nevertheless, in the case of drying, the moisture content only decreases and some simplifications apply. Here, only the most common way to express creep and mechanosorptive effect will be presented. The general formulation of the time dependency of the creep property involves the whole stress history:

$$\varepsilon_{ij} = \int_{-\infty}^t J_{ijkl}(t-t') \frac{d\sigma_{kl}}{dt'} dt' \quad (36.18)$$

where $J_{ijkl}(t)$ is the creep compliance tensor and t is the actual time. The experimental creep function is often analyzed as a number of Kelvin elements in series (Figure 36.32), each having the property of

a spring and dashpot in parallel (Genevaux, 1989; Martensson, 1992; Mohager and Toratti, 1993; Hanhijärvi, 1999 Passard and Perre, 2005). In the case of uniaxial load, this leads to

$$J(t) = J_0 \left(1 + \sum_{n=1}^N a_n (1 - e^{-\frac{t}{\tau_n}}) \right) \quad (36.19)$$

The temperature and moisture dependency of that function can be expressed using a material time or changing the characteristic time τ_n . The thermal activation, for example, is often expressed with the aid of an Arrhenius law:

$$\tau_n = \tau_n^\infty \exp\left(-\frac{\Delta W_n}{RT}\right) \quad (36.20)$$

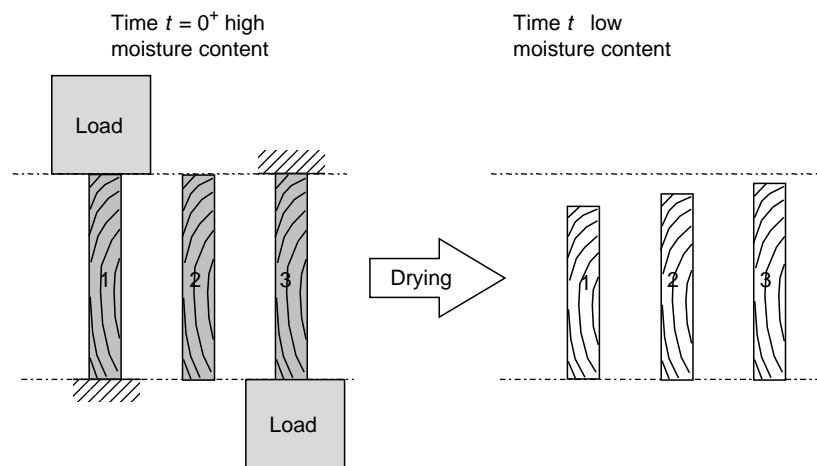


FIGURE 36.28 Dimension changes of a specimen loaded during drying. (Adapted from Perré, P., The numerical modeling of physical and mechanical phenomena involved in wood drying: an excellent tool for assisting with the study of new processes, Tutorial, *Proceedings of the Fifth International IUFRO Wood Drying Conference*, Québec, Canada, 1996, 9–38.)

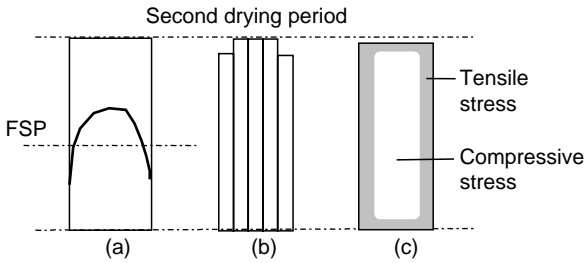


FIGURE 36.29 Appearance of drying stresses following shrinkage. (Adapted from Perré, P., The numerical modeling of physical and mechanical phenomena involved in wood drying: an excellent tool for assisting with the study of new processes, Tutorial, *Proceedings of the Fifth International IUFRO Wood Drying Conference*, Québec, Canada, 1996, 9–38.)

where ΔW_n is the activation energy associated to element n .

The mechanosorptive effect occurs as soon as the moisture content changes during load. The simpler way to express this effect consists in assuming that the strain rate depends linearly on both the stress field and the time derivative of the moisture content \dot{h} :

$$\dot{\epsilon}_{ij}^{ms} = m_{ijkl} \sigma_{kl} (\text{sign } \dot{h}) \dot{h} \quad (36.21)$$

The mechanosorptive strain rate is always in the direction of the stress field, hence the factor $\text{sign } \dot{h}$ in Equation 36.21.

A three-dimensional resolution is very costly in terms of calculation time and computer memory space. The need for such a cost is justified whenever the objective of the calculation lies in evaluating the overall deformation of the board. In this way, the effect of reaction wood, fiber angle, and property

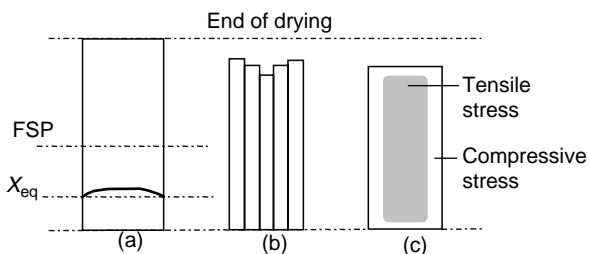


FIGURE 36.30 Stress reversal due to the memory effect of wood. (Adapted from Perré, P., The numerical modeling of physical and mechanical phenomena involved in wood drying: an excellent tool for assisting with the study of new processes, Tutorial, *Proceedings of the Fifth International IUFRO Wood Drying Conference*, Québec, Canada, 1996, 9–38.)

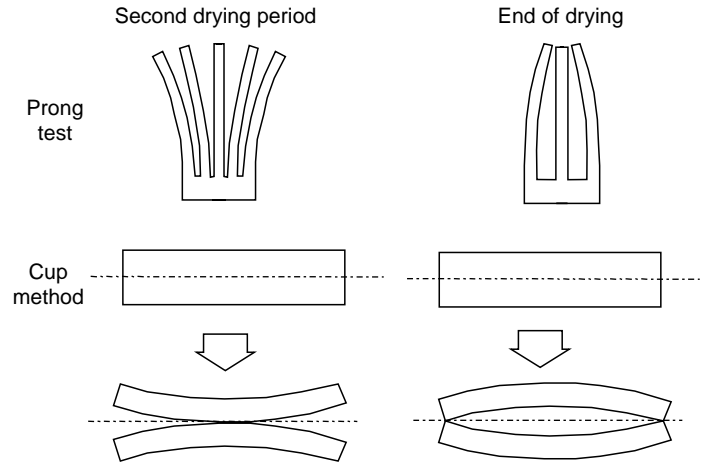


FIGURE 36.31 Two experimental methods used to assess drying stresses.

variations can be analyzed. Good examples of these possibilities can be found in the literature (Dahlblom et al., 1994, 1996; Ormarsson, 1999). Nevertheless, to study the stress development within a section far from the ends of the board, a two-dimensional simulation is sufficient. A “planar displacement” formulation has to be used in this case, assuming small displacements (Perré and Passard, 1995; Chen et al., 1997a) or large displacements (Mauget and Perré, 1999).

36.2.4.4 Stress Development during Drying: Some Examples

All examples presented in this section, except the non-symmetric case, refer to a flatsawn board of heartwood, 20-mm thick and 40-mm wide. They have been computed using the computer code Transpore (Perré and Turner, 1996b and c, 1999; Mauget and Perré, 1999; Perré and Passard, 2002;). [Figure 36.33](#) depicts the moisture and stress fields calculated at different drying stages for quite severe conditions at a medium temperature level ($T_d = 80^\circ\text{C}$, $T_w = 60^\circ\text{C}$). After 6 h of drying, the external part of the section is within the hygroscopic range, which gives rise to tensile stress due to shrinkage in the zones close to the exchange surface. As a consequence of the mechanical equilibrium, internal zones undergo compressive stress. A negative shear stress exists close to the edge (the right angles have decreased). At the end of the drying process (60 h) the moisture content is almost equal to the EMC (7%) throughout the section. The stress reversal due to the memory effect of wood is clearly exhibited. It can be noted that the shear stress also changes in sign.

When a face of a board is insulated, the drying conditions are not symmetrical anymore and

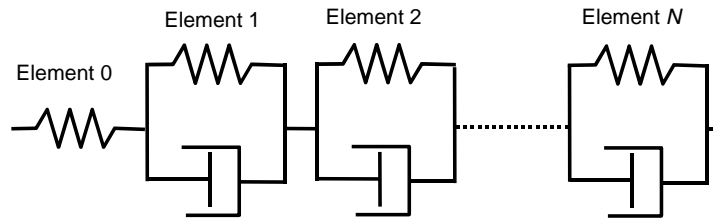


FIGURE 36.32 Modeling the viscoelastic behavior of wood with the number of Kelvin elements in series.

significant section deformations can be observed experimentally (Brandão and Perré, 1996). [Figure 36.34](#) is an example of nonsymmetrical drying. In order to increase the section deformation, a thin quartersawn board has been simulated, here 5 mm by 80 mm. The

large displacement formulation is essential in this case. In this configuration, one part of the drying stress is transformed into section deformation; hence the stress-reversal phenomenon induces a negative final curvature.

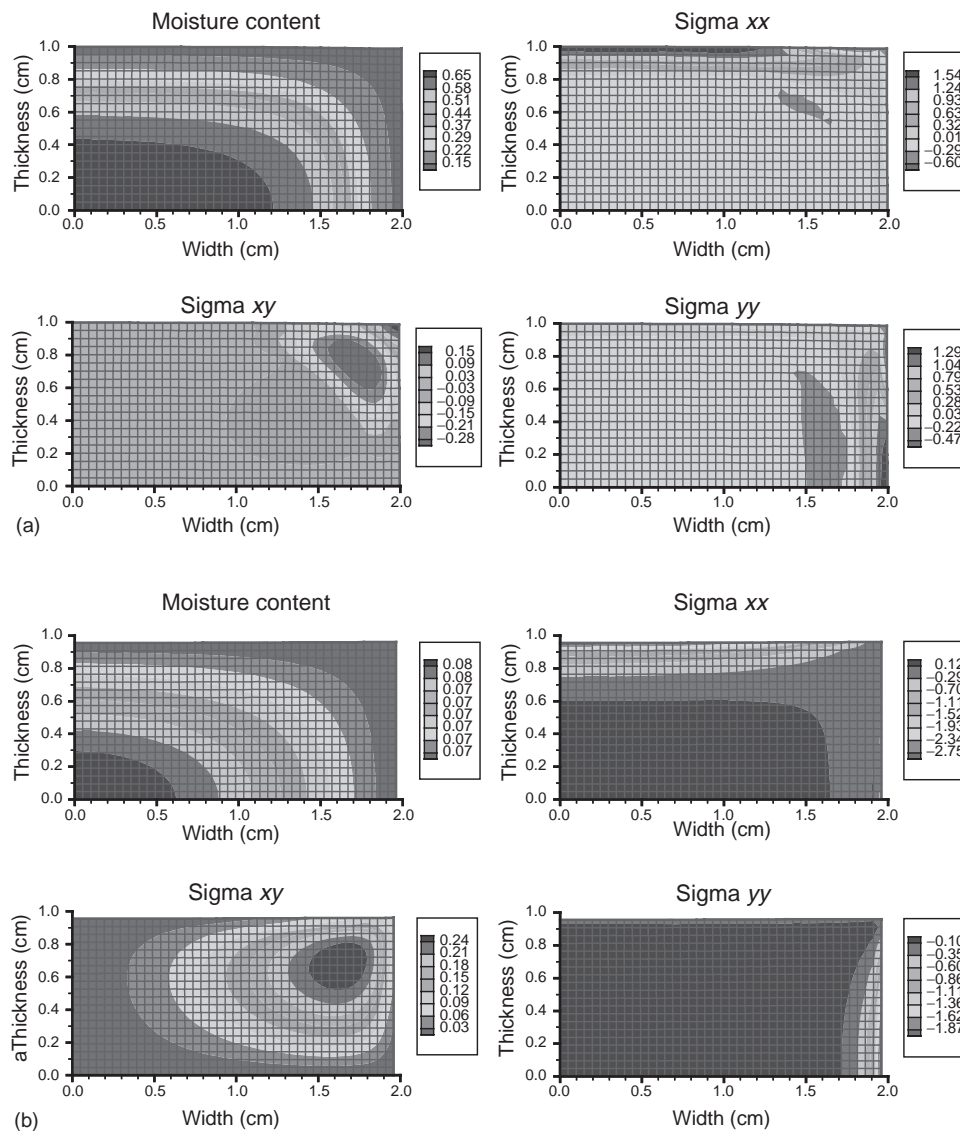


FIGURE 36.33 Moisture content and stress fields after (a) 6 h and (b) 60 h ($T_d = 80^\circ\text{C}$, $T_w = 60^\circ\text{C}$).

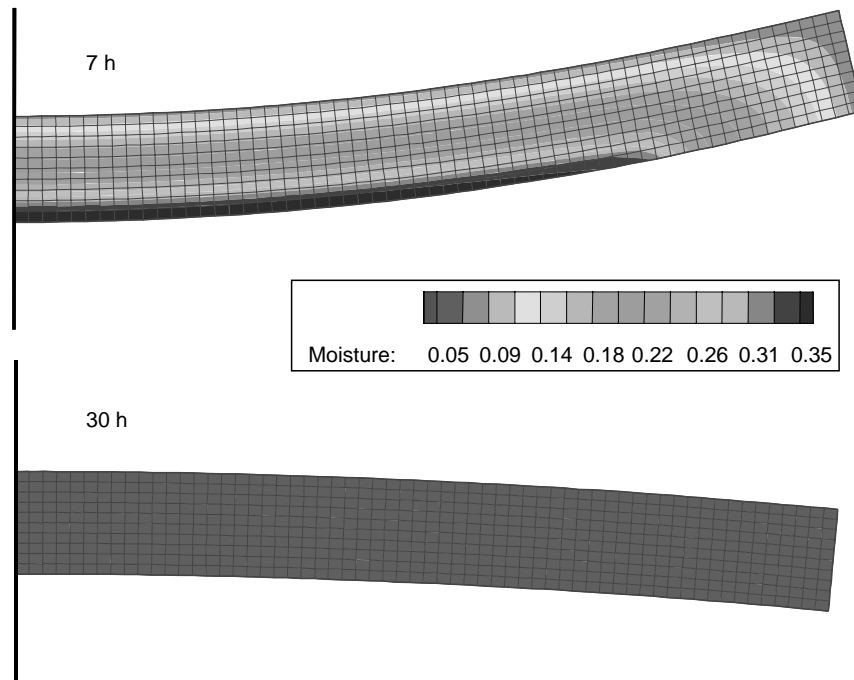


FIGURE 36.34 Example of nonsymmetrical drying: moisture content and section deformation (80/60°C).

The next example illustrates three different constant drying conditions chosen to analyze the possibilities of prediction (T_d stands for dry-bulb temperature and T_w for wet-bulb temperature):

1. $T_d = 40^\circ\text{C}$, $T_w = 35^\circ\text{C}$; mild conditions at low temperature (EMC = 15%)
2. $T_d = 80^\circ\text{C}$, $T_w = 60^\circ\text{C}$; rather severe conditions at medium temperature (EMC = 7%)
3. $T_d = 80^\circ\text{C}$, $T_w = 76^\circ\text{C}$; mild conditions at medium temperature (EMC = 14%)

Figure 36.35 depicts the variations of the averaged moisture content and the stress level (direction parallel to the exchange surface) at different positions vs. time. All tests show a first drying period, without drying stress, then a stage with tensile stress in the peripheral zones, and finally the last drying stage that exhibits the stress reversal. However, the duration of each stage and the stress level depend strongly on the drying conditions.

For the mild drying conditions at low temperature ($T_d = 40^\circ\text{C}$, $T_w = 35^\circ\text{C}$), the drying time is rather important. The first drying period lasts for around 10 h and the stress level is high for both the second drying stage and the final drying stage. The drying conditions are mild concerning heat and mass transfer, while the temperature level is not high enough for the creep field to relax the stress field; the final stress level reveals the importance of the memory effect.

As a consequence of the low relative humidity of air, the second test ($T_d = 80^\circ\text{C}$, $T_w = 60^\circ\text{C}$) is very fast. However, both the maximum tensile stress level and the final stress reversal are important; the rapid external transfer imposes a high moisture content gradient within the board. In addition, the viscoelastic creep is not sufficient to cancel the memory effect. At the beginning of drying, the board temperature is close to the wet-bulb temperature, which is below the glass-transition zone (Geoffroy 1984, Goreng 1963, Salmen 1984, Ostberg et al. 1990, Passard and Perre 2001). At the end of drying, the temperature level is sufficient for greenwood, but not for the dry part of the board to activate the viscoelastic behavior. Consequently, the outer parts, which are close to EMC, are below the glass-transition zone.

In the third test ($T_d = 80^\circ\text{C}$, $T_w = 76^\circ\text{C}$), the difference in moisture content between surface and core remains low. The first drying period lasts an important part of the total drying time. Due to the high value of EMC, the board temperature is always above the softening zone; consequently, all stress levels remain very low. These conditions allow wood of good quality to be obtained relatively free of stress reversal with a moderate drying time (less than 150 h against 400 h for the low-temperature test).

These simulations are in good agreement with nonsymmetrical drying experiments performed on oak (*Quercus rubra*) boards using the same drying conditions (Figure 36.36; Perré, 2001). However, a

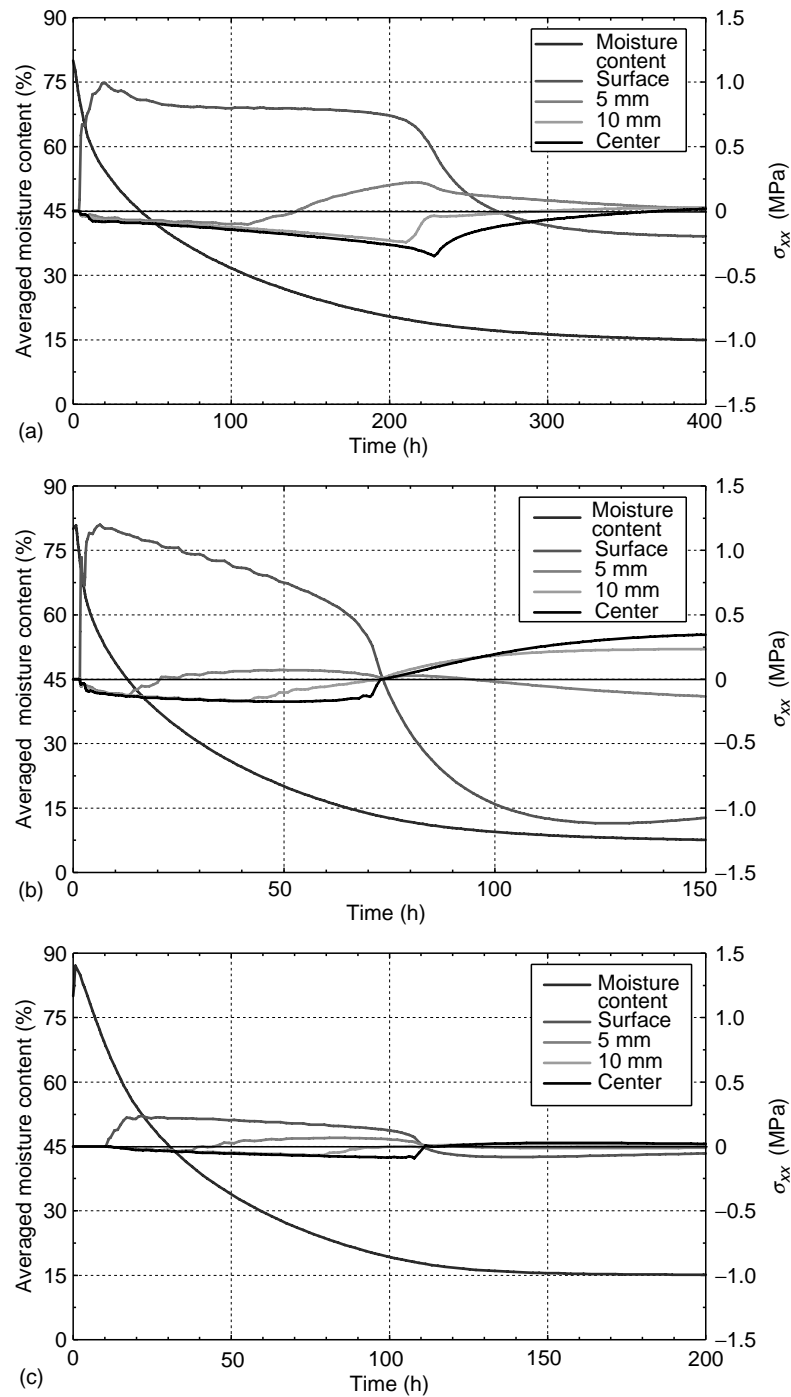


FIGURE 36.35 Averaged moisture content and σ_{xx} vs. time: (a) $T_d = 40^\circ\text{C}$, $T_w = 35^\circ\text{C}$; (b) $T_d = 80^\circ\text{C}$, $T_w = 60^\circ\text{C}$; and (c) $T_d = 80^\circ\text{C}$, $T_w = 76^\circ\text{C}$.

closely similar schedule for another hardwood (*Nothofagus truncata*) resulted in gross deformations and thermal degradation due to the high extractives content of the wood (Grace, 1996).

Based on this reasoning, new drying procedures have been devised and tested on different tropical species, including numerous tests in industrial kiln of 100-m³ capacity (Aguar and Perré, 2000b). The prod-

uct quality was always very good, often with very little checking and rather less deformation than with conventional drying. Most importantly, this method needs only one half to one third of the time required for drying according to conventional schedules.

The code can also be used to test different drying schedules (Figure 36.37). The first one (Schedule A) is recommended for softwoods while the second one

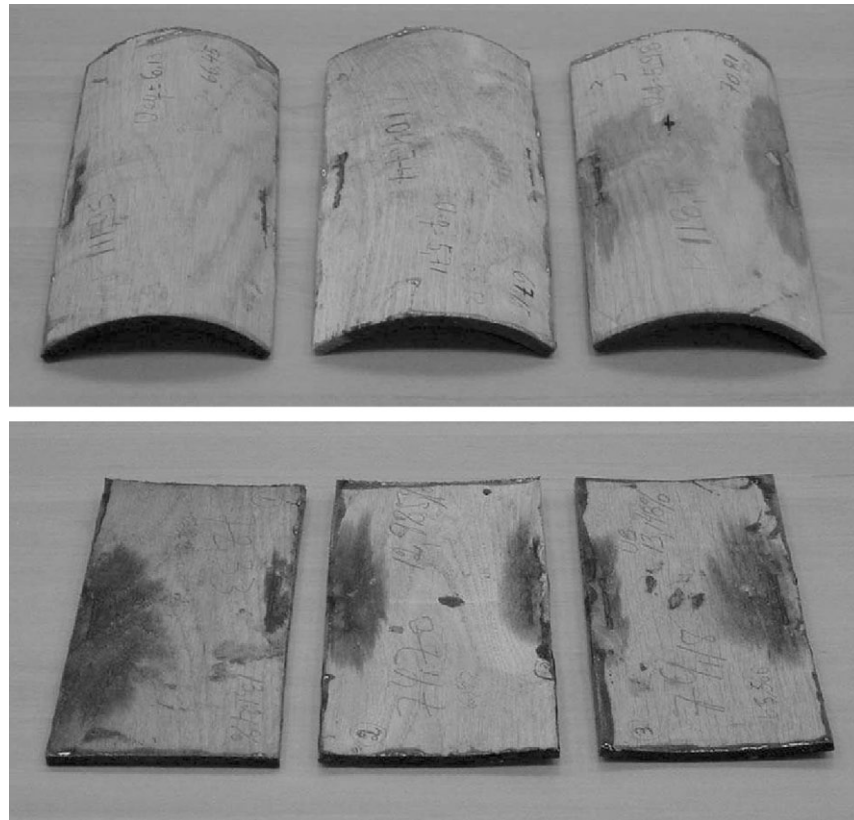


FIGURE 36.36 Final section curvature of oak samples dried on their upper face with conditions n 2 ($T_d = 80^\circ\text{C}$, $T_w = 60^\circ\text{C}$) and n3 ($T_d = 80^\circ\text{C}$, $T_w = 76^\circ\text{C}$), respectively.

(Schedule B) is recommended for hardwoods. Schedule B proposes lower temperature levels and higher relative humidity values. In this drying schedule, EMC decreases significantly only at the end of the process, when the board is supposed to be dry with a low moisture content gradient. As the first consequence of these drying conditions, one can notice a much longer drying time for Schedule B (130 h against 50 h). The first drying period lasts also for a longer time for the second procedure (30 h instead of 10 h). It may be noted that the first drying-period duration represents about the same percentage of the total drying time for both schedules. This remark still stands for the stress level. One can easily consider a relative time (current time over total drying time) at which all curves have the same shape and the same stress magnitude.

In the first example, the advantages to be gained from using a high relative humidity level (low moisture content gradient, high hygroactivation of the viscoelastic behavior) hardly offset the negative effect of the low temperature levels (slow moisture migration and low thermoactivation of the viscoelastic behavior). A careful analysis of these approaches is very promising. New rules can be

derived to improve existing drying schedules or to devise innovative drying procedures.

Schedule A			
Average Moisture Content (%)	Dry-Bulb Temperature (°C)	Wet-Bulb Temperature (°C)	Relative Humidity (%)
Green	71	66	80
50	76.5	68.5	70
30	82	70.5	60
20	88	67.5	40
Schedule B			
Average Moisture Content (%)	Dry-Bulb Temperature (°C)	Wet-Bulb Temperature (°C)	Relative Humidity (%)
Green	40.5	38	85
60	40.5	37	80
40	43.5	39	75
35	43.5	38	70
30	46	39.5	65
25	51.5	43	60
20	60	47.5	50
15	65.5	49	40

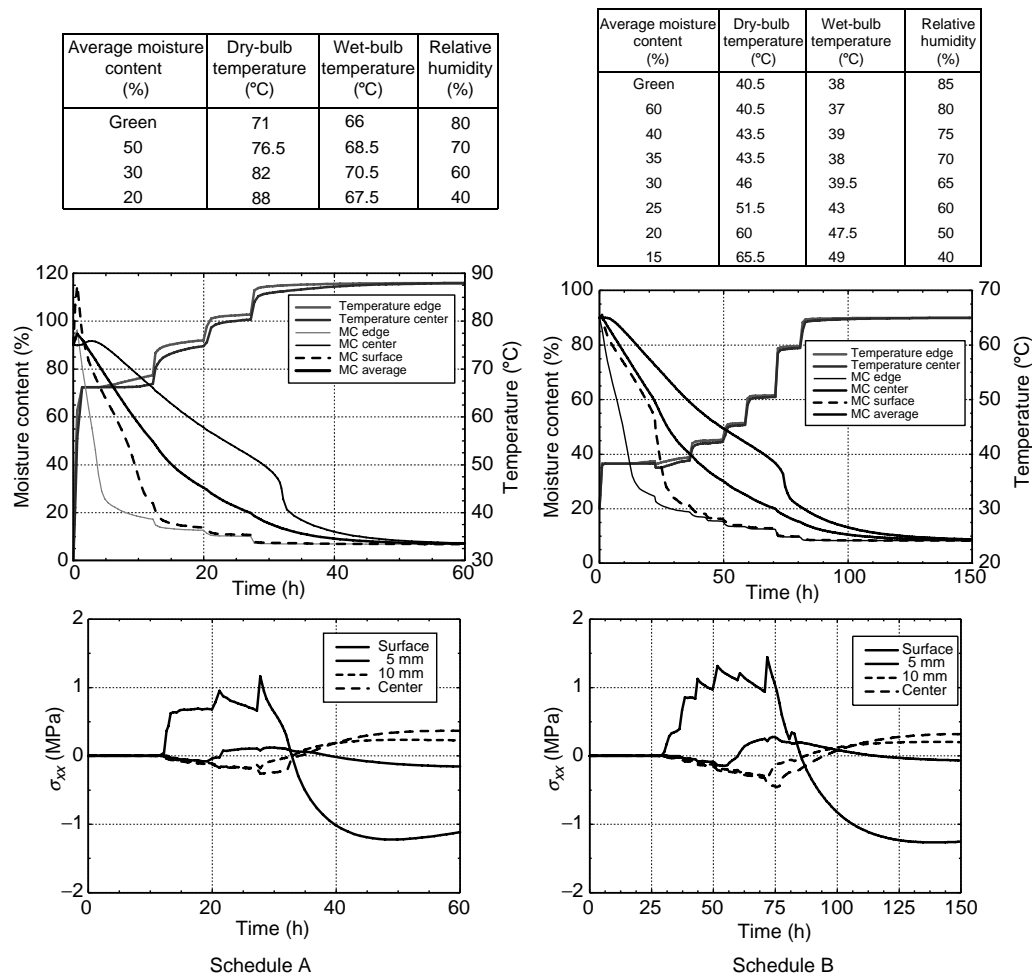


FIGURE 36.37 Simulation of two different drying schedules: moisture content, temperature, and stress level at different positions vs. time.

An alternative procedure in improving kiln schedules has been the estimation of strain levels to provide a safe envelope of dry- and wet-bulb temperatures in kiln operation. One industrial method uses acoustic emissions on sample boards to determine a stress threshold to keep the surface strain under 50 to 75% of the estimated ultimate value (Doe et al., 1996b). Later optimized schedules have been developed by Langrish et al. (1997) using a model predictive control technique to keep within the strain criterion. The technique reduced the number of small- and medium-sized cracks, both internally and at the surface, to less than one quarter of those observed in the original conventional schedule.

36.2.5 DRYING QUALITY

36.2.5.1 Factors Affecting the Drying Duration

Let us assume that the duration of drying is the single important factor. In this case, heat and mass transfer

only are to be considered (Figure 36.38). Some parameters depending on the load are important, but they are not under control:

- The thickness is a very important parameter (roughly speaking, the drying time increases as the thickness doubled)
- The transfer properties of the wood (diffusivity, permeability, capillary pressure, thermal conductivity, etc.)

In conventional drying, the controlled parameters are the dry- and wet-bulb temperatures as well as the velocity of the airflow. These three parameters determine the external heat- and mass-transfer rates:

- The “drying potential” of the air flow is the heat-transfer coefficient (which increases with the air velocity) times the difference between dry- and wet-bulb temperatures.

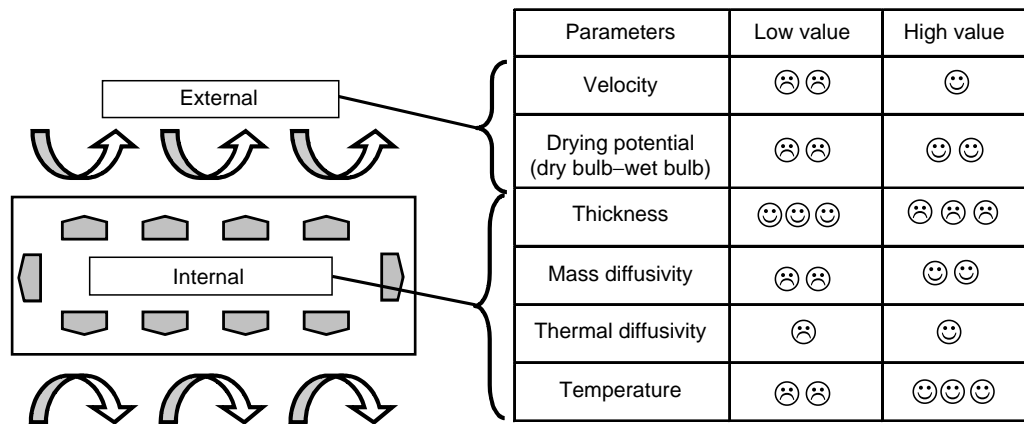


FIGURE 36.38 Guidelines on how to obtain a fast drying operation (from ☺ good to ☺☺☺ excellent and from ☹ poor to ☹☹☹ disastrous). (Adapted from Perré, P., The drying of wood: the benefit of fundamental research to shift from improvement to innovation, *Proceedings of the Seventh International IUFRO Wood Drying Conference*, Tokyo, Japan, 2001, 2–13.)

- The air velocity also plays an important role in uniform drying within the stack. However, its effect becomes less important as the drying progresses when internal transfer mainly controls the moisture migration.

In addition, we have to keep in mind some more subtle effects:

- The internal transfer (diffusion, liquid migration) becomes easier when the temperature level increases.
- Above the boiling point of water, an additional driving force, the gradient of total pressure, acts with a dramatic effect (Perré, 1995). Drying by internal vaporization takes place in such conditions.
- The internal transfer rates depend on the local moisture content (liquid migration is usually much more effective than bound or vapor diffusion). In addition, diffusion becomes very slow when the bound-water content decreases toward zero.

In general, the drying time is reduced when the velocity and the temperature of air are high and its relative humidity is low. However, an excessively low relative humidity may produce a surface zone with low moisture content, thus reducing moisture migration close to the surface. All high-temperature arrangements (convective drying at high temperature, vacuum drying, contact drying, etc.) are processes that accelerate internal moisture migration due to the overpressures generated within the product.

Finally, drying with electromagnetic heating (microwave or radio frequency) offers an entirely

new possibility: any internal temperature can be attained without resorting to a heating medium such as a hot gas.

36.2.5.2 Factors Affecting the Drying Quality

Drying stresses originate from shrinkage; as soon as the shrinkage field within the board is not geometrically compatible, a stress field develops in the material, which is responsible for mechanical degradation. In order to reduce the stress level throughout the process, and thereby the surface checking, the internal checking, and the residual stress, several conditions should be fulfilled (Figure 36.39):

- Low shrinkage coefficients, not under control
- Small thickness, not under control
- Low moisture content values between surface and core
- Retaining important possibilities of viscoelastic creep (mechanosorptive creep is always a source of stress reversal); such an effect is obtained at high temperatures, provided the moisture content is sufficiently high (Irvine, 1984)

It may be noted that a low-temperature level is sometimes desired (for example to avoid collapse), because a high-temperature level may produce thermal degradation or discoloration.

36.2.5.3 Criteria for Obtaining a Fast and Good Drying Process

A fast and good drying process should incorporate the criteria listed in Section 36.2.5.1 and Section 36.2.5.2,

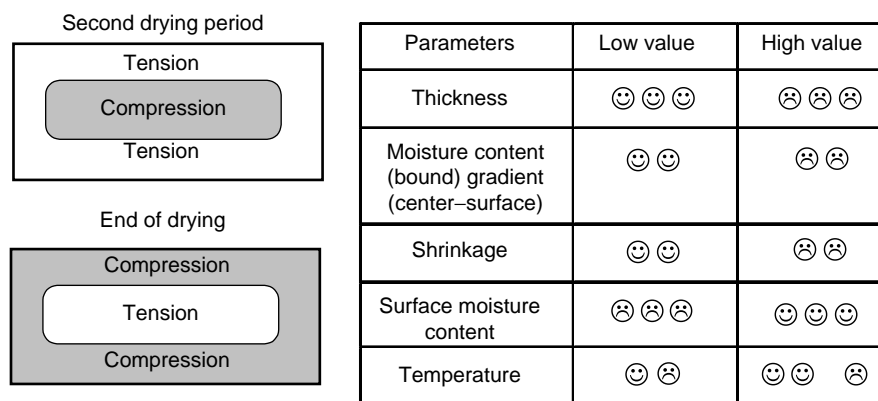


FIGURE 36.39 Guidelines on how to obtain a good-quality product (from ☺ good to ☺☺☺ excellent and from ☹ poor to ☹☹☹ disastrous). (Adapted from Perré, P., The drying of wood: the benefit of fundamental research to shift from improvement to innovation, *Proceedings of the Seventh International IUFRO Wood Drying Conference*, Tokyo, Japan, 2001, 2–13.)

which are, to a large extent, contradictory. Numerous mechanisms involved during drying have to be considered (Figure 36.40).

Because of this complexity, compromises have to be found. Nevertheless, some general rules can be listed (Figure 36.41):

- **Rule 1: high relative humidity.** To ensure a low moisture content gradient, one way is to reduce the drying potential (wet-bulb depression) as much as possible. In addition, this condition imposes a relatively high value of EMC (only one part of shrinkage is effected and the influence of temperature on the viscoelastic creep is not inhibited by a relatively low moisture content level). However, a high relative humidity value can activate the development of fungi.
- **Rule 2: high temperature.** A high value of temperature is most often a positive factor. This accelerates the internal moisture transfer and activates the viscoelastic creep. However, care should be taken with sensitive species; high temperature levels can increase the risk of collapse, problems of color, or even thermal degradation of the wood constituents.
- **Rule 3: high air velocity.** A high air velocity promotes good uniformity of drying throughout the stack. However, a higher velocity increases the electricity consumption and may produce, by the heat-transfer coefficient, an excessively high external transfer flux, which is opposite to the effect intended in Rule 1.

Concerning moisture transfer, Rule 1 and Rule 2 mean that internal transfer has to be increased whereas external transfer should be reduced. Exceeding the

boiling point of water is decisive for internal transfer. However, these rules also require the temperature to be high with a high value of relative humidity. Such conditions may be difficult to ensure for certain dryers. Innovative drying procedures may need new dryer designs. Finally, too often, the effect of temperature and moisture content on the viscoelastic behavior is disregarded in the optimization of drying schedules. The situation strongly differs from one species to the other. Usually, softwood species are quite easily dried. On the other hand, hardwoods are often intractable because of their low permeability.

36.3 KILN SCALE

36.3.1 LUMBER QUALITY

The ultimate fitness for the purpose of dried lumber depends not only on the chosen drying conditions but also on the lumber quality itself. This quality may be thought of in terms of gross defects such as knots as well as intrinsic wood properties such as the degree of anisotropy. Drying, which causes anisotropic shrinkage, interacts with various wood features in various ways. The objective of kiln seasoning, then, is to acknowledge this interaction by setting process conditions that yield dried lumber to the specifications in terms of a grade for an end use. There is a world of difference between drying decorative hardwoods and drying structural softwoods.

Increasingly, kiln operators are drying wood from ever younger, fast-growing stands rather than from mature, old-growth forest. The drying behavior of this new kind of wood is requiring operators to adapt traditional processes on the basis of better understanding of the drying mechanism, as outlined in the previous sections.

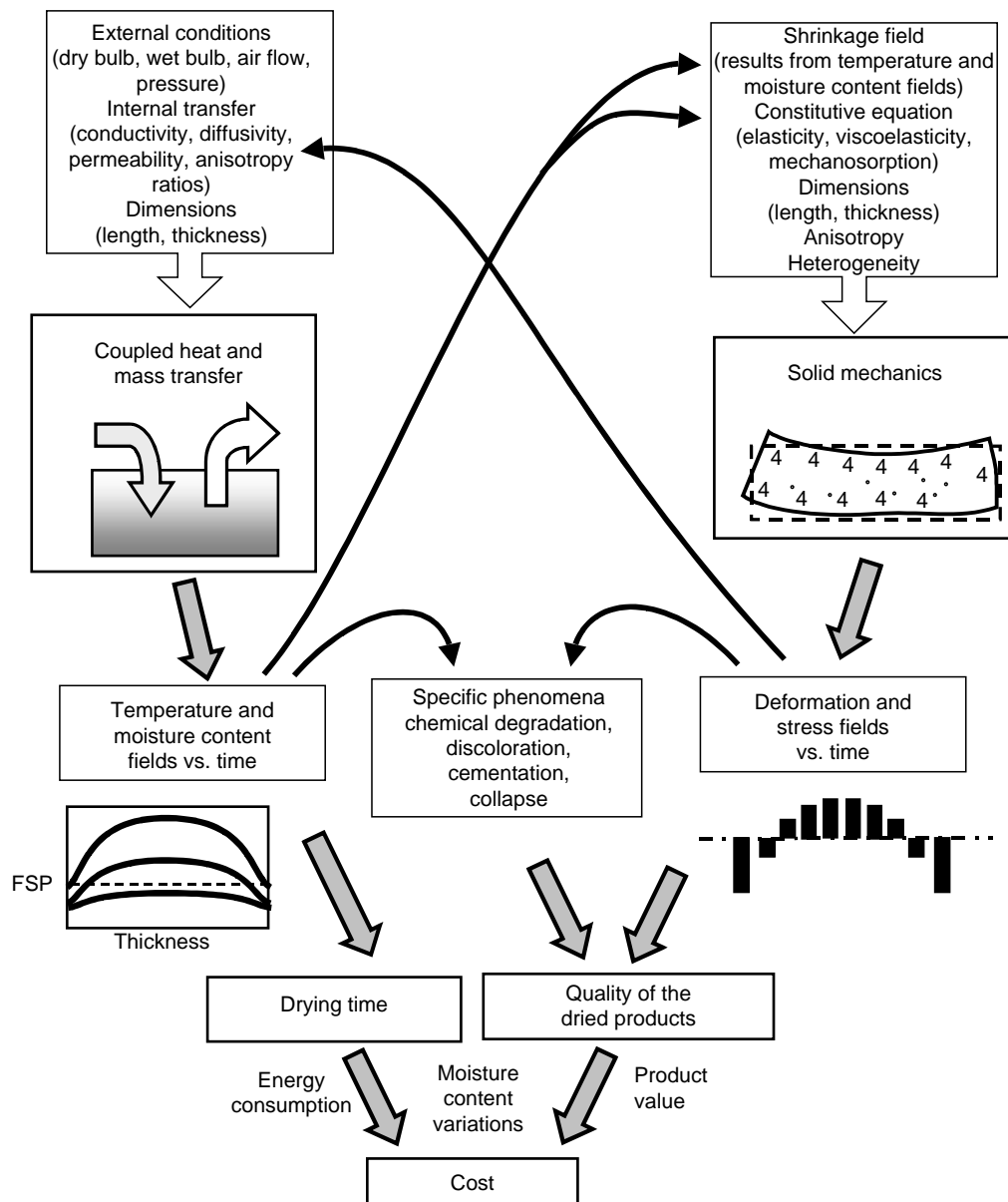


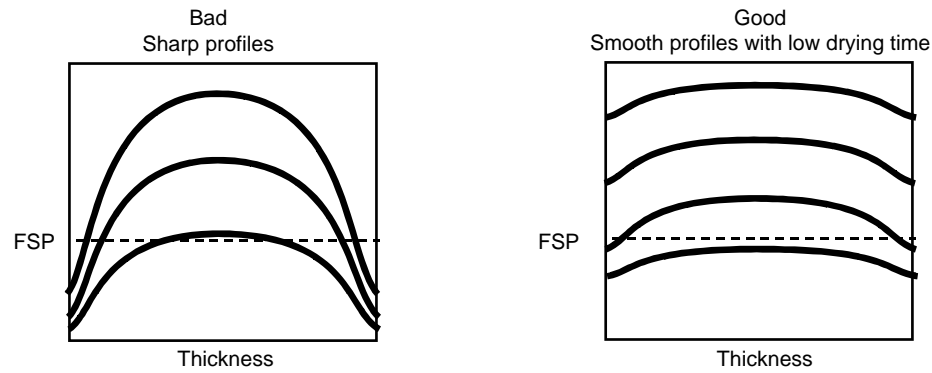
FIGURE 36.40 Some of the phenomena involved in the drying quality. (Adapted from Perré, P., The drying of wood: the benefit of fundamental research to shift from improvement to innovation, *Proceedings of the Seventh International IUFRO Wood Drying Conference*, Tokyo, Japan, 2001, 2–13.)

36.3.1.1 Gross Features of Wood

Most hard pines and Douglas-fir (*Pseudotsuga menziesii*) show large differences in green density and the amount of water in logs, varying with age and height up the stem. Figure 36.42 illustrates these variations for *P. radiata* (Cown, 1992). The large amount of water in the younger trees and the top logs of older trees add to both the costs of transport from the forest to the mill and the energy costs of kiln

drying. Generally, the green moisture content in softwoods lies in the range of 150 to 200% in sapwood and in the range of 30 to 80% in heartwood.

However, unlike softwoods, the green moisture contents of sapwood and heartwood of hardwoods are roughly comparable, and there is little variation with age and position in the tree's stem (Forest Products Laboratory, 1999). Depending on the species and its density, the green moisture content for hardwoods ranges from 50% to about 100%.



Parameters	Value	Why?	Problems
Drying potential (dry bulb–wet bulb)	Low	Unique way obtain a low moisture content gradient Imposes a high equilibrium moisture content (few shrinkage, high creep)	May be difficult to ensure for certain dryers Other problems (coloration, fungi)
Temperature	High	Easy internal transfer + high viscoelastic effect	Can increase the risk of collapse
Velocity	High	Good homogeneity throughout the stack	Electricity consumption

FIGURE 36.41 Guidelines on how to obtain a fast drying operation together with a good product quality. (Adapted from Perré, P., The drying of wood: the benefit of fundamental research to shift from improvement to innovation, *Proceedings of the Seventh International IUFRO Wood Drying Conference*, Tokyo, Japan, 2001, 2–13.)

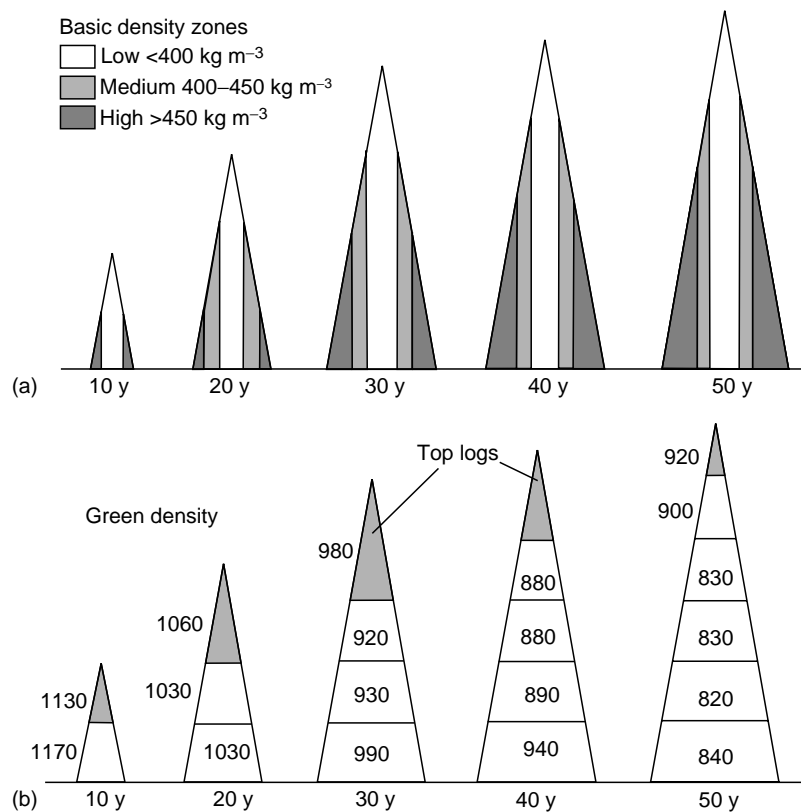


FIGURE 36.42 (a) Density zones, (b) within-tree variations in green density for *Pinus radiata* of various ages. (Adapted from Cown, D.J., New Zealand radiata pine and Douglas-fir suitability for processing, *FRI Bull.*, 168, NZFRI, Rotorua, 1992.)

The term wetwood refers to the discoloration that occurs with high moisture contents in both soft- and hardwoods. Wetwood differs from the wood held in wet storage, as it is a slow-drying tissue confined mainly to heartwood. An acrid odor is associated with anaerobic and other bacteria that can live with very little oxygen in the wetwood, which is found with ring failure and frost damage in the standing forest. Weakened cell walls enhance the susceptibility of the wood to collapse and check, even under mild schedules. The practical difficulty in drying wetwood lies in the danger of overdrying the bulk of lumber.

In the drying of softwoods, it may be important to segregate heartwood from sapwood because of the large differences in moisture saturation and drying behavior. In spite of the high difference in the green moisture content, the drying time can sometimes be comparable for sapwood and heartwood (Salin, 1989; Perré and Martin, 1994). However, with species like *P. radiata*, even though the heartwood is much less permeable than the sapwood, the latter takes twice the time to dry (Haslett, 1998). The presence of heartwood influences the drying of lumber in other ways. Sometimes, for some tropical hardwood species, there is reduced shrinkage, a desirable feature, due to the bulking of cell walls by extractives. There are also greater volatile emissions from the kiln's exhaust, especially in high-temperature drying, with implications for meeting ever increasing environmental and health standards.

Although many lumber-grading procedures presume that each knot is a defect and that boards are graded on the basis of the extent of clearwood, many highly valued decorative woods contain knots and other blemishes. Small knots, which may add character to the wood, may create no difficulties in drying, whereas large knots or knotty clusters can cause problems in both drying and subsequent manufacturing. Whenever visual characteristics are unimportant, such as the requirements for structural timber, knots and knotholes are of little concern.

Large dead knots (where the growing stem has encased a dead branch over the years), however, can be a problem, either falling out during drying or causing difficulties in later millwork. The width of checks associated with knots increases with the knot diameter, and chipping out of knots during machining has been found to treble the level of board rejection for unacceptable knot checks (Haslett, 1998).

The gradual exhaustion of natural forest resources has forced the industries to accept juvenile wood from plantations and second-growth forests. Although juvenile wood is often considered unsatisfactory, resulting in inferior solid wood products, some juvenile wood, being of higher density, can be of better quality than some mature wood. Some of the deficiencies of juvenile wood for fast-grown pines are set out in Table 36.9, ranked subjectively in the decreasing order of significance.

Similar differences between mature and juvenile wood occur in hardwoods. Further, large growth-stress gradients are often present in young hardwood trees. To reduce this problem, it is wise to prolong the rotation of plantation hardwoods until the butt-log diameter is at least 0.6 m. (Keey et al., 2000). Although juvenile wood is still there as long as there are living branches where the wood is formed, the proportion of such wood is much less in the larger-diameter trees.

36.3.1.2 Intrinsic Features of Wood

Wood density is regarded as the easiest and most reliable measure of wood quality. Data from the drying of 106 North American hardwoods show that the volumetric shrinkage of wood increases with its basic density (Stamm, 1964). There is also a correlation between the permeability and the basic density for the sapwood of a softwood (Nijdam and Keey, 2000). Presorting a sapwood load into high-density and low-density groups can reduce the moisture variability of the final recombined kiln-dried boards.

TABLE 36.9
Undesirable Features in Juvenile Wood of *Pinus* spp. as They Affect Solid Wood Quality

Features	Effects
Above-average amounts of compression wood, especially in butt log	Problems with crook, bow, and stiffness
Large microfibril angle in initial growth rings, especially in butt log	Problems with crook, bow, and stiffness
Severe spiral grain in the first few growth rings, especially above butt log	Problems with twist
Above-average hemicellulose content	Enhances any instability in lumber
High moisture content in sapwood of young trees	Higher transport and drying-energy costs
Lower basic density	Increased potential for collapse during high-temperature drying

Source: Keey, R.B., Langrish, T.A.L., and Walker, J.C.F., *The Kiln-Drying of Lumber*, Springer, Berlin, 2000.

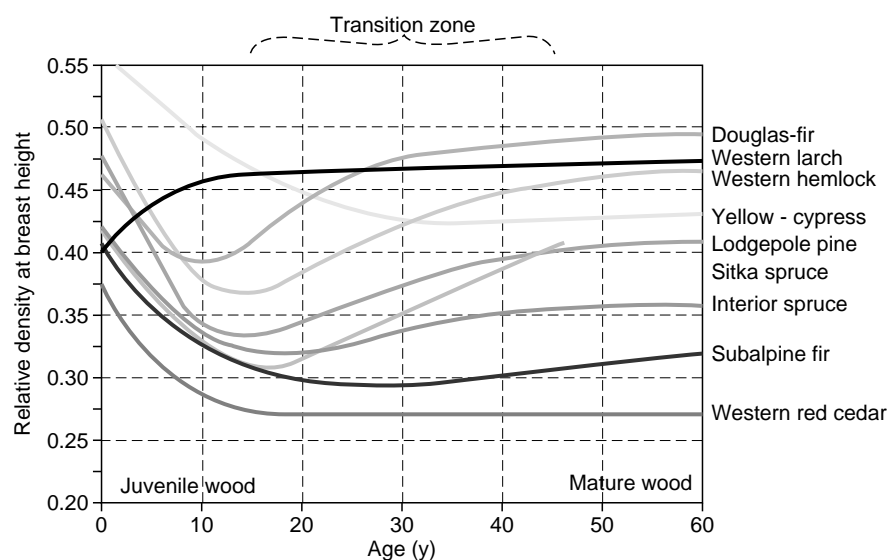


FIGURE 36.43 Trends in basic density at breast height for commercial second-growth softwoods in British Columbia. (Adapted from Josza, L.A. and Middleton, G.R., *A discussion of wood quality attributes and their practical implications, Forintek Canada Special Publ. SP-34, 1994.*)

In practice, however, it is not easy to distinguish the effect of density with that of moisture content.

Within-ring density variations can cause problems because of the differential shrinkage at the ring boundary. Subsequently, severe drying stresses may cause deformation and internal checking (Booker, 1994). Further, a low absolute density in earlywood can result in collapse on drying, particularly under high-temperature conditions (Booker, 1996).

Within-tree variations in density can be highly significant. Cown and McConchie (1983) show that the density in a 24-year-old radiata pine tree can vary from 300 kg m^{-3} in the top log to greater than 450 kg m^{-3} in the outer wood of the butt log. Consequently, the drying kinetics of boards taken from the same log may be markedly different (Davis, 2001). Trends in

basic density for a number of second-growth softwoods are illustrated in Figure 36.43.

The situation is more complex with hardwoods. The growth rate has little effect on the wood properties of diffuse-porous hardwoods, but has a marked impact on the density of ring-porous hardwoods. Unlike softwoods, these produce denser wood when fast grown.

Regardless of the species or where the forests are established, the variation in wood properties between trees is very great and can be great even in boards sawn from the same tree. In particular, the social status of the tree in the stand (whether dominated or dominant trees) has a great effect on the growth rate in diameter and the occurrence of reaction wood. Table 36.10 lists some of the characteristics of

TABLE 36.10
Characteristics and Properties of Reaction Wood Compared with Corresponding Normal Wood

Features	Compression Wood	Tension Wood
Physical characteristics	Darker in color and very hard	Darker in color and silvery sheen in most temperate hardwoods
Density	10–100% greater	10–30% greater
Longitudinal shrinkage	Order of magnitude greater (up to several fold)	About fivefold greater
Warp on drying	Liable to warp badly	Can warp and is liable to collapse
Strength	Comparable strength, does not reflect higher density	Superior strength

Source: Keey, R.B., Langrish, T.A.L., and Walker, J.C.F., *The Kiln-Drying of Lumber*, Springer, Berlin, 2000.

reaction wood produced by environmental factors in the forest. Compression wood has been found to have a significant effect on drying, with lower drying rates over the moisture content range of 40 to 100% for boards of *P. radiata* (Davis et al., 2002). This reduction is attributed to the lower permeability of the compression-wood zone, with its denser wood and more thick-walled cells.

Large microfibril angles are found in both compression wood and normal wood near the pith, inducing larger than normal longitudinal shrinkage and a greater tendency to warp. The longitudinal shrinkage of tension wood, although larger than normal wood, is less than in compression wood.

36.3.1.3 Sawmilling Strategies

Timell (1986) describes sawmilling strategies to reduce warp in juvenile lumber and compression wood. The cutting pattern influences the quality of drying either by releasing stresses on sawing or the cutting induces a stress pattern in the board that can be balanced on drying. Warp has become more of a problem with harvesting from second-growth forests of short rotation, which contain proportionately more juvenile wood than lumber from old-growth forests. Crook and bow are most severe in the core wood of the butt log and where compression wood is encountered; whereas twist is most severe in the core wood of the upper logs where spiral-grain angles are large and changing rapidly. The variation of properties about the mean is the critical factor with core wood.

Vázquez (2001) examines silvicultural practice and sawmilling strategies to counteract the effect of growth stresses in fast-grown Iberian eucalypts (*Eucalyptus globulus*). A model of the stress distribution that enables to determine the deformations on sawing

is given. The appearance of checks and warps can be limited by the choice of a suitable sawing pattern, as shown in Figure 36.44. Pang and Haslett (2002) note that the residual drying stresses in quartersawn *P. radiata* boards are less than flatsawn boards, particularly under low-temperature drying conditions when the effect of mechanosorptive stress relief is relatively minor. The difference in behavior is attributed to the lesser shrinkage in the width direction with quartersawn boards.

36.3.2 KILN DESIGN

Although there are differences in detail between manufacturers, a lumber kiln is essentially a special-purpose room fitted with overhead fans for circulating the drying air and heating coils for maintaining the air (and thus the wood) temperature at the set levels. The moisture in the air is controlled by means of opening the vents in the kiln's roof, thus governing the amount of moist air that returns to the fan which is to be mixed with the fresh air drawn in. Although many kilns are operated batchwise for ease of controlling the drying conditions, which may change through out the drying schedule, a kiln can be continuously worked by arranging the lumber to be slowly railed through the chamber. In this latter case, the drying schedule is maintained by varying the temperature and humidity settings along the length of the kiln. There is some increase in interest in continuous kilns, which might provide energy savings and better quality control (McLean, 2003).

The lumber is stacked externally in a rectangular pile on a low, flat-bed trolley, with rows of boards separated by wooden stickers of uniform thickness to provide duct-like spaces between the boards for the kiln air to flow through. The boards may be stacked

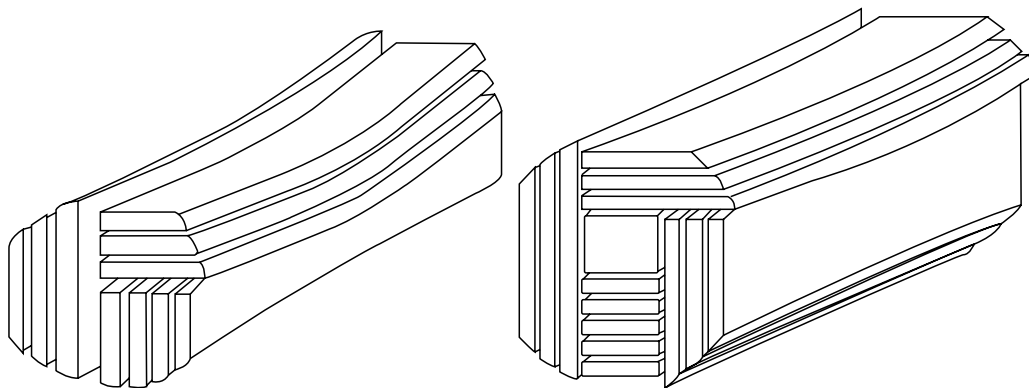


FIGURE 36.44 Sawing pattern to limit the appearance of checks. (Adapted from Vázquez, M.C.T., Tensiones de crecimiento en *Eucalyptus globulus* de Galicia (España). Influencia de la silvicultura y estrategias de aserrado (Growth stresses in *E. globulus* from Galicia (Spain). Influence of silviculture and sawing strategies), *Maderas: Ciencia Tecnología*, 2(1), 68–89, 2001.)

in separate packages on bearers and loaded into the kiln by a forklift vehicle. The boards are butted up, with their long faces incident to the airflow. The stack is squared off as far as possible to provide a uniform resistance to the airflow, and thus minimize variations in drying throughout the kiln. In Scandinavian practice, kiln stacks are normally built from boards of random length, so that every second board is placed flush at one end of the stack and the other boards flush at the other end (Salin, 2001). Kilns may be single-tracked with a 2.4-m wide stack, or twin-tracked with two stacks side by side, to yield a double-width stack of 4.8 m. Figure 36.45 illustrates a vertical cross section through a single-tracked, batch kiln.

To obtain a uniform air distribution to the air-inlet face of the stack as possible, the plenum spaces at each side of the stack must be sufficiently wide (Nijdam and Keey, 2000). An internal ceiling directs the air through the lumber stack or stacks, with baffles or curtains to direct the airflow through the lumber pile. Inward-swinging baffles and contoured, right-angled bends to the plenum space from the ceiling zone improve the uniformity of this airflow (Nijdam and Keey, 2002). Bypass of air around the stack is minimized by the fitting of side baffles or curtains. The kiln is designed so that the pressure loss through the heating coils and other ancillary fittings is small compared with that through the stack of lumber.

36.3.2.1 Airflow Considerations

Kröll (1978) reports the air distribution in a batch dryer fitted with 15 shelves, a heat exchanger, and a fan above a false ceiling, which thus resembles a box-type lumber kiln in a number of respects. In the fifth gap from the top, the air velocity was over twice the average, whereas in the top gap there was a small backflow with air streaming toward the inlet. The flow reversal appeared to be the result of a vortex generated at the top of inlet plenum below the half-circle bend out of the ceiling space. Although such air maldistribution may be extreme in a modern lumber kiln, existing kiln designs may still yield a nonuniform airflow through well-stacked lumber loads, as may be inferred from the kiln audits reported by Nijdam and Keey (1996). Haslett (1998) recommends that the coefficient of variation across the outlet face of the stack should not exceed 0.12 at velocities through the stack between 4.5 and 8 m s⁻¹ when the dry-bulb temperature is greater than 90°C.

Industrial rules of thumb generally equate the ceiling-space height to the plenum-space width and to the combined sticker-spacing height. Hydraulic tests on a model kiln confirm the former rule (Nijdam and Keey, 1999), and the latter is verified by a pressure-drop analysis (Nijdam, 1998).

The transverse gaps between the boards are not simple smooth ducts, but there may be irregularities

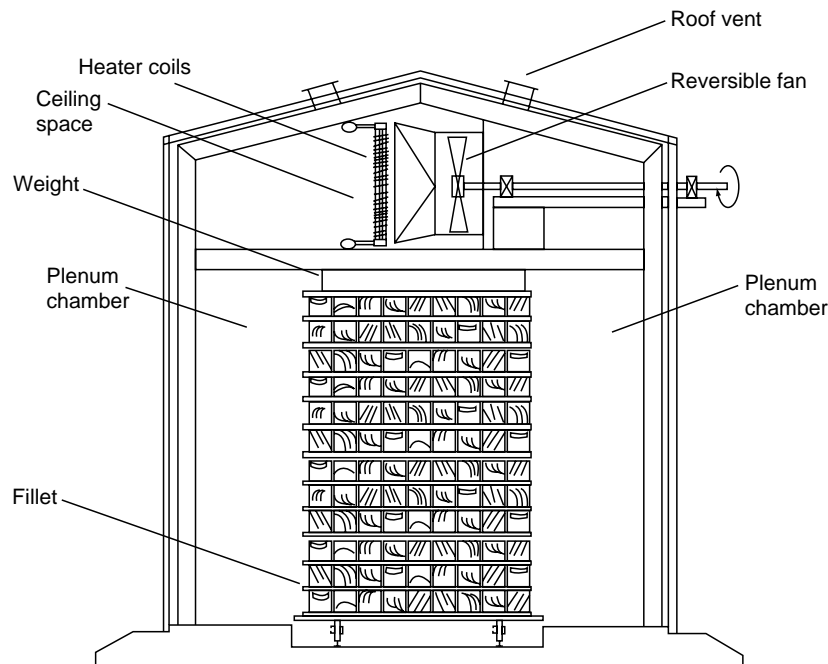


FIGURE 36.45 A vertical cross section through a single-tracked, box kiln. (Adapted from Keey, R.B., Langrish, T.A.L., and Walker J.C.F., *The Kiln-Drying of Lumber*, Springer, Berlin, 2000.)

where the boards butt up due to shrinkage on drying or unevenness in thickness and there may be gaps due to the presence of boards with uneven length. Fluid-dynamic simulation of the flow over inline slabs (Langrish et al., 1993) has suggested that gaps as small as 1 mm might be sufficient to disrupt the flow, with circulation within the gaps themselves. The magnitude of the side gaps in the board influences both the drying rates and the development of drying stresses (Langrish, 1999), so that experiments on the drying of single boards (as often done) may yield uncertain information about the drying of a load of the same wood in a kiln. With regard to variations in board thickness, Haslett (1998) recommends that the coefficient of variation for the board thickness must be under 0.04 for successful high-temperature drying.

Whenever kiln stacks are built from random-length lumber so that every second board is flush at each end of the stack, variations in openness of the stack result. This gives two different zones: (1) a central zone in which all the available space is filled and (2) two end zones where alternate boards are missing (Salin, 2001). This arrangement results in higher within-stack velocities (about 30% higher) in the center than in the end zones, with corresponding implications in the variation in drying behavior.

36.3.2.2 Moisture-Evaporation Considerations

The airflow through the stack influences the magnitude of the local airside mass-transfer coefficient, and thus the evaporation into the airstream. Particularly, at the higher air velocities used in high-temperature drying, any variations in these transfer coefficients have a significant effect on the uniformity of drying throughout the stack.

The air-inlet face of the lumber stack presents a set of blunt edges to the incident airflow, resulting in an enhancement of the mass-transfer coefficients near the leading edges (Kho et al., 1990). Computational studies (Sun, 2001) of the flow over a series of slabs with inline gaps suggest that for gaps greater than about 2 mm there will be similar, but lesser, enhancements at subsequent boards downstream.

With Scandinavian stacking practice, the end zones of the stack dry faster than the central, fully filled part (Salin and Öhman, 1998). The lower air velocity in the ends is more than compensated by higher heat-transfer coefficients associated with the flow disturbance and smaller wood volume. (There is a smaller decrease in temperature and increase in humidity along the stack in the airflow direction.) In general, it is expected that the local transfer coefficients diminish with distance in the airflow direction

due to a thickening of the boundary layer. This variation and the downwind accumulation of moisture in the airstream result in the maximum possible evaporation rate dwindling with distance from the air inlet to the air outlet from the stack.

Traditionally, the variation of evaporative rates across the stack has been counteracted by the installation of bidirectional fans and by periodically reversing the airflow direction through the stack. This policy has minimal effect on the drying rates in the center of the stack, but reduces the variation in behavior between the two end zones. If only moisture content variations are considered, many reversals are not needed to achieve this equalization (Pang et al., 1995; Nijdam and Keey, 1996; Wagner et al., 1996). However, if stress development in the surface layer with the likelihood of checking is taken into account, then the flow reversals for a timber such as *Pinus sylvestris* should be less than 2 h apart (Salin and Öhman, 1998). A period of 4 h is a common industrial practice for permeable softwoods such as *P. radiata*.

36.3.3 KILN OPERATION

To understand kiln-wide behavior, it is useful to invoke the concept of the characteristic drying curve (van Meel, 1958; Keey, 1978). The concept reduces the drying kinetics for a specific material of specific geometry to a single function of the local averaged moisture content. The concept when applied to the kiln drying of lumber boards is rough, not only due to variations in drying behavior between boards (Davis et al., 2001) but also due to embedded assumptions in the concept itself. Nevertheless, it is a sufficient representation of drying behavior to determine the effect of kiln parameters on the course of drying. These things, such as the uniformity of the airflow, the number of airflow reversals, the velocity, temperature, and humidity settings, are all those under the control of the kiln operator.

The concept of a characteristic drying curve leads to the following expression for the moisture-evaporation rate per unit of exposed board surface:

$$N_v = f\beta\phi(Y_W - Y_G) \quad (36.22)$$

where f is the evaporation rate relative to that at a given moisture content (either the initial or some critical value of transition from unhindered drying), and is a unique function of the mean free moisture content; β is the external (airside) mass-transfer coefficient; ϕ is the humidity-potential coefficient, which takes a constant value when the wet-bulb temperature remains the same throughout the kiln; Y_W is the saturation humidity at the wet-bulb temperature; and Y_G is the bulk-air

humidity between the boards. Although the values of the parameter f depend on the extent of drying that has taken place and the nature of the wood itself, the other parameters are under the control of the kiln operator by varying either the stack extent, the kiln-air velocity, or the dry- and wet-bulb temperatures.

Consideration (Keey et al., 2000) of the moisture transfer over a small zone in the kiln leads to the equations, which can be conveniently expressed in nondimensional form as follows:

$$\frac{\partial \Phi}{\partial \theta} = \frac{\partial \Pi}{\partial \zeta} = -f\Pi \quad (36.23)$$

where Φ is the moisture content relative to unit value when f is 1, Π is the humidity potential ($Y_W - Y_G$) relative to unit value at the air inlet, ζ is a nondimensional extent of the kiln in the airflow direction and is a weak function of the kiln-air velocity, and θ is the relative time of drying which itself depends upon the value of ζ for the kiln stack and the capacity of the air to pick up moisture. These equations can be solved if the parameter f is known as a function of the moisture content (averaged over the board thickness). They imply that the rate of change of moisture content with time (i.e., the drying rate) directly depends on the rate at which the bulk air humidifies in its passage through the kiln. The drying rate is also directly dependent upon the humidity potential (the driving force for the evaporation) and the parameter f , which reflects the ease of moisture movement through the wood.

These equations have been used to examine the influence of kiln variables on the course of drying, including the impact of exhaust-air recycle and the switching of airflow direction (e.g., Tetzlaff, 1967; Ashworth, 1977; Ashworth and Keey, 1979; Keey and Pang, 1994; Nijdam and Keey, 1996). A summary of this work is given by Keey et al. (2000).

Figure 36.46 shows the variation in the dimensionless drying rate, $\partial\Phi/\partial\theta$, as a function of the normalized moisture content Φ for one-way flow through a single-tracked kiln, for which the nondimensional extent ζ in the airflow direction is 1. In the case of a timber, whose initial green moisture content is equal to the critical point of transition between unhindered and hindered drying by the rate of moisture movement through the wood, the drying rate falls monotonically with both time and distance in the airflow direction. The effect with time is due to the intrinsic drying rate as the wood dries out, whereas that with distance is due to the progressive humidification in the kiln. Whenever there is free moisture content in the wood above the critical point, the drying-rate profiles are more complex. As soon as the critical

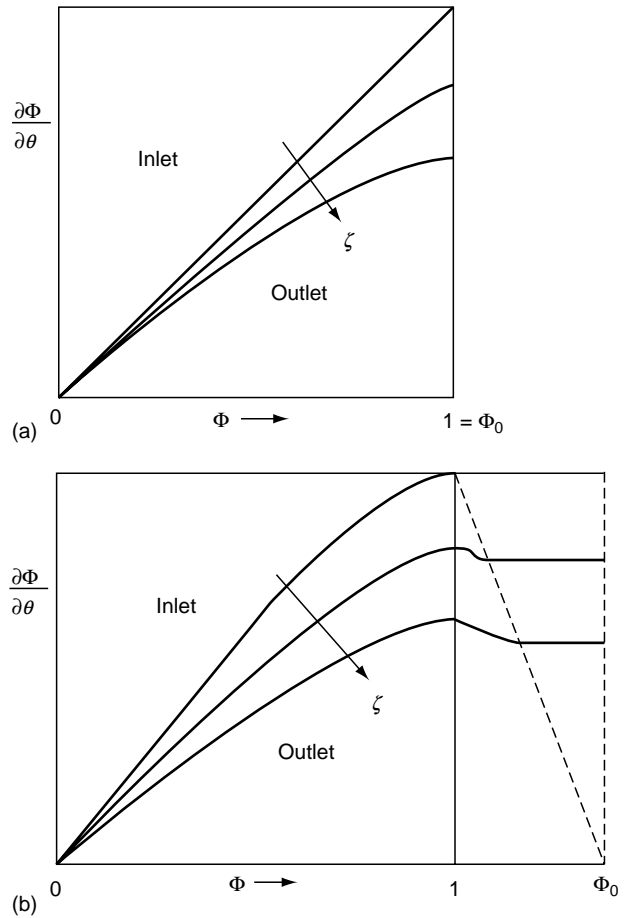


FIGURE 36.46 Normalized drying rates in a kiln with an extent ζ of 1 and one-way airflow as a function of board-averaged moisture contents: (a) an indicative hardwood with $f = \Phi$, (b) an indicative sapwood of a softwood with $f = \Phi^{0.5}$. (Adapted from Keey, R.B., Langrish, T.A.L., and Walker, J.C.F., *The Kiln-Drying of Lumber*, Springer, Berlin, 2000.)

point is reached at the air-inlet face of the stack, the intrinsic drying rate falls there, resulting in less progressive humidification in the kiln; the downstream drying rates can now rise until the local critical point is attained. The greatest effect is seen at the outlet face of the stack.

Figure 36.47 shows the effect of reversing the airflow direction through the stack. On switching over the flow, what was once the “inlet” face now becomes the “outlet” face, and vice versa, giving a temporary boost to the drying rate at the former outlet and a moderation to that at the former inlet. The rates within the center of the kiln are essentially unaffected. With flow switchovers, the leaflike moisture content profiles become more pinched with a lesser variation in moisture content across the kiln.

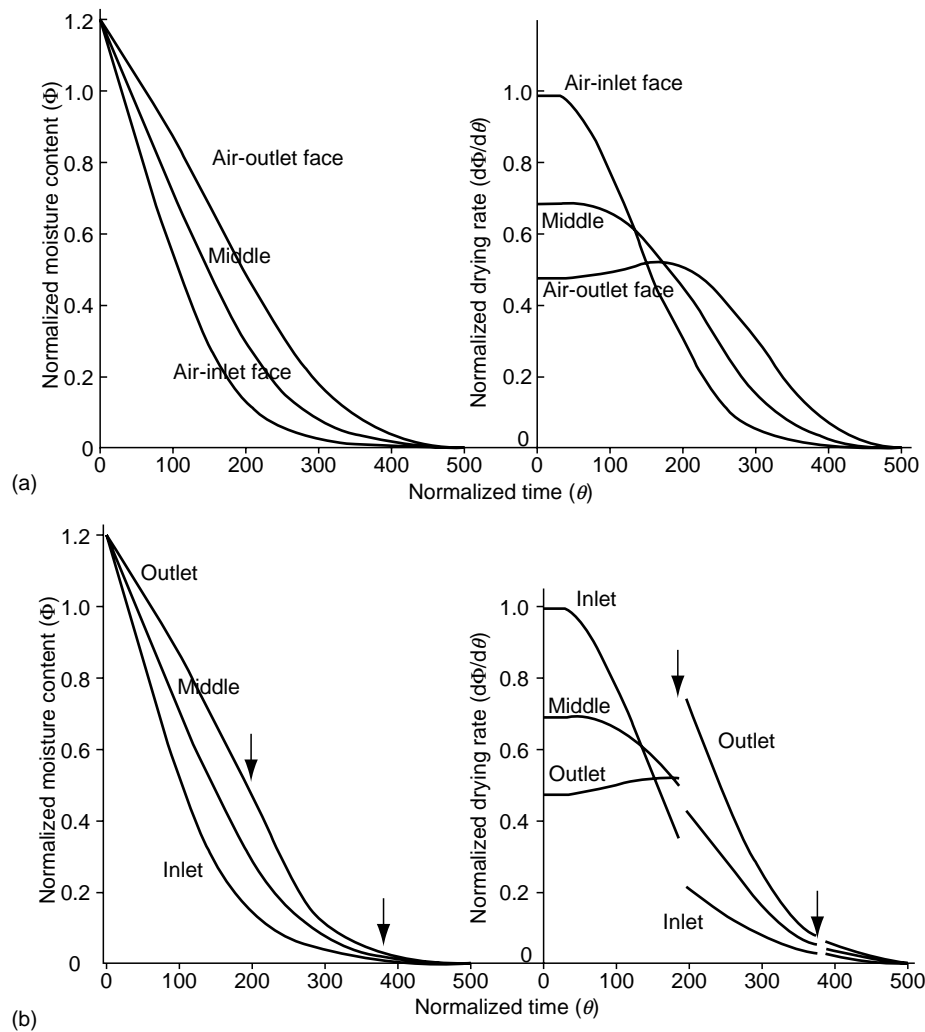


FIGURE 36.47 Board-average moisture contents and normalized drying rates as a function of time and distance in the airflow direction for a twin-stack kiln without reversals in the schedule (a) and for two flow reversals in the schedule (b). The profiles represent the drying of 100×50 -mm sapwood *Pinus radiata* dried at $77/65.5^\circ\text{C}$ and an air velocity of 2.5 m s^{-1} . (Adapted from Tetzlaff, A.R., An investigation of drying schedules when kiln-drying radiata pine, B.E. Report, University of Canterbury, New Zealand, 1967.)

Typically, lumber kilns operate at very high ratios of recycled air to that discharged through the vent to the outside air to maintain the wet-bulb temperature at the scheduled values. For that reason, commercial kilns appear to operate under very steamy conditions to the casual observer. The high degree of air recycle means that small deviations in evaporation are fed back to the air-inlet face of the stack, disturbing the conditions there. This disturbance is then propagated through the stack. In theory, with high recycle ratios and extensive dryers ($\zeta > 1$), it is possible, once the wood has reached the critical moisture content at the air-inlet face, for internal drying rates to rise above those for very wet greenwood at the air-inlet face of

the stack initially and even to exceed them (Keey, 1968). Although the necessary combination of factors to get substantial rate enhancements is unlikely in the kiln drying of most lumber species, the potential for increases in drying rate (and strain development) should be borne in mind.

36.3.4 PRACTICAL CONSIDERATIONS

Various works give a detailed overview of kiln practice. Such overviews include those by Pratt and Turner (1986), Boone et al. (1988), Hilderbrand (1989), Mackay and Oliveira (1989), Simpson (1992), and Haslett (1998).

36.3.4.1 Schedule Development

Kiln schedules are of two kinds. A tropical forest contains many species, often diffusely spread, so that individual identification of species and separate drying is rarely economic or feasible. Economic drying of mixed species involves the grouping of these according to some criteria. Simpson and Baah (1989), for example, choose basic density and initial moisture content, on the grounds that high-density, moist lumber is the most susceptible to drying defects and requires the longest drying times. Using this system, Hidayat and Simpson (1994) were about to define drying-time factors for 12 species groups. On the other hand, temperate production forests may be confined to a single kind of tree or a limited range of species, and comprehensive species-specific schedules can be developed for these. For example, Haslett (1998) specifies various schedules for a single species, *P. radiata*, according to the grade and end use of the wood.

Traditionally, schedules have been developed on a cautious, trial-and-error basis from small-scale tests on the woods of interest. Pandey (2001) notes that the seven standard, empirically derived schedules for nearly 150 Indian woods could be fitted to a diffusional model. Brandaõ and Perré (1996) propose a drying test at 90°C on small boards to provide information on drying rates and deformation criteria. This leads to an alternative species-grouping approach to that put forward by Hidayat and Simpson (1994). Increasingly, however, detailed simulation of the mass-transfer and strain-development behavior is used as a basis for determining appropriate kiln-temperature settings (e.g., Langrish et al., 1997; Aguiar and Perré, 2000a; Nijdam et al., 2000).

Kiln drying involves both the transfer of moisture through the boards and the evaporation from the exposed surfaces into the air circulating through the kiln. These processes must be kept in balance to maintain the fastest possible drying rate. With impermeable timbers, the transfer processes within the wood are rate-limiting. The wood temperature is the important variable. With permeable timbers, particularly under high-temperature conditions, external transfer mechanisms become important and high kiln-air velocities (7 m s^{-1} and higher) are used to enhance the heat- and mass-transfer coefficients.

To sustain the rate of drying as the wood dries out, either the dry-bulb temperature or the wet-bulb depression may be raised, thus providing a greater overall driving force for the drying process. However, faster drying rates can lead to steeper moisture content gradients with the risk of excessive strain development and checking. Thus most schedules specify relatively small wet-bulb depressions, corresponding

to relatively high EMC at kiln temperatures, producing modest surface shrinkage. In a conventional schedule the wet-bulb depression is typically about 5 to 10°C, whereas for high-temperature drying wet-bulb depressions can be 50°C or more.

Whenever it is important to avoid significant color development in the wood on drying, low kiln temperatures have to be used. Even though in New Zealand where accelerated conventional schedules (with dry/wet-bulb temperatures of 90/60°C) are used to dry appearance-grade timbers, commercially, kiln temperatures not greater than 50°C are employed to get very pale wood (Keey, 2003).

Although higher airflows seem justifiable at the start of a schedule when the evaporation from the exposed surface is more controlling, there is some doubt whether there is an economical advantage in designing kilns with variable-speed fans (Riley and Haslett, 1996). However, one manufacturer claims significant fan-energy savings by reducing the rate of circulating air toward the end of the schedule in the drying of a softwood without causing any significant extension of drying time (Fogarty, Priv. Comm., 2002).

36.3.4.2 Kiln Control

As the basic control of the drying process depends upon the wet- and dry-bulb readings, control of these temperatures has been the normal method of maintaining the drying schedule. Most single-zone, steam-heated kilns mount split bulbs under the overhead heating coils, one in each plenum some distance off the floor (Culpepper, 1990). In kilns divided into two zones along their length, the bulbs are placed midway in each zone. The thermometer bulbs must be protected from radiation from hot surfaces and have an adequate air circulation over them. Careful location of the bulbs is important, as significant air-temperature gradients can exist in the plenum spaces.

Simple lumber-drying installations having one or two kilns might have simple electronic-control systems with discrete programmable controls, timers, and chart recorders for temperature and steaming control. A more sophisticated system for a medium-sized facility might have a computer-based system running on windows-type software to give a visual readout of kiln conditions during the schedule (Keey et al., 2000). Endpoint determination can be based on the temperature drop across the load (TDAL), which is a function of the extent of drying (Taylor and Landoch, 1990; Martin et al., 1995).

As pointed out by Morén (2001), TDAL reflects the average drying rate in the kiln directly. He suggests that monitoring of the TDAL provides a method of “adaptive kiln control,” with the TDAL

set at a constant value at constant wet-bulb temperature after an initialization period, followed by a period at constant dry-bulb temperature.

The measurement of the boards' moisture content, at the end of drying, has frequently been done with hand-held resistivity meters, often complemented with more time-consuming gravimetric measurements with sample boards (EN 13183-1). The Standard EN 13183-2, *Round and Sawn Lumber—Procedure to Measure the Moisture Content*, employs a hammer probe with insulated measuring pins, which are inserted to a depth of approximately one third of the board's thickness, and is valid over the moisture content range from about 7 to 30%. This technique has been adapted by inserting probes in the kiln stacks to give a continuous reading of the moisture content in the later stages of drying. Nonpenetrative systems are available in which the resistivity probes are aluminum fillets placed along the length of the lumber charge to measure the resistance of all the boards that the probes are in contact with. All resistive methods, however, are limited to measuring moisture contents from just above fiber saturation and below, with one system having a claimed repeatability of $\pm 2\%$ when the average moisture content is less than 16% (Furniss, Priv. Comm., 2002).

The alternative use of microwave-sensing technologies is attractive in enabling a wider range of moisture contents to be determined from green to dry (Holmes and Riley, 1996; Riley and Holmes, 2001). The first prototype tested used a waveguide, $40 \times 20 \times 1.2$ mm, with angled slots in the long face, connected by a high-temperature coaxial cable to the oscillator and placed in a stack so that the slots were in contact with the lower face of a board. A later development enabled the possibility of the stack's moisture distribution to be determined. The experiments have demonstrated the potential for the manufacture of a rugged industrial system. However, unless the timber is very uniform in properties, this technique is uncertain as the relationship between the microwave signal and the moisture content becomes fuzzy because of factors such as variations in wood density among the boards.

An online contact-free method of measuring the moisture content of wood, which involves traversing the boards on a band between sensor heads, could be developed based on techniques used in papermaking. The method would be suitable for sorting boards after drying. At least one such method is commercially available, but the measurement principle has not been divulged (Smith, Priv. Comm., 2002).

The benefits of even a relatively simple kiln-control system are illustrated in an example quoted by Culpepper (1990). The incorporation of a

programmable logic controller at one site increased the overall grade recovery from 70.7 to 81.9%, with a reduction in energy costs of 43% for steam and 10% for electrical power. The total saving represented a 1.2-y payback on the investment.

36.3.4.3 Volatile Emissions

Acidic corrosion is a well-recognized feature of kiln drying, particularly with certain species such as oaks. Packman (1960) notes that, of some 150 species studied of both hardwoods and softwoods, the majority (80%) has a pH between 4 and 6 and only one had a pH consistently above 7. In an extreme case, with a hardwood of high extractives content, sufficiently severe rusting of steel trolley wheels and other ferrous fittings in a pilot-plant kiln required their replacement after a 2-week period. Extensive use of aluminum linings with plastic piping in modern kilns has largely avoided the problems of corrosion.

The corrosive nature of the volatile substances released during kiln drying is derived from the presence of free acetic acid in the wood and from the hydrolysis of the various acetyl groups attached to the hemicelluloses. Softwoods generally yield greater emissions than hardwoods, because the latter are normally dried at lower temperatures and have lower resin content. Some volatile substances, such as formaldehyde, come from the thermal degradation of the hemicelluloses and lignin, whereas green pines release terpenes and their derivatives. Table 36.11 lists the concentrations of volatile emissions from two high-temperature schedules.

Milota (2001) compared the emissions from various North American lumber species in a small-scale kiln

TABLE 36.11
Concentrations of Volatile Emissions Arising from Two High-Temperature Kiln Schedules for *Pinus radiata*

Compound	Concentration, g m^{-3} at dry/ wet-bulb temperatures ($^{\circ}\text{C}$)	
	120/70	140/90
Formaldehyde	19.5	31.0
Acetic acid	21.7	38.2
Monoterpenes	34.8	66.4
Hydroxylated monoterpenes	12.6	10.7
Condenser residues (resins and fatty acids)	16.4	16.4

Source: McDonald, K.A. and Wastney, S., Analysis of volatile emission from kiln drying of radiata pine, *Proceedings of the Eighth International Symposium on Wood Pulp Chemistry*, Vol. 3, 431–436, 1995.

with those from the same wood in a commercial kiln. The laboratory kiln functioned like the mill kiln with respect to venting characteristics, temperature, and humidity, but it dried the wood faster, possibly because the load is narrower. In general, the small-scale method was fairly repeatable for the determination of volatile organic chemicals, with a standard deviation of 8 to 16% of the mean value, but the repeatability was poorer for methanol and formaldehyde.

36.3.4.4 Equalization and Stress Relief

At the end of kiln drying there is always some board-to-board variations in moisture content, arising from the inherent differences in drying characteristics of individual boards and the progressive humidification of the circulating air. Because kiln drying is basically a process in which the boards are forced to reach a new equilibrium, moisture content profiles also exist within each piece of wood. With moisture-based schedules, the kiln conditions are reset at the end of the schedule to give an EMC that is 2 to 3% below the desired value. This strategy prevents overdrying of the drier boards while allowing the wetter boards to dry further toward the target end moisture content. Some authorities (e.g., Haslett, 1998) recommend that the softwood lumber should be slightly overdried to ensure that the moisture content variation both within and between boards becomes small. Any subsequent steaming, which is undertaken for stress relief, will raise the moisture content of the overdried load toward the specified value.

Kiln schedules are designed to dry the lumber as fast as possible without causing unacceptable defects to appear due to excessive strain development. Residual stresses in the wood must be relieved if the lumber is to be further processed. Steaming is a common method of doing this.

In smaller installations, steaming may be done in the kiln; but, in larger installations, a separate chamber supplied with low-pressure, saturated steam is used for this purpose. Such chambers are not normally supplied with fans, and stratification of the steam and air can be a problem. Lumber that is high-temperature dried should be cooled so that the wood temperature falls to about 70 to 95°C to enable the wood to pick up moisture. The steaming induces a reversal of the moisture content profile through the wood, with concomitant reduction in the residual stresses.

Chen et al. (1997b) examined various stress-relief strategies for sapwood boards of softwoods and have found that other procedures are suitable in addition to final cooling and steaming. These include simple cooling under cover or the use of a schedule consisting of intermittent drying and conditioning cycles.

36.3.5 LESS-COMMON DRYING METHODS

The majority of lumber-drying installations are convective drying chambers worked at atmospheric conditions and temperatures not greatly above ambient. For structural-grade, permeable softwoods, however, when color development is not a prime concern, elevated temperatures can be used to get very fast drying processes. Designs under consideration include kilns being worked to temperatures of 200°C, with air velocities between the boards up to 15 m s⁻¹. Continuously worked kilns then become attractive with very fast drying. Such designs will require particular attention to thermal expansion, reliable heat-exchange equipment, and venting of moisture vapor.

Many of the advantages of high-temperature drying can be obtained by working under vacuum. In particular, such a process gives rise to an internal overpressure and an additional and efficient driving force for internal moisture migration. The lower operating temperatures are an advantage in drying heat-sensitive woods and in minimizing color development when pale products are required.

Similar advantages of lower working temperatures are obtained in the use of dehumidifying heat-pump systems, with the added bonus of lower thermal-energy use to compensate for extended drying times. The drying principle of these kilns, however, is the same as vented conventional kilns.

Most kilns, either direct-fired or steam-heated, use wastewood as the primary fuel. Other heating arrangements have been advocated such as the use of microwave, radio frequency, and solar energy, with the latter being attractive in remote locations for small kilns.

36.3.5.1 Vacuum Drying

Vacuum dryers have been commercially available for many years and their use is regarded as a standard practice in Europe for the drying of high-quality hardwoods economically, which would otherwise be difficult to dry (Hilderbrand, 1989). Descriptions of vacuum drying are given by Ressel (1994), Audebert and Temmar (1997), and Jomaa and Baixeras (1997).

Because of the enhanced internal moisture migration under vacuum, the rate of drying can be as rapid as that at a much higher temperature at atmospheric pressure. However, the higher specific volume of vapor associated with the reduced pressure is a severe limitation for heat transport by convection (Perré et al., 1995).

Several industrial solutions have been proposed:

- The use of plates heated by electrical resistance or circulation of heated water are placed between each layer of boards; heat is supplied to

boards by conduction whereas the vacuum level is used to enhance the internal mass transfer.

- Discontinuously operating kilns, with two periods of about 1 h alternate: a heating period at atmospheric pressure and a drying period at reduced pressure.
- Finally, the most recent “high-vac” kilns use a slightly higher pressure level (more than 100 mbar), together a very high linear air velocity (10 m s^{-1} or more), to compensate for the loss of thermal capacity of the air; this method has proved to be very effective.

In the latter method, uniformity of the air distribution through the load is important to ensure evenness of drying, with regions of low velocities resulting in higher final moisture contents (Ledig and Militzer, 1999). The positions of fans and heating coils have an important bearing on the temperature and on the final moisture content of the load (Hedlund, 1996). Vacuum dryers with overhead fans provide a fairly controlled airflow path through the load, but other fan locations can result in ill-defined pressure and suction sides. An overhead-fan dryer, however, was found to yield a systematic variation in temperature between the door and the other end of the dryer, which might have been reduced by dividing the unit into separate temperature-control zones. Techniques to overcome the inherent poor heat transfer in vacuum dryers include the use of heated plates between the boards or intermittent heating with superheated steam. Another suggested technique employs a heating cycle at atmospheric pressure, when the heat transfer is better, followed by a vacuum-drying cycle (Guilmain et al., 1996). Tests on drying oakwood at pilot and industrial scales showed that the discontinuous process was faster, with less susceptibility to mechanical damage of the wood, but the thermal consumption was higher than under continuous vacuum conditions.

Behnke and Militzer (1996) have produced a vacuum-dryer model for design and process-control purposes based on a characteristic drying curve for the wood's drying behavior. Hilderbrand (1989) claims that commercial drying time in vacuum kilns varies between one half and one third of those found in conventional convective kilns under atmospheric pressure.

36.3.5.2 Dehumidifier Kilns

Drying at low temperatures, which is a feature of seasoning refractory timbers, is energy-inefficient. A dehumidifier kiln reduces the thermal-energy consumption by incorporating an air-conditioning unit that recovers heat by cooling the kiln air below its

dew point and, in effect, recycling the latent heat of condensation. As moisture is removed as condensed liquid rather than vapor in warm discharged air, the associated thermal loss is avoided. However, a small amount of venting is needed for humidity-control purposes. Volatile organic chemicals normally removed with the vented moist air now appear in the condensate stream, which potentially could be sent to a separate unit for chemical recovery.

Figure 36.48 shows the layout of a heat-pump dehumidifying kiln. Moist air is drawn over the evaporator and condenser consecutively in a Rankine cycle heat pump. Besides these basic elements such as an evaporator, a condenser with its associated compressor, and expansion valve, there is an accumulator that prevents the refrigerant from entering and damaging the compressor and a subcooling heat exchanger to enhance the effectiveness of the heat pump.

The performance of dehumidifier kilns is normally expressed in terms of the specific moisture-extraction rate (SMER), which is the amount of moisture extracted per unit energy input. Two such ratios may be defined: one representing how effectively the dehumidifier extracts moisture from the air as condensate and the other (the kiln SMER) representing how efficiently the kiln removes moisture from the lumber including the condensate and venting. The kiln SMER for convective kilns is limited to about 0.8 to 0.9 kg kWh^{-1} , compared with values in the range of 1.5 to 2.5 kg kWh^{-1} for commercial dehumidifier kilns (Davis, 2001). Some of the lower values may reflect the poor insulation of the tested kilns rather than a defect in the process.

Dehumidifying kilns are limited in operating temperature by the working limits of the compressor ($<70^\circ\text{C}$) and the size of the unit ($<40 \text{ HP}$). However, dehumidifying heat-pump units can be coupled together and a mill with as many as 14 units is operating (Kerr, Priv. Comm., 2002).

Dehumidifier drying has attractions when high-quality solid wood is sought. Because of the relatively low operating temperatures, color and kiln brown stain development is reduced significantly while checking is less than in standard alternative schedules.

36.3.5.3 High-Frequency Electrical Heating

In wood drying, heating with high-frequency electric fields in both the dielectric and microwave range has been considered and some specialized applications have been found. Schiffmann (1995) defines the dielectric frequencies of electromagnetic radiation as those covering the range from 1 to 100 MHz , whereas microwave frequencies are considered to span from 300 MHz to 300 GHz . He suggests that, as a rule of

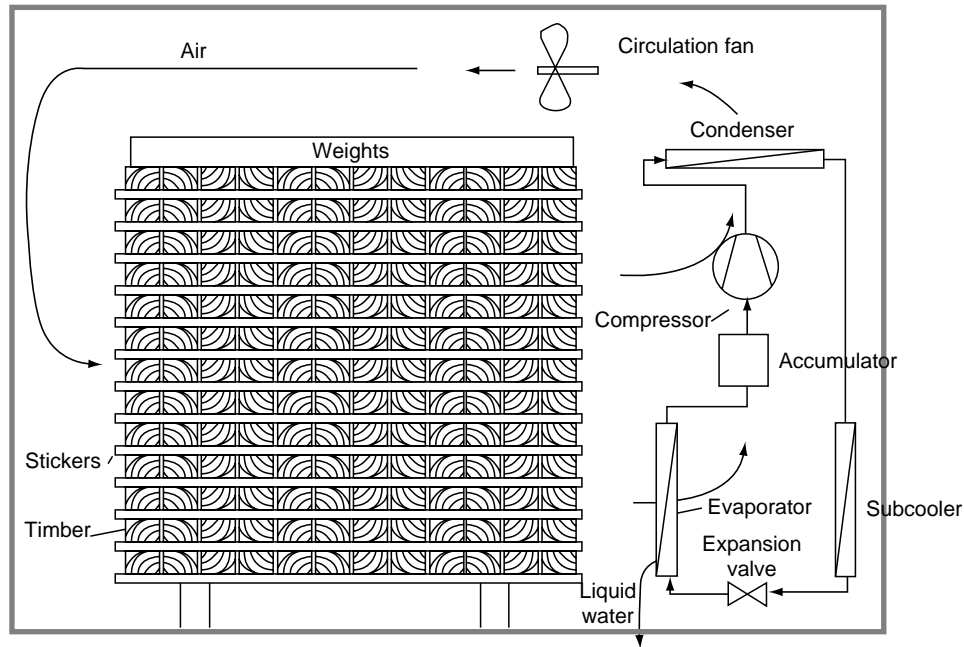


FIGURE 36.48 A typical configuration for a heat-pump dehumidifying kiln. (Adapted from Davis, C.P., Drying *Pinus radiata* boards in dehumidifier conditions, Ph.D. thesis, Otago University, New Zealand, 2001.)

thumb, these high-frequency heating methods become economically attractive for new kilns if the drying rate is increased fourfold over that for conventional drying. In general, the use of dielectric and microwave heating may become attractive for the small-scale drying of high-value hardwood species that are difficult to dry by conventional means. For example, Smith and Smith (1994) report the use of radio-frequency heating for the drying of oakwood in a small vacuum kiln of 23-m³ capacity, which had a lower capital cost but higher energy costs than a conventional dryer for the same duty. For very small power requirements, microwave heating is more attractive; when the power requirement exceeds 50 kW, however, economics favor the higher-power tubes in the radio-frequency range. In one Canadian system, radio-frequency drying is used to finish the seasoning of conventionally dried lumber that has not met target moisture content.

Heating is generated in the dipolar rotation of water molecules as they try to orient themselves in the rapidly changing polarity of the applied electrical field. The power developed per unit volume is given by

$$P = kE^2f\epsilon' \tan \delta \quad (36.24)$$

where k is the dielectric constant, E is the electric field strength, f is the field's frequency, ϵ' is the relative permeability, and $\tan \delta$ is the loss tangent or

dissipation factor. The field's strength and its frequency are fixed by the equipment, whereas the other parameters are material-dependent. As the dielectric constant of water is over an order of magnitude greater than the woody materials, moisture is preferentially heated, a process that leads to a more uniformly moist product with time. This feature is one of the attractions of the technique, for example, in moisture leveling in the manufacture of plywood to avoid delamination during subsequent hot pressing (Schiffmann, 1995).

There is also a contribution due to ionic conduction because of the presence of ions in the sap. This mode of heating is not significantly dependent on either the temperature or the frequency of the applied field, but is directly dependent on the charge density and mobility of the ions.

Because the heating is internally generated, rather than convectively warmed at the exposed surface of the boards, high and damaging internal pressures can be created in the process. For example, internal overpressures of 60 kPa have been reported by Antti (1992) for power inputs of 1.25 kW on drying 100 × 50 × 1660-mm boards. Under vacuum drying, such overpressures become less damaging. Thus, high-frequency heating has been advocated for use with vacuum drying because of the difficulty in achieving adequate convective heating under vacuum, and a summary of its historic development is given by Resch and Gautsch (2001). This

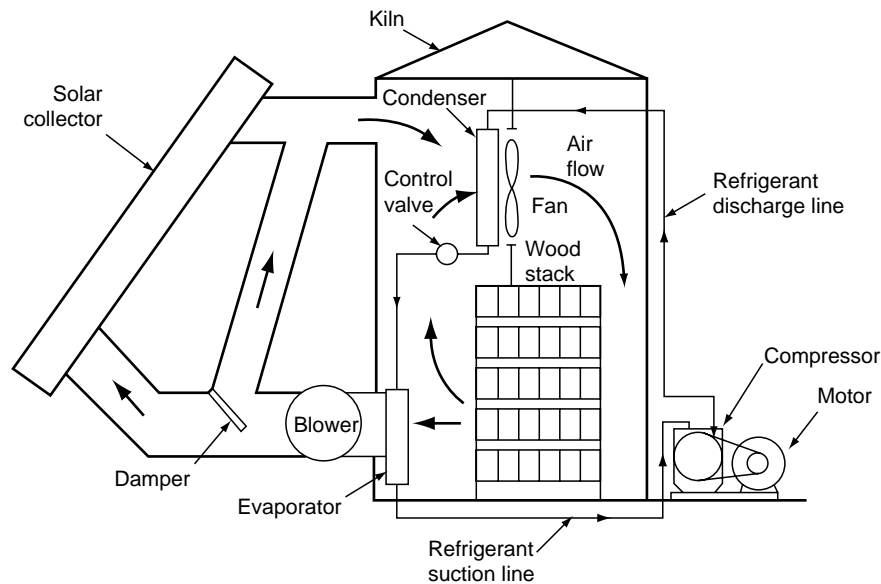


FIGURE 36.49 A solar-dehumidifier dryer. (Adapted from Chen, P.Y.S., Helmer, W.A., and Rosen, H.N., Experimental solar-dehumidifier kiln for drying lumber, *Forest Prod. J.*, 32(9): 35–41, 1982.)

technique is attractive for beech and oak timbers in the European market because of the retention of their natural light color with low-temperature drying.

Perré and Turner (1997, 1999a) have described a numerical model of microwave drying of softwood with an oversized waveguide. In this work, internal overpressure reaching two to three times the atmospheric pressure has been reported both in experimental and numerical results.

36.3.5.4 Solar Drying

Solar drying of lumber has attractions in remote locations with favorable climates because of the “free” nature of the energy source. Imré (1995) has classified solar-heated dryers into three main groups:

1. Solar natural dryers that use only the sun
2. Semiartificial solar dryers with a fan to supply a continuous flow of air through the load
3. Solar-assisted artificial dryers, which may use an auxiliary energy source for boosting the heating rate

Mixed types include a solar-dehumidifier dryer with forced-air recirculation, as shown in Figure 36.49.

Many solar dryers described in the literature are simple greenhouse kilns (e.g., Langrish and Keey, 1992). In these units, the solar collector is fitted within the structure that holds the load and the airflow is maintained by fans. The solar energy, in other cases,

is collected externally in heat-storage systems or panels, as illustrated in Figure 36.49. Greenhouse kilns have attractions in simplicity of construction and operation.

The daily world-average solar radiation on a horizontal surface is 3.82 kWh m^{-2} (McDaniels, 1984), with values in tropical countries being higher (up to 7.15 kWh m^{-2}) (Imré, 1995). However, Plumptre (1989), reported by Keey et al. (2000) on reviewing 35 solar kiln designs, notes that the location of these was spread almost uniformly over the range in latitude from 0 to 50° .

Langrish and Keey (1992) observed one operational feature of the use of a greenhouse kiln. With the kiln’s vents shut overnight and with the drop in ambient temperature, the relative humidity in the kiln would rise sufficiently for moisture to condense on the wood’s surface. This provided a degree of conditioning, which prevented the development of excessive checking in a refractory hardwood being dried.

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37 Drying in Mineral Processing

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37.1 INTRODUCTION

Thermal removal of water from solids or slurries is an important operation carried out in numerous mineral-processing and metallurgical-processing applications. Although drying is a highly energy-intensive operation that is also increasingly difficult at lower moisture contents, no special attention is generally given to the technical and economical aspects of the drying process employed in the mineral- or metallurgical-processing industry. It is therefore not surprising that most dryers found in these industries are of the conventional type, as discussed later in this chapter. Detailed descriptions and design considerations of specific dryer types (e.g., rotary, fluid bed, and spray) are presented elsewhere in this handbook. The interested reader is referred to relevant sections for further information. The objective of this chapter is to summarize the types of dryers currently used in practice, to discuss any special aspects with illustrations, and to identify possible new concepts that may be applicable in the mineral industry.

In the mineral and metallurgical industry, drying is generally carried out at the raw-material or product-handling stage; for example, after beneficiation or concentration the ore may require drying to some optimum level to facilitate handling. Among the various possible reasons for resorting to drying, some are listed below:

1. Saving freight charges if the ore is transported over large distances.
2. Facilitating handling in conveyors, cars, bins, and so on.
3. Improvement of efficiency of subsequent drying processes, such as screening, air classification, and electrostatic precipitation.
4. Enhancing efficiency or cutting fuel consumption in roasting and calcination, for example.
5. Minimizing or eliminating problems of handling wet solids due to freezing in cold climates.
6. Some processes, such as flash smelting, may require bone-dry solids as a necessity for efficient operation.

As a general rule, one must establish the necessity of drying before undertaking further action. If it is deemed necessary, it is essential to maximize mechanical dewatering (by filtration, centrifugation, pressing, and other processes) to reduce the thermal load in drying. It may be noted that below a typical solid moisture content of 4%, dusting may pose serious handling and environmental problems. Fire and explosion hazards as well as hazards associated with the handling of toxic materials are commonly encountered in the drying of solids in the mineral and metallurgical industry. Needless to say, as in the latter industry the final product is typically molten metal, drying is not required at the product-handling stage.

Sometimes drying may be achieved simultaneously with another operation. For example, a rotary kiln-pulverized coal firing system requires coal that is dried to an optimum moisture level, which depends upon the rank of the coal as well as the type of grinding system used. Between 2 and 4% moisture in the coal leaving the mill is best for ball mills. Lower-rank coals with higher hygroscopic moisture can generally be ground at higher moisture levels without seriously affecting their handling and grinding characteristics. The upper level of moisture in the coal is determined by the stability of the flame; for a rotary kiln, about 15% represents the upper limit [1].

Conventional dryers used in the mineral-processing industry are classified as hearth type, shaft type, and grate type. Other types of dryers used less commonly in current practice are the spray type, fluid-bed type, pneumatic or flash type, conveyor type, drum type, stationary- and rotating-tray type, infrared type, and others. Only the more commonly used dryers are discussed in this chapter.

After a brief overview of the essential characteristics of the various major types of dryers found in the mineral industry, we discuss the rotary dryers in some depth, followed by a summary of industrial applications of other, less frequently used dryers that have the potential for increased application in the near future.

37.2 CONVENTIONAL DRYERS

37.2.1 HEARTH TYPE

For the hearth type of dryers, the bed of solids to be dried is supported on a floor or hearth that is heated directly. Heat for drying is supplied primarily by conduction. These dryers are in use for the drying of flotation concentrates of zinc and lead, copper sludge, washed kaolin, and so on. These dryers are inherently slow and highly labor-intensive. These are popularly employed for dusty solids, such as bauxite. Following

are the ranges of thermal-performance characteristics found in practice:

1. Evaporative capacity: 5 to 25 kg water evaporated/h/m² (hearth area)
2. Fuel consumption: 1500 to 4500 kcal/kg water evaporated
3. Thermal efficiency: 10 to 30%

Because of their low thermal efficiency and large space requirements, hearth dryers are less popular now.

37.2.2 GRATE TYPE

In the grate type of dryers, the wet granular solids are supported on and conveyed by a reciprocating grate through which drying air is passed. Such dryers are extensively used in the drying of “green balls” in iron, copper, and chromium manufacture. These dryers are sometimes operated with alternating updraft and downdraft to prevent possible condensation of water on the downstream layer of balls. Drying-air temperature varies from 300 to 450°C. Although thermal efficiency improves with temperature, higher temperatures may cause thermally induced cracking of the pellets, resulting in poor mechanical strength, excessive dusting, and severe fan wear. The typical bed depth ranges from 2 to 5 m; the air velocity (superficial) ranges from 50 to 125 m/min. Its evaporative capacity is at least three times that of hearth-type dryers, i.e., 25 to 75 kg water evaporated/h/m² of grate area. The thermal efficiency is also much superior from 30 to 60%. Gentle handling (of pelletized solids), better quality control of the product, and relatively good thermal performance are some of the advantages of grate dryers. They are also used for cooling solids after drying. Among their disadvantages are high maintenance costs due to grate wear and high material- and air-handling costs. They cannot be used for sticky or pasty feeds.

37.2.3 SHAFT DRYERS

Shaft dryers are convective dryers in which the granular material is “showered” through a current of the drying medium, which may flow in the vertical or horizontal direction. The former are often termed tower dryers and the latter constitute, perhaps, the most widely used “rotary” dryers.

The tower dryer is basically a chimneylike structure made up of bricks, insulated steel, or concrete. Hot gases from a furnace are led through the bottom whereas the wet solids flow downward countercurrently to the drying medium. The flow of the solid

material is retarded (to allow necessary residence time in the dryer) by means of properly designed baffles. Such dryers are in use in the zinc and iron industries. Typical sizes are in the range of $2 \times 2 \times 20$ m. The air velocity ranges from 60 to 150 m/min; the thermal efficiency is about 35 to 60%. The design is simple and the operation is trouble-free for nonsticky materials. For free-flowing solids with little residual moisture, these types of dryers offer operational advantages. Such applications are exceptions rather than the rule in the industry.

As the rotary dryer is by far the most commonly encountered dryer in the mineral industry, it is discussed in some detail in this chapter. Additional information on rotary dryers is available elsewhere in this handbook. Essentially, rotary dryers are cylindrical drums normally supported on two or more tires running on a set of rollers at a slight incline to help convey the material. Wet feed enters the dryer through a feed chute, belt conveyor, screw conveyor, vibrating conveyor, or other suitable means. Drying air flows through the shell in either a cocurrent or countercurrent fashion. Lifters in the drum shower the material over the cross section of the shell to induce intimate contact with the drying air. Rotary dryers are very versatile and economic for large-tonnage drying of relatively low-value materials. In the mineral industry, they range from 1.5 to 4 m in diameter and are 6- to 30-m long. Evaporative capacities are of the order of 30 to 120 kg/h/m³ (dryer volume); thermal efficiencies range from 35 to 70%. Low capital costs, close quality control, low maintenance costs, and trouble-free operation over extended periods are some of the advantages of rotary dryers. Among the disadvantages are the difficulty of sealing, high structural load, and nonuniform residence times. For fine materials that tend to cause a dusting problem when dry, it is difficult to operate rotary dryers at high efficiency without more complicated mechanical modifications, which are mentioned later in this chapter.

37.3 VARIOUS DRYERS USED IN MINERAL PROCESSING

37.3.1 ROTARY DRYERS

It is customary to classify rotary dryers as direct heat and indirect heat. Hot gases are brought into direct contact with the wet solids in the former type, and the heat for evaporation is supplied through metal ducts or tubes located within the shell in the latter variety. Except for special operations, the rotary dryer operates at near atmospheric pressure. Vacuum operation is recommended for special applications, such as

drying of yellow cake in the *in situ* uranium leaching process. In the indirect type, heat is transferred by conduction and radiation (with a minor contribution by convection); in the direct type, the predominant mode is convection, though radiation may also be significant depending upon the operating temperature.

It is impossible to use one type of dryer for all applications. Pilot-plant tests and prior experience are the best guides in the selection of the proper type as well as the optimal operating conditions for the specific problem at hand. All elements of drying costs, namely, drying system (including pretreatment of feed and posttreatment of dry product), fuel, power, and maintenance costs, must be examined carefully before selecting the drying equipment. Efficient use of heat generally requires the control of entering- and exit-gas temperature and the control of the drying time to minimize heat losses. In the mineral industry, it is not common to use heat-recovery equipment even with modern installations. Maintenance costs are related to the first costs. Inadequate attention to mechanical and structural considerations, which may lower initial costs, may in the long run lead to higher maintenance costs.

For directly heated counterflow or parallel-flow rotary dryers, the typical heat requirement for evaporation ranges from 1200 to 2200 Btu/lb. For efficient operation, it is desirable to use the highest possible air temperature; in the mineral industry the normal range is 1400 to 1800°F. For heat-sensitive materials, lower temperatures may be needed. For best heat transfer, it is necessary to use the highest possible air velocity; for dusty materials this poses a problem. For product moistures below 1%, a counterflow dryer is recommended if the product is not heat-sensitive. For thermolabile materials with product moisture in excess of 1%, a parallel-flow dryer is desirable as the feed is in contact with hot gases while drying the surface moisture. The solid temperature does not exceed the corresponding wet-bulb temperature (corrected for radiation effects). For sticky or caking solids, which do not shower well, it is often necessary to allow more residence time in the dryer. New designs include special internals within the shell to break up the formation of large agglomerates in the drum so that a rotary dryer could be used to dry sticky or pastelike feeds.

Dryer loading may vary from 3 to 15%, of which 8 to 12% is the most common. This represents the volume occupied by the material within the dryer as a percentage of the total volume of the shell. Rotary dryers have a retention time of 5 to 25 min, with the normal range of 7 to 15 min. Dryer speed and inclination can be adjusted to attain the desired retention time for a given material. Kramm has provided a

table of some field installations to illustrate the size, operating parameters, and types of materials handled in direct-heat rotary dryers [1].

37.3.2 SPRAY DRYING

Spray-drying technology has found a number of important applications in the mining industry, particularly in drying a wide assortment of mineral flotation concentrates. Spray drying can substitute for conventional dewatering methods followed by drying in rotary dryers or kilns. By means of spray drying, flotation concentrates (pulp, sludge, and precipitate) can be dried to a bone-dry state (less than 0.5% moisture) in a single continuous operation. The following is a list of advantages of spray drying in the mining industry:

1. Feed can be taken from thickener underflow directly; this is important when handling colloidal suspensions, which are difficult to filter.
2. Spray drying yields bone-dry product that is essential for flash smelting. It also cuts down fuel use in smelting or refining operations.
3. Product handling is easy. Product loss and environmental pollution can be minimized.
4. Pyrophoric materials, which tend to undergo spontaneous combustion at moisture levels of 3 to 5%, can be dried to a bone-dry level at which they are inert.
5. Plant operation and control are easy. Thermal efficiencies of spray-drying systems are inherently better than those of most other dryers. Extra equipment to recover low-grade heat in exhaust is not necessary.

The solids content in the slurry is typically between 55 and 70%. A pumpable slurry is needed to accomplish atomization. A specially designed rotary atomizer converts the slurry into a cloud of droplets that contact the hot drying gases, which may be at 500 to 1000°C, thus giving excellent heat economy. The exit-gas temperature is typically between 110 and 150°C. The dried products attain a temperature well below that of the exit air. It can be noted that at inlet temperatures above 750°C, the inlet duct to the dryer is refractory-lined. If available, exhaust gases can be used as a supplementary heat source, thus reducing the overall heat consumption. Once the pulp is bone-dry, it is discharged through the bottom of the conical drying chamber into a conveying system. Fines entrained in the exhaust gases are collected in high-efficiency electrostatic precipitators and then discharged into the conveyor system. Product recovery can also be achieved by high-efficiency cyclones combined with a wet scrubbing system.

The heart of the spray-drying system is the atomizer. As mineral slurries tend to be abrasive, conventional atomizers cannot be used for this application. New designs that are suitable for abrasive slurries have appeared in the market during the last decade. In one commercial design, the atomizer wheel is equipped with the wear-resistant, sintered, bushings so arranged that a protective layer of the concentrate builds up in the inner wheel chamber as a result of the fluid mechanics of the flow generated. Thus the concentrate provides its own protective cover for abrasion control. The bushings can be turned through 90° to quadruple their operating life.

Commercial installations based on spray-drying technology are currently in use in the mining industries around the world. Examples of products that are spray dried include chromium concentrate, copper concentrate, lead-manganese-molybdenum-tin-zinc concentrate, yellow cake, precious metal slime, zinc leach residue, and rock phosphate.

The cost of spray drying depends upon the prices of fuel, power, and labor, which vary with location. Typically, the energy consumed per metric ton of bone-dry concentrate—starting from 70% solids in pulp—is about 325,000 kcal (or equivalent to about 9 U.S. gal of fuel oil). About 10 kWh of electrical energy is needed for air handling and atomization. As the atomizer is the only rotating part, maintenance costs are small, aside from replacement of the expendable inserts in the atomizer wheel. Abrasion may occur in the drying chamber or air ducts where air velocities are high.

For more details about the spray-drying process, the reader is referred to the appropriate sections of this handbook and to the literature cited there.

37.3.3 VACUUM DRYERS

Development of new processes often leads to new requirements on the drying equipment needed. One example of this is the yellow cake-precipitation process of International Energy Corporation (IEC) [2], which uses neither chlorides nor sulfates in extraction and precipitation for *in situ* uranium leaching. This process avoids the formation of uranium salts that require high-temperature calcination. IEC uses a solution of ammonium carbonate and hydrogen peroxide for leaching and live steam to sparge the eluate to precipitate the so-called yellow cake. Yellow cake slurry with 30% solids is fed to a rotary vacuum dryer to produce a dry product ready for packaging. Compared with the conventional open-hearth dryer, vacuum drying offers the advantages of low capital costs, low installation and maintenance costs, and environmentally safe operation with no dust problems.

Figure 37.1 shows the vacuum drying system schematically. The dryer is 3 ft in diameter with an active length of 10 ft to dry 1200 lb/d of yellow cake. The approximate utility requirements are 250 lb/h of steam at 50 psig, 40 gpm of condenser cooling water, and electricity to drive a 10-hp motor. Water vapor from the dryer is condensed in vertical tube-type condensers and recycled to allow nearly total water recovery; this system thus conforms closely to the zero-discharge concept advocated by regulatory bodies. Because the dryer operates under vacuum, a leak results in air being drawn into rather than out of the system.

The batch-type rotary vacuum dryer has an internally heated helical agitator that provides excellent heat transfer. For sticky or adhesive materials, addition of a spring-loaded scraper-blade device prevents the product buildup during the drying cycle.

37.3.4 FLUIDIZED-BED DRYERS

Batch or continuous drying in an upward flow of hot gas, with or without additional heat transfer provided by immersed heat exchangers in the bed, provides an efficient method of drying for fluidizable, nonsticky

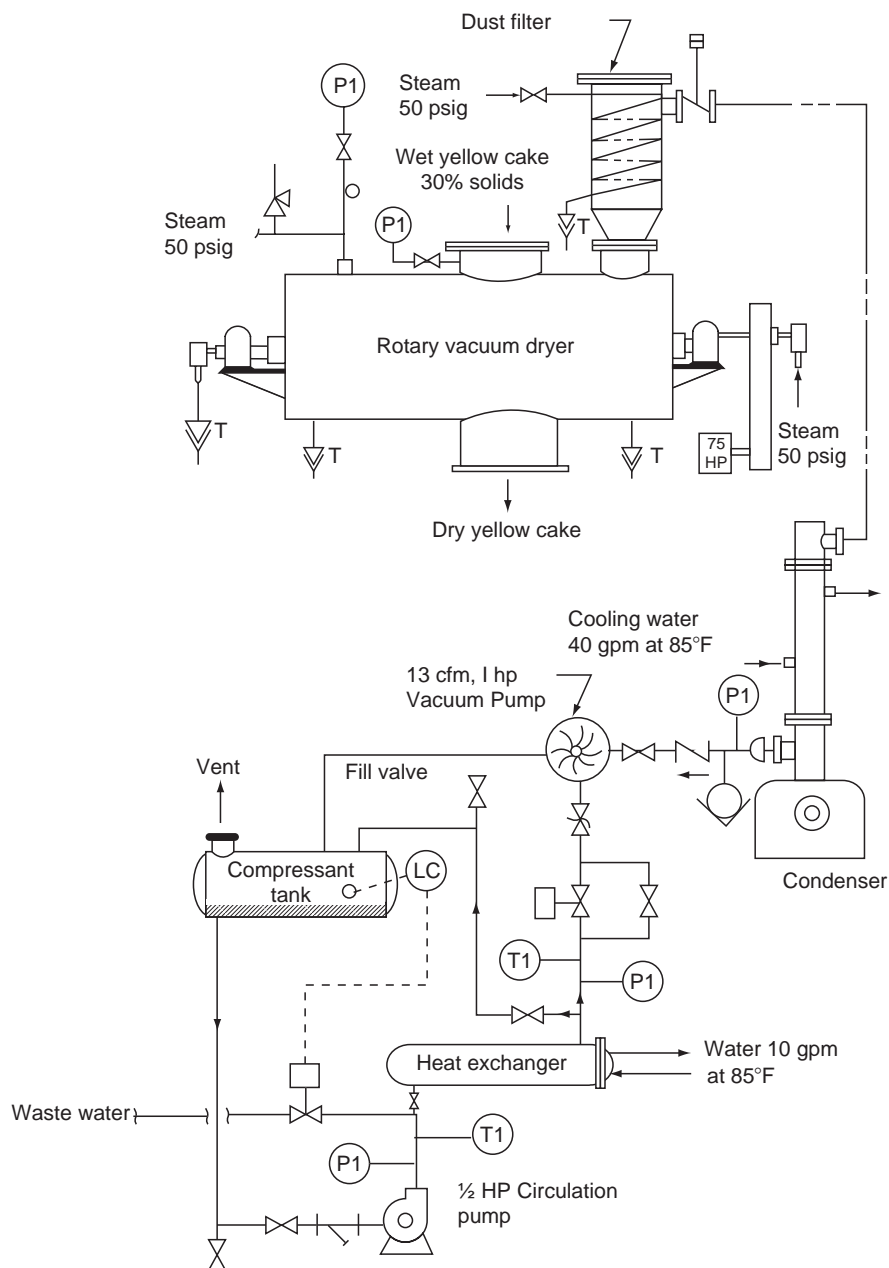


FIGURE 37.1 Drying of yellow cake slurry using a vacuum dryer.

solids or even slurries. Details of the fluid-bed drying process are presented elsewhere in this handbook. In mineral drying, the plenum temperature ranges from 1200 to 1800°F but the exhaust-air temperature of the discharge material varies from 180 to 250°F. Residence time varies from 5 to 25 min. Fluid-bed dryers are energy-efficient, consuming as little as 1500 Btu/lb of water evaporated. Capital and maintenance costs are low, but the feed size is limited to 3/4 in. Larger feed sizes are limited to lighter materials. Fluid beds are in commercial use for drying titanium dioxide, zirconium silicate, zircon, ilmenite concentrate, coal, sand, and other products.

For sticky solids with a tendency to cake or form lumps, mechanically stirred or vibrated fluid beds can be used for low- or medium-scale applications.

37.3.5 FLASH DRYERS

Flash or pneumatic dryers transport wet, pulverized solids in a hot airstream. For fine feed that is loose and readily conveyed pneumatically, flash dryers provide a means of drying as well as transport. Clearly, only surface moisture is typically removed in the short time span available in the dryer, say, several seconds. If the feed is not free-flowing, it is possible to integrate a hammer mill into the gas circuit to reduce feed size to small particles. Feeds that tend to cake can be mixed with dry solids to make them pneumatically transportable.

37.3.6 CONVEYOR DRYERS

For conveyor dryers, the wet feed is carried on a belt or a vibrating conveyor that is made to pass slowly through a heating chamber. Depending upon the design, hot air may or may not pass through the bed of solids on the conveyor.

37.3.7 SCREW CONVEYOR DRYERS

Screw conveyor dryers have hollow shafts or flights through which the heated gas flows. The trough that carries the material is similarly heated. The free-flowing wet feed is transported and dried as it is conveyed to the exit. It is especially well suited as a supplemental dryer for an existing dryer.

37.3.8 DRUM DRYERS

Drum dryers are used for slurries or pastes. The wet material is coated on the outer surface of a slowly rotating drum that is steam-heated. The slurry is either sprayed onto the drum surface or the drum is placed in a trough filled with the liquid slurry. The dryer must be designed to ensure that total drying is

accomplished in one revolution; the dry solid is scraped off before a new coating of wet material is applied.

37.3.9 ROTATING SHELF OR DISK DRYERS

For rotating shelf or disk dryers, the wet solid is supported on circular trays that rotate slowly in a fixed housing. The trays or shelves are cut out in the center and have radial slots. In the TURBO-DRYER version, the shelves are mounted rigidly to a single vertical shaft.* The fan assembly fits within the shaft assembly and can rotate independently of the shaft. Hot air or another drying medium is fed from suitably designed air distributors mounted on the side and is exhausted from the roof. Material flows from one shelf to the other below it. Stationary wiper blades wipe the tray clean of the material, which drops to the shelf below it through the radial slots. The dryer must be designed to yield dry product as the material exits from the lowest shelf.

These dryers are recommended for granular solids that need gentle handling, which are dusty when dry, or when drying with solvent recovery is necessary. Operating temperatures as high as 1100°F and as low as 0°F are feasible. Dryer operation and control are simple. Maintenance costs are low. High thermal efficiencies can be attained in practice.

These types of dryers are in commercial service for drying materials such as bauxite, borax, calcium carbonate, powdered coal, kaolin, mica, soda ash, thorium dioxide, zinc powder, and iron ore concentrate.

37.3.10 FUTURE TRENDS

Most dryers, especially those that directly contact hot combustion gases with the wet solids, operate best with liquid or gaseous fuels. With escalating prices of fuel oils and natural gas, dryers will increasingly use coal when possible. Direct-fired rotary dryers are the easiest to adapt to this source of fuel. Another trend is the development of more efficient drying systems for the use of nontraditional dryers. Better thermal economy is obtained with fluid-bed, spray, and flash dryers, but the rotary is currently more economical for very-high-tonnage operation of products with low unit value.

The use of solar energy to supplement the heat required for drying is unlikely to be economical in mineral-processing applications. One area that needs to be examined very carefully is the possible use of superheated steam instead of heated air as the drying

*TURBO-DRYER is a trademark of Wyssmont Co. Inc., Fort Lee, NJ.

medium. This concept is already in commercial use for drying pulp in the paper industry and for drying textiles. Technoeconomic studies have shown favorable indications for the use of superheated steam spray drying as well as fluid-bed drying (of granulated coal). With superheated steam, the mass-transfer resistance offered by the air film through which the evaporated moisture must diffuse is eliminated. The exhaust is also steam, which can be condensed to recover heat or compressed or reheated for total or nearly total recycling. Thus, the efficiency of superheated steam-based drying systems can be very high, though there are a number of mechanical-design constraints and operational problems that must be resolved before employment of such new technology in a large-scale operation. Finally, with high thermal efficiency dryers it is generally not economically justifiable to install heat-recovery equipment to recover the low-grade heat in the exhaust gases. If the exhaust gas is at high temperature for whatever reason, one must carefully evaluate the possibility of installing heat-recovery systems for such dryers. It is also likely that heat-pump technology will develop rapidly in the next decade and may find applications in the drying of minerals.

37.4 CONCLUSION

This chapter was intended to summarize the various dryer types found in the mineral-processing industry for the benefit of process engineers involved in the selection and specification of such equipments. Several competing dryer systems can generally accomplish the desired performance, albeit with different capital and operating costs. Energy costs should be considered, and energy-conservation measures should be employed when possible.

Several different dryer types can be used in many of the mineral-processing applications, e.g., drying

of concentrated nickel ore is carried out on industrial scale using spray dryers, flash dryers, rotary dryers, as well as fluid-bed dryers. In some cases, the selection of the dryer type depends on the cost of energy that is available locally as well as the availability of waste heat at the site. The relative cost of drying using these widely different drying systems must be quite different as well. Care should be exercised in selecting the most cost-effective drying system (e.g., nonthermal dewatering using filtration or centrifugation preceding thermal drying). Selecting a drying system used by a similar operation elsewhere in the world may not necessarily be the appropriate system. The interested reader is referred to a recent monograph by Kudra and Mujumdar [3] for discussion of some of the novel drying technologies available.

ACKNOWLEDGMENT

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*Note: See other chapters of this handbook for details and cross-references for specific types of dryers.

38 Dewatering and Drying of Wastewater Treatment Sludge

Guohua Chen, Po Lock Yue, and Arun S. Mujumdar

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38.1 INTRODUCTION

Sludge is the name that describes a muddy or slushy mass, deposit, or sediment as (a) the precipitated solid matter produced by water and sewage treatment processes; (b) mud from a drill hole in boring; (c) the muddy sediment in a steam boiler; (d) waste from a coal washery; or (e) the precipitated or settled matter from industrial processes. Water treatment sludge consists of suspended solids, coagulation chemicals, usually an alum or polymers with a limited amount of biological materials. A comprehensive review of industrial sludge can be found elsewhere [1]. In that review, sludge from petroleum, metal-finishing, flue gas cleaning, water treatment, pulp and paper processing, polymer plants, chemical plants, as well as mineral and metallurgical industries are discussed. The sludge addressed in this chapter is the by-product of a wastewater treatment plant. Brief reviews on the treatment, usage, and disposal of this type of sludge are available [2,3]. In this chapter, we take a comprehensive approach to examine the literature on sludge dewatering and drying in order to give readers a relatively detailed and complete picture of this area

that is growing with the expenditures on environmental cleanup and control, amounting to about US \$150 billion in the United States and about US \$400 billion globally in 1997 [4].

38.2 WASTEWATER TREATMENT PROCESS AND SLUDGE PRODUCTION

Among the sludges generated from wastewater treatment plants, there are two types: (a) municipal sewage sludge and (b) industrial sludge, depending on the sources of the wastewater. By taking a municipal wastewater treatment plant as an example, there is primary sludge obtained from gravity settling at the beginning of the process, secondary sludge generated from biological treatment (usually activated sludge), [see Figure 38.1](#), and also tertiary sludge generated by processes such as chemical precipitation and filtration [5]. The sludge produced may undergo further biological digestion either aerobically or anaerobically in order to reduce the volume as well as the pathogens in the sludge [6]. The digestion of raw sludge will also convert some of the organic carbon into methane or CO₂.

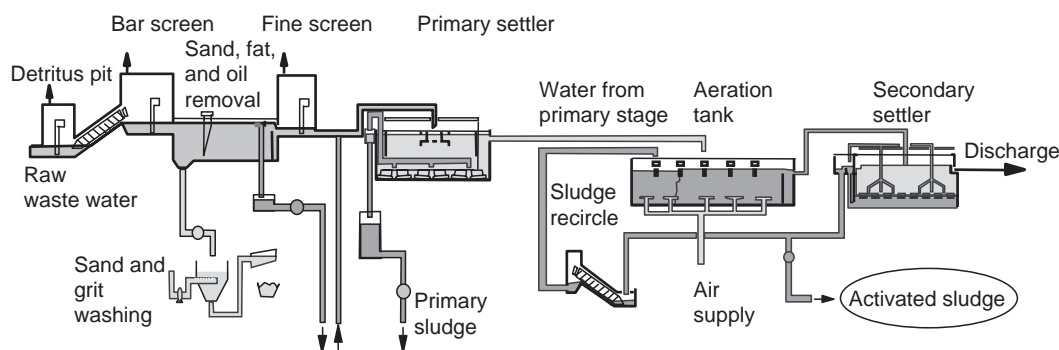


FIGURE 38.1 Typical wastewater treatment process.

Approximately 5.4 million tons of dry sewage sludge are generated annually from about 12,750 public owned treatment works (POTW) in the United States [7]. The average sludge production is about 0.090 kg dry sludge/person/day in European countries [8]. Since raw sludge comprises over 90% water, the volume of the raw sludge produced is enormous. Typical raw-activated sludge production is in the range of 1.2 to 2.4 kg/person/day. Table 38.1 shows the production and utilization of sludge produced in EU, the United States, and Japan. The data on the EU and the United States are from Ref. [5], while the data on Japan are from Ref. [9]. Currently, the sludge is used in agriculture, placed in landfills, incinerated and, in some places, dumped in the sea. Sea dumping has been banned by many countries and is a trend being followed worldwide. Landfill remains the primary sludge disposal practice in most of the countries where suitable sites are avail-

able. Since the volume of sludge is very high worldwide, there is a growing interest in either increasing agricultural usage or processing the sludge to produce other useful products.

The resources found in municipal wastewater treatment sludge, more recently called biosolids, are nutrients (N and P) and energy (C) along with some minor components as shown in Table 38.2 [10]. Therefore, 1.17 million dry tons, or 21.6% of total wastewater treatment sludge generated in the United States, are used to improve the soil for crops and pastures. The application of sludge for agricultural use in developing countries should be even more widespread. In addition, there are also applications for nonagricultural land, such as forest land, parks, golf courses, and reclamation sites. Agricultural use includes the direct spray of raw sludge, the use of digested sludge as fertilizer, or composting the sludge with some municipal solid waste (MSW)

TABLE 38.1
Global Sludge Production and Utilization

Country	Dry t/y ×1000	Agriculture Use (%)	Landfill (%)	Incineration (%)	Sea (%)	Other (%)
United States	5358	33.3	34	16.1	6.3	10.3
Germany	2700	27	54	14		5
Japan	~2000		25	75		
UK	1107	42	8	14	30	13
France	852	60	20	20		
Italy	816	33	55	4		8
Spain	350	50	35	5	10	
The Netherlands	323	26	50	3	2	19
Denmark	170	54	20	24		2
Belgium	200	29	55	15		1
Greece	48	10	90			
Ireland	37	12	45		35	8
Portugal	25	11	29		60	
Luxembourg	8	12	88			

TABLE 38.2
Municipal Sludge Characteristics

Parameter	Primary	Secondary
Total solids (TS), %	3.0–7.0	0.5–2.0
Volatile solids (% of TS)	60–80	50–60
Nitrogen (N, % of TS)	1.5–4.0	2.4–5.0
Phosphorus (P ₂ O ₅ , % of TS)	0.8–2.8	0.5–0.7
Potash (K ₂ O, % TS)	0–1.0	0.5–0.7
Heat value (kJ/kg, dry basis)	23,000–30,000	18,500–23,000
pH	5.0–8.0	6.5–8.0
Alkalinity (mg/l as CaCO ₃)	500–1500	580–1100
<i>Metal contents (mg/kg, dry basis)</i>		
	<i>Range</i>	<i>Median</i>
Arsenic	1.1–230	10
Cadmium	1–3410	10
Chromium	10–99,000	500
Copper	84–17,000	800
Lead	13–26,000	500
Mercury	0.6–56	6
Molybdenum	0.1–214	4
Nickel	2–5300	80
Selenium	1.7–17.2	5
Zinc	101–49,000	1700
Iron	1000–154,000	17,000
Cobalt	11.3–2490	30
Tin	2.6–329	14
Manganese	32–9870	260

to form humus-like material for land remediation [11]. The limitation of large-scale agricultural utilization of sludge lies in the high heavy metal contents and pathogens in the sludge. Biosolids can also be used as landfill cover material. They can be used as an energy source as is the case for incineration of sludge together with MSW to generate electricity. Incineration can greatly reduce the volume of sludge, thus relieving the demand for landfill. This option is suitable for regions such as Hong Kong, Singapore, and Japan where landfill sites are not available. However, the well-known air pollution problem arising from incineration has to be addressed properly. The ash produced can be used as raw material for the production of building and road-surfacing products. Other methods of sludge treatment include the production of oils and fertilizers from sludge [12,13] and construction materials from melted incinerated sludge ashes [14]. An extensive report on alternatives for sludge use is available elsewhere [15].

38.3 MOISTURE DISTRIBUTION IN SLUDGE

The moisture in sludge can be as high as 99%. Figure 38.2 shows a conceptual visualization of moisture distribution in sludge. This distribution takes the following forms [16]: (a) free moisture that is not attached to the sludge particles and can be removed by gravitational settling; (b) interstitial moisture that is trapped within the flocs of solids or exists in the capillaries of the dewatered cake and can be removed by strong mechanical forces; (c) surface moisture that is held on the surface of the solid particles by adsorption and adhesion; and (d) intra-cellular and chemically bound moisture. The amount of water that can be removed depends on the dewatering process and also the status of the water in the sludge. The water that can be removed by mechanical dewatering is usually termed as “free water” and the remaining as “bound water.” The bound water content is the theoretical limit of mechanical dewatering [17]. The free water so defined includes truly the free, interstitial, and partially the surface moisture. The bound water includes the chemically bound moisture and partially the surface moisture.

The bound water content can be determined by methods such as *dilatometric determination, vacuum filtration, expression, drying, and thermal analysis*. The dilatometric method is based on the assumption that free water can be frozen at -20°C . By placing a known amount of sludge into a scaled container called a dilatometer and measuring the volume of expansion at -20°C , the free water content can be calculated [18]. The part that does not freeze is the bound water. The sludge is usually mixed with a fluid that does not freeze at -20°C , is immiscible with water, has specific gravity less than 1.0, and shows

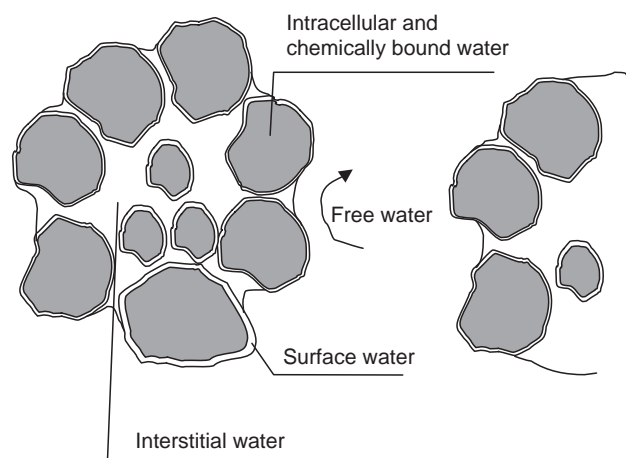


FIGURE 38.2 Water distribution in sludge.

linear expansion/contraction over the temperature range of -20 to 20°C . Thus, the volume change in the frozen sludge can be easily determined. Xylene was the fluid suggested for use in this process [19]. Hydraulic oil has been found to be suitable for such a purpose [20]. The rate of sample cooling is usually controlled at $1^{\circ}\text{C}/\text{min}$ although it has an insignificant effect. The vacuum filtration and expression methods force the free water to pass through porous media, which retain the solids with the bound water. For the expression test, either a constant head piston press [21] or a centrifuge can be used [22,23]. The bound water content can also be determined by drying a thin layer of sludge under constant air temperature (30 to 40°C) and constant humidity (RH of 50 to 60%). The drying of the free water is accomplished at a constant rate while the bound water drying rate decreases with a decrease in the moisture content. The critical moisture content equals the bound water content [16,24–28]. Thermal analysis may be used to measure the amount of heat released when the free water is frozen or the amount of heat required to melt the frozen water. By assuming that the free water would freeze at -20°C , the heat released or required during the freezing or thawing process can be determined by a calorimeter [29] or by a differential scanning calorimeter [30,31].

Obviously, the bound water content often depends on the operation because of the differences in the analytical methods [20,31,32]. Such a dependence was found to be related to the mechanical force applied and the bond strength between the water molecules and the solid particles. In vacuum filtration, the dependence corresponds to a bond strength close to zero (1 kJ/kg); in the drying test, it corresponds to a bond strength of 20 to 40 kJ/kg ; and in expression, it corresponds approximately to 70 kJ/kg , which is almost the value of the heat of physisorption [21]. A detailed sludge testing and handling guide can be found elsewhere [33].

38.4 SLUDGE DEWATERING

Sludge drying is an energy intensive process. Consequently dewatering, i.e., removing water without evaporation, becomes an important process before thermal drying is utilized. Prior to sludge dewatering, the digested sludge is usually conditioned to generate flocs that are easy to filtrate. Chemical conditioning using polyelectrolytes, Fe (III), Fe (II), lime, or Al (III), is the most common method (see Figure 38.3). The chemicals act as either coagulants by reducing the zeta potential of the solid particles or flocculants through the bridging effect to form proper-sized flocs. Thermal conditioning by simple heating or wet air oxidation can also be used [34]. An effective method that can be used in cold regions in winter is freezing/thawing [35]. During the freezing period, the ice crystals formed will push the contaminants, in this case the sludge particles, out of the lattices to form solid blocks. The sludge should be fully frozen. The thawing and quick removal of water will end up with a dewatered sludge of more than 20% solid content [36,37]. The thawed sludge should not be agitated in order to keep its filterability [38].

There are several alternatives to mechanical dewatering of sludge. They are: vacuum filters, belt filter presses, centrifuges, and membrane filter presses. Vacuum filters are among the earliest mechanical devices employed (Figure 38.4a). Although many of its installations are being replaced by the more energy-efficient belt filters, the use of vacuum filters is popular when applied with precoat filtration. Another configuration is the rotary belt vacuum filter. The difference between a belt vacuum filter and a rotary drum vacuum filter is that the filter medium belt is wrapped around the surface of the drum rather than fixed to it. This gives the advantages of continuous belt washing and more efficient cake discharge [39].

The operation of belt filter presses is similar to pressing in papermaking (Figure 38.4b). The

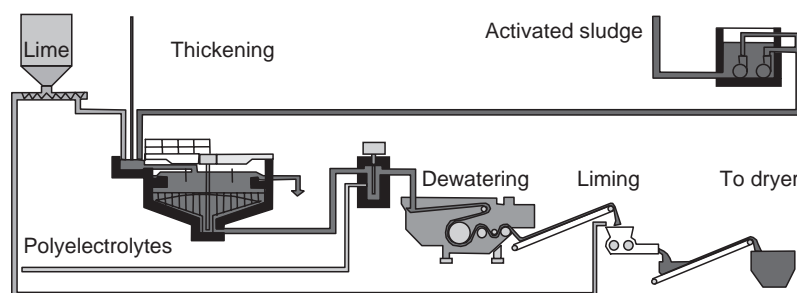


FIGURE 38.3 Typical mechanical dewatering of sludge.

conditioned sludge is first drained under gravity before being sandwiched between two endless filtering belts. The pressure from the tensioned belts squeezes the water out of the sludge with a continuous cake discharged at about 15 to 25% solid content when the feeding solid concentration varies from 2 to 5%. Plate-and-frame filter presses can also be used if the volume of the sludge to be treated is not high (Figure 38.4c). The rotary press is a recent development (Figure 38.4d). Biosolids are pumped into a peripheral channel that has walls made up of rotating filter elements. As the mechanism rotates, compression is created and liquid is forced through the filter elements. A cake is formed in the interior channel and then extruded [34]. The belt filter press can be enhanced by acoustics [40] or electrical osmotic dewatering techniques ([41]; Figure 38.4e).

Centrifugal dewatering uses the centrifugal force developed by spinning a bowl or basket to separate the sludge solids from the liquids. Disc, basket, and solid bowl centrifuges are all used for sludge dewatering with the latter being the most common. Solid bowl centrifuges are available in cocurrent or countercurrent flow designs ([42]; Figure 38.4f, Figure 38.4g). A typical centrifuge has two operating zones: a submerged pool and a drainage zone. The current generation of centrifuges can achieve solid contents of 25 to 35% [34].

Membrane filter presses are modified filter presses with the introduction of inflatable membrane systems,

automatic cloth washing, controlled filling techniques, dual-speed plate separation, and automatic discharge [43]. The use of cross-flow microfiltration technology has been used in South Africa and in the UK to concentrate sludge solids [3]. Even with technical advances over the last century, sludge dewatering technology still continues to be improved. No simple method exists for selecting the best dewatering process for a wastewater treatment plant. The dewatering process chosen will be a function of sludge disposal and capacity, along with the plant's operation and maintenance capabilities [44].

38.5 SLUDGE DRYING

38.5.1 SOLAR ENERGY DRYING

Thermal drying of dewatered sludge is necessary when the required final product water content is low as in fertilizer production, when the heating value of the sludge needs to be improved for efficient incineration, or in the case of reduction of transportation costs in disposal such as landfilling. Thermal drying can also help to inactivate indigenous viruses [45]. As mentioned before, the thermal drying of sludge is a very energy-intensive process, hence utilizing solar energy is often the first natural choice [46]. It has been found that a solar energy collector can help increase the drying rate in order to decrease the

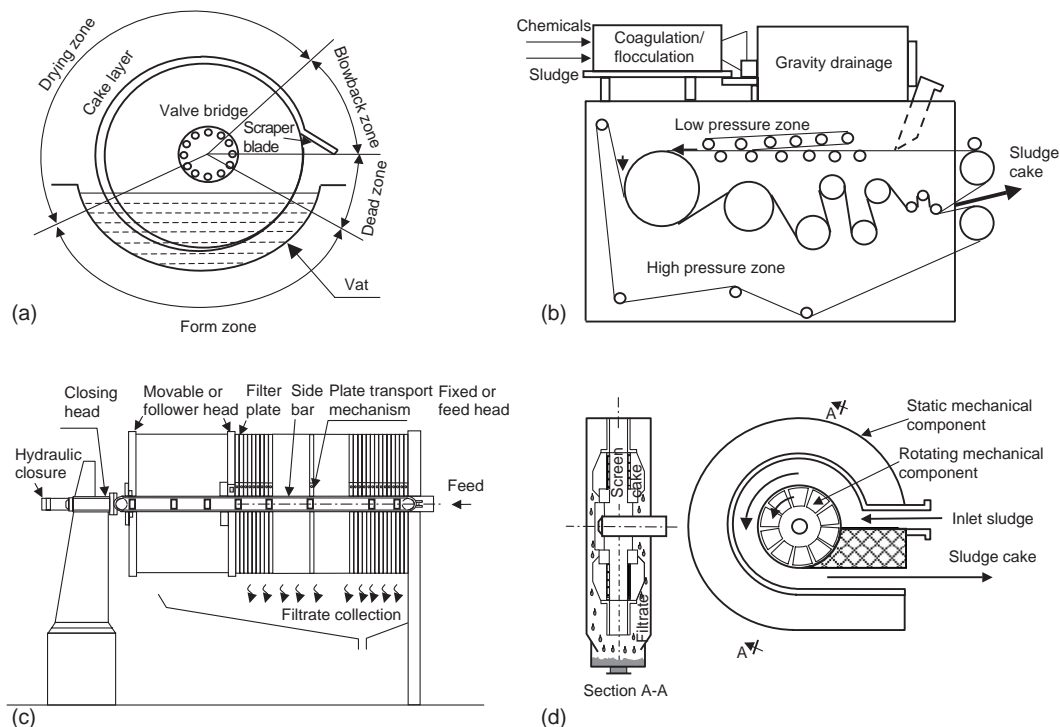


FIGURE 38.4 Typical sludge dewatering equipments: (a) vacuum filter; (b) belt press filter; (c) press-filter; (d) rotary press;

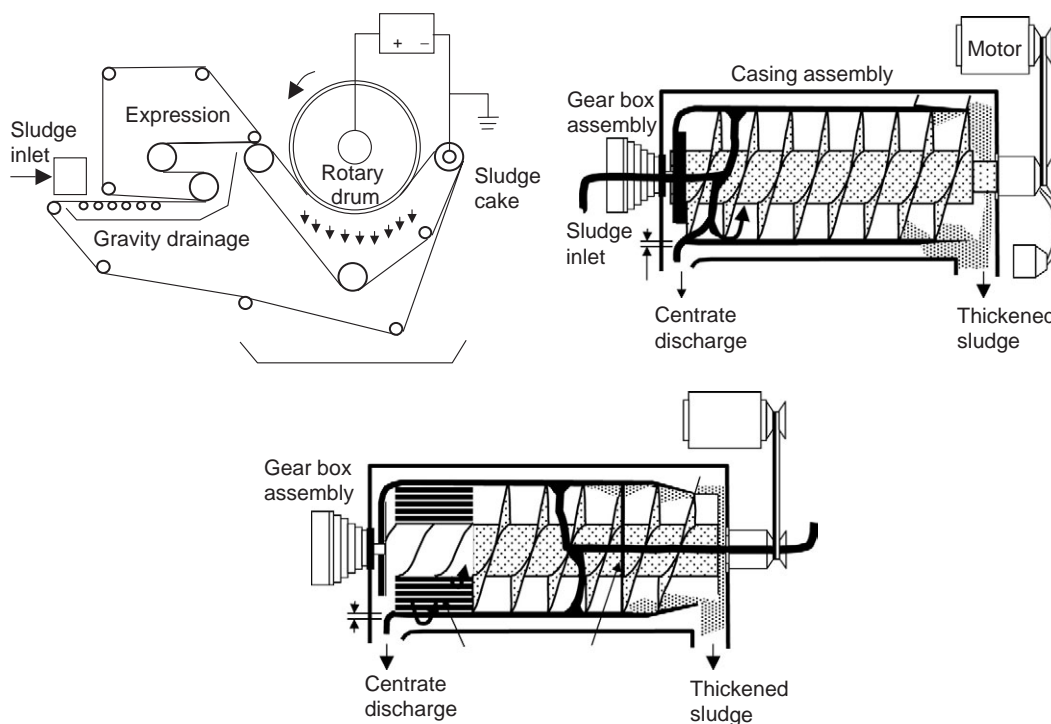


FIGURE 38.4 (Continued) (e) rotary drum press combined with EOD; (f) cocurrent centrifuge; (g) countercurrent centrifuge.

drying time at some locations [47,48], whereas at other locations such a system can offer only limited help [49]. A drying bed with asphalt pads can make sludge processing easier than if clay-lined sites are used [50]. A filtration sand bed that drains sludge at a small angle around 1° was found to perform better by placing a cellular confinement system at 3, 4, 6, and 8 inches cell depth [51]. This grid-confined sand easily supports a small loader used to remove dried sludge that typically has a solid content of 35% and a consistency similar to soil. In open drying beds, periodical turning of the sludge has been found to increase the drying rate, decrease the insect populations, and reduce the odor problem through increased aeration and the disturbed incubation of insects [52]. Otherwise, the formation of thin surface layers would hinder the drying of the remaining sludge [53].

38.5.2 THERMAL DRYING FUNDAMENTALS

The drying of sludge using solar energy requires a considerable amount of land and may give rise to an odor problem that is difficult to solve. In recent years, thermal drying has received much attention and is becoming a major sludge-processing technology [54]. The thermal drying of sludge will normally undergo a constant rate with a first falling rate followed by a second falling rate period as shown in Figure 38.5

[55]. During the constant drying rate period, free water is removed. The first falling rate period is believed to remove the interstitial water and the second falling rate period removes the surface water. The final moisture content retained within the sludge is mostly chemically bound water in amounts that depend on the type of sludge and the drying conditions. Besides the variations in moisture content and drying rates, the physical status of the sludge will change from a wet zone to a sticky zone before granulation depending on the solid content [54]. In the wet zone,

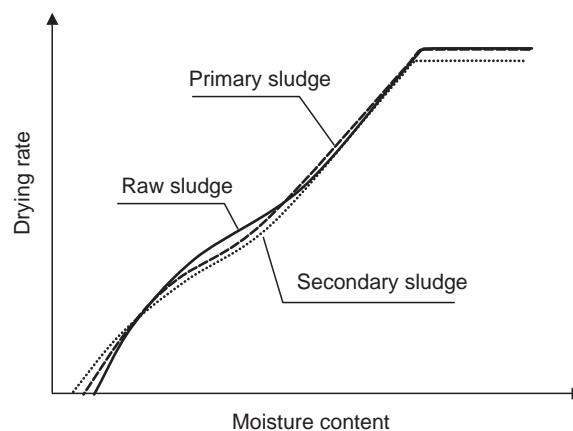


FIGURE 38.5 Typical drying rate curve.

the sludge can be spread easily on to a heated tube in an indirect dryer. Once the sticky zone (solid content between 55 and 70%) is reached, the sludge turns into a paste and strong shearing is necessary in order to have a high degree of mixing. Therefore, in an indirect drying system, the dried sludge is frequently mixed with raw sludge to have a solid content of about 70% in order to avoid the sticky zone. With further removal of moisture, the sludge becomes crumbly in nature and can mix much more freely.

There are numerous dryers available globally that claim to be able to dry sludge. In terms of heat and mass transfer, the available dryers can be classified as (a) direct drying systems, (b) indirect drying systems, and (c) combined systems that found in fluidized-bed dryers. Direct dryers are simple in design but the vapors released from the sludge have to be separated from the drying medium, especially in the situation when the drying medium is to be recycled to save energy. Direct dryers are typically rotary-drum, flash, moving-belt dryers, or centridryer types. Typical indirect dryers are thin-film, rotary-disc, or rotary-tray dryers. The indirect drying system has the advantage of producing minimal amounts of vapors and is therefore easy to manage [3]. The drying rate of indirect dryers may be lower than that of direct dryers because the latter can operate at much higher temperatures.

38.5.3 DIRECT SLUDGE DRYERS

Figure 38.6a shows a rotary-drum dryer. Inlet temperatures of up to 1000°C can be used without the risk of ignition with a water evaporation rate ranging from at least 800 kg/h to as much as 50,000 kg/h. Recycling part of the vent gas may be used to improve the thermal efficiency of the dryer. In industrial sludge drying in Japan, rotary-drum dryers equipped with a disintegration device are widely employed (Figure 38.6b). The disintegrator in the drum consists of antenna-like bars and rotates at 200 to 400 rpm. Such a dryer can give an evaporation rate of 100 to 5000 kg/h [56]. One variation is to introduce hot gas into the rotary drum through perforated tubes to ensure uniform gas distribution while the drum wall is equipped with spiral flights for sufficient mixing [57]. If the incoming sludge contains a significant amount of fibers such as sludge from the papermaking industry, an attrition mill may be installed ahead of the dryer [58]. Figure 38.6c and Figure 38.6d show a triple-pass rotary drum dryer. It consists of two concentric cylinders within one large outer cylinder. These cylinders are aligned to rotate at the same speed. As sludge passes through the most inner cylinder, it is exposed to the highest gas temperatures. When the sludge is conveyed pneumatically to the

outside cylinders by the fan, it is exposed to lower gas temperatures and air velocities. Depending on the hot gas temperature and the dryer size, the evaporation rate can vary from a few hundred kilograms per hour to more than 25,000 kg/h.

The most basic flash drying system for use with materials that require disintegration, which is the case for sludge, is the cage mill system. The feed is agitated in the hot gas stream by the cage mill, thereby increasing turbulence and retention time. The circular motion of the rotor assists the partially dried sludge to move up the dryer where additional drying occurs. A mixer may be used for wet feed where a portion of the already dried sludge is diverted back into the mixer. In flash drying, heat recovery is helpful in improving the energy efficiency. Depending on the application, this is normally accompanied either by using vent gas recirculation (Figure 38.7a), by employing a deodorizing preheater as a heat exchanger (Figure 38.7b), or tying the flash drying system directly with a steam-generating boiler that can incinerate both the dried sludge and any odorous drying gases (Figure 38.7c). The maximum water evaporation rate is approximately 0.1 kg water/m³ air flow at the vent. A recent development in flash dryers is presented in Figure 38.8. The strong impact force and air turbulence generated by a high speed dispersion rotor enables better performance of the feed material dispersion in the drying chamber without a cage mill. A remarkable increase in drying efficiency has been observed. By adjusting the classifying rotor located at the top of the dryer, the moisture content and the particle size of the product can be regulated. Such a dryer is claimed to be adhesion-free with an evaporation rate of up to 6000 to 8000 kg/h. It operates similar to that of a hot-gas-grinder dryer (Figure 38.9). The difference is that the former can handle wet sludge while the latter is reported to operate with dried sludge of moisture content around 20 to 30% [56]. A spin flash drying system has recently been patented based on similar principles [59].

Belt dryers are available in open, semi-open, and closed-loop systems. Figure 38.10a shows a closed-loop system with the heating air dried and recirculated [60]. Figure 38.10b shows another system that was designed for sludge drying using infrared (IR) heating. Air from the supply plenum passes across the face of the IR heaters and is directed downward toward the sludge in a belt passage way. A lower plenum is provided below the upper run of the belt. It is maintained in a vacuum to draw air through the sludge. About 10 to 30% of the air is exhausted from the upper plenum to release moisture from the apparatus with the same portion of air made up by drawing air from the entrance and exit of the belt passageway

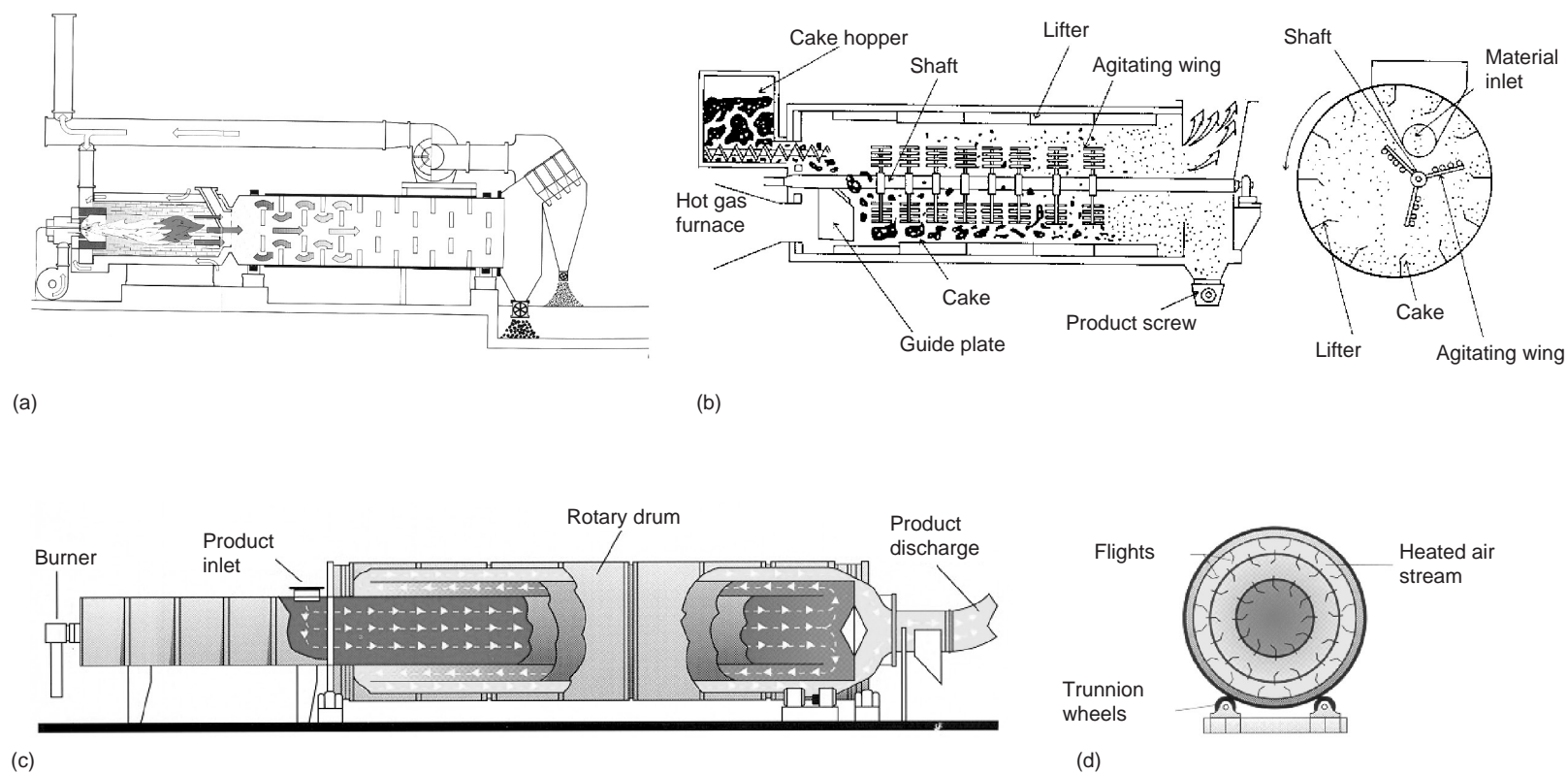


FIGURE 38.6 Rotary drum dryers: (a) single-pass (courtesy of Vadeb); (b) dryer equipped with disintegration device (adapted from Imoto, Y., Kasakura, T., and Hasatani, M., *Drying Technology*, 11(7):1495–1522, 1993); (c) triple-pass (courtesy of M-E-C Company); (d) cross-section of a triple-pass dryer (courtesy of M-E-C Company).

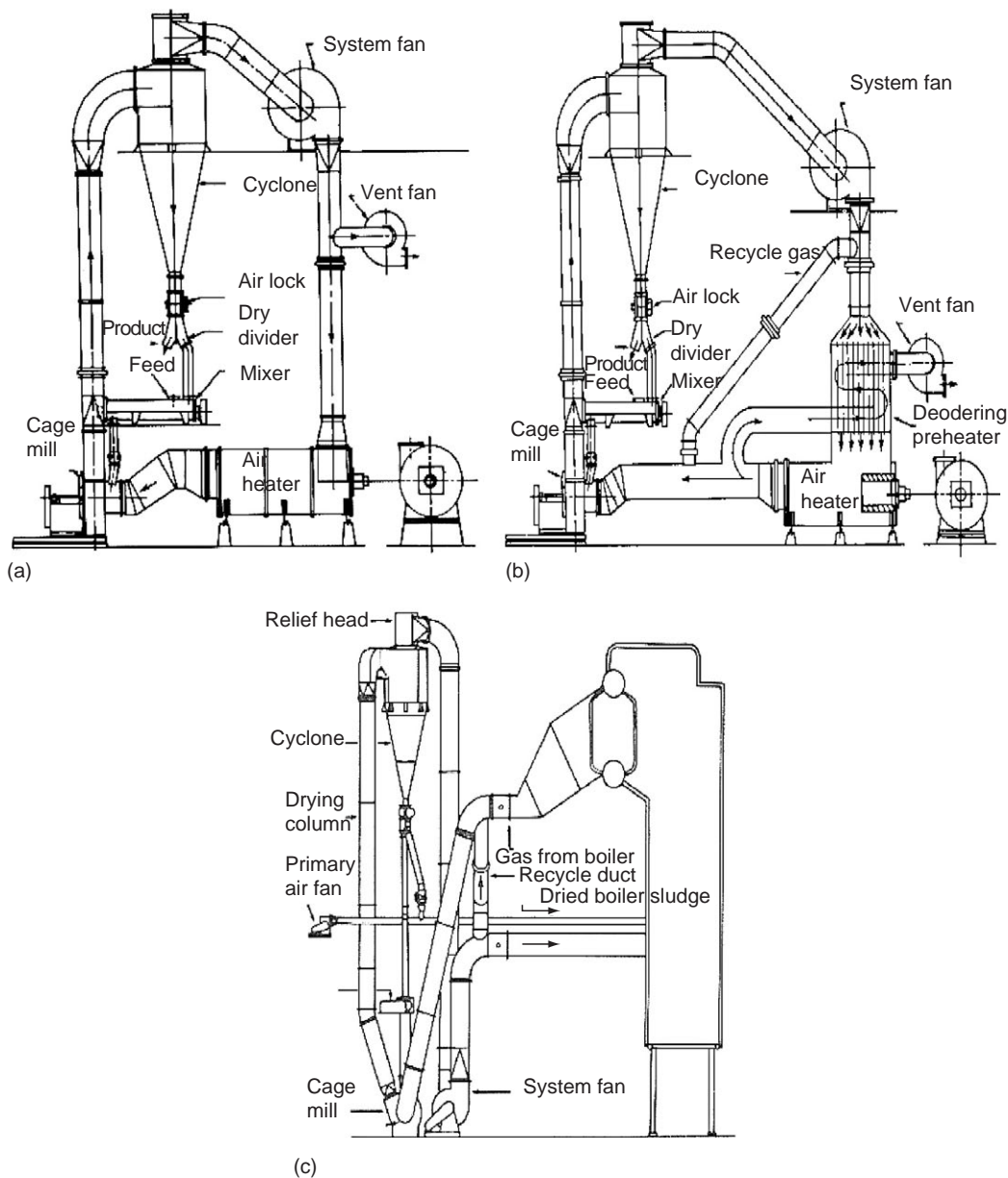


FIGURE 38.7 Cage-mill flash drying system (courtesy of ABB Raymond): (a) with recirculation; (b) with deodorizing preheater; (c) with accompanying fired boiler.

[61]. Figure 38.10c shows a moving-belt sludge dryer with the belt made up of cellular pockets. The sludge-filled pockets are supported on a heated pan. Heated air is supplied from both the top and the bottom surfaces of the pocket. Heat transfer is achieved by convection and conduction. The sludge may undergo multiruns before it is dried and rejected from the pockets [62]. A belt dryer with stacked heating chambers claiming to have a better thermal efficiency (Figure 38.10d) has been patented [63].

Another type of direct dryer is the spray dryer [64] but there is no such installation in the United States for sludge treatment [65]. A recent patent of cyclonic dryer may be used in sludge drying [66]. In this cyclonic dryer, a high-speed heated air stream is introduced to the cylinder such that it rotates tangentially around a central axis immediately upon entry into the cylinder. The wet material such as sludge is fed to the dryer directly into the outer spiral air stream with turbulence generated by ramps on the wall of the

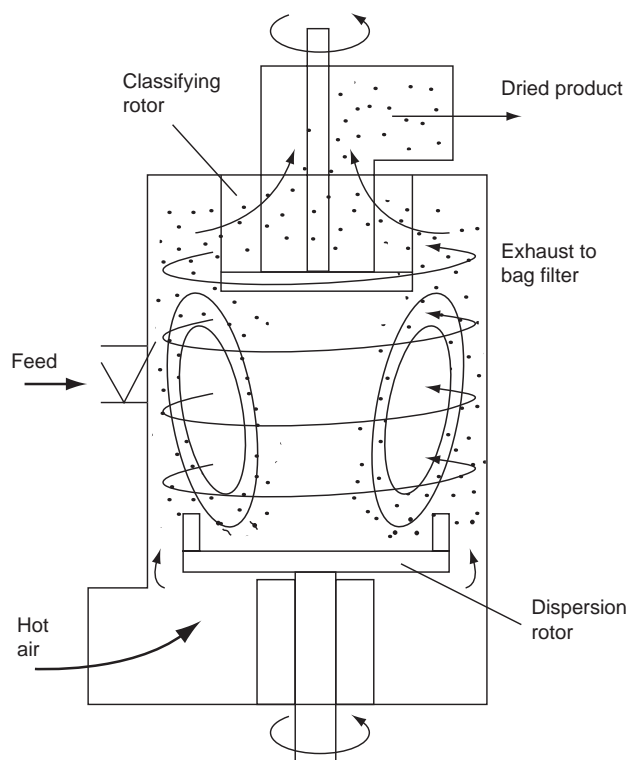


FIGURE 38.8 Schematic diagram of Drymeister (DMR). (Courtesy of Hosokawa Micron B.V.)

cyclone. The concept of combining drying and dewatering has been realized by Centridry[®] using a centrifuge in the heart of the dryer and the dewatered sludge granules are directly dried immediately after exiting the centrifuge. The dewatering and drying parts are separated by the stator case of the centrifuge ([67]; Figure 38.11).

38.5.4 INDIRECT SLUDGE DRYERS

Since indirect dryers generate less gas, most of their designs are closed-loop with either heat recovery and/or odor removal units. Because indirect dryers depend on heat being transferred from a heated surface and the dewatered sludge is still relatively wet (around 25% solid content for activated sludge), the handling of the interfacial behavior of the sludge and the heated surface is an important issue. While there is no air flow to disperse or disintegrate the wet sludge, mechanical agitation has to be designed to prevent the heating surface from being fouled, especially in the sticky zone with the solid content ranging between 55 and 70%. Depending on the requirement of the final moisture content, which is related to the fate of the dried sludge, there are single or two-stage indirect drying systems employed. If the sludge is dried to

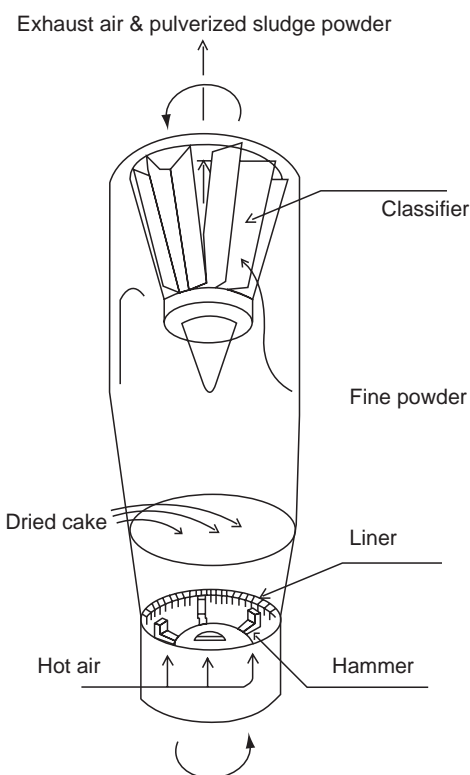


FIGURE 38.9 Hot gas grinder dryer. (Adapted from Imoto, Y., Kasakura, T., and Hasatani, M., *Drying Technology*, 11(7):1495–1522, 1993.)

improve its burning properties in an incinerator, a single-stage thin film evaporator can accomplish the job with a final solid content up to 65%. This technology has been established in the processing of other products [68]. For sludge processing, usually a horizontal agitated thin film evaporator is selected. Its drying rate lies between 20 and 160 kg/m²/h [54].

When higher final solid contents are desired, a rotary paddle or disc dryers may be used either alone or as the second stage following a thin-film evaporator. Figure 38.12 shows the design principle and necessary parts usually found in an indirect dryer. It consists of a stator with scrapers and a rotor with agitators and driving units. The heat transfer surface can be self-cleaned with the gentle friction forces created from the rotor motion. Adequate mixing can be achieved by the action of the agitators and the scrapers. The nearly consecutive chambers ensure almost a plug flow of the sludge through the dryer, which is an important issue in sludge sterilization [69]. The dryers so designed are available as shown in Figure 38.13a through Figure 38.13d. Some of the dryers shown in Figure 38.13 claim to be able to dry sludge with moisture contents down to less than 10%.

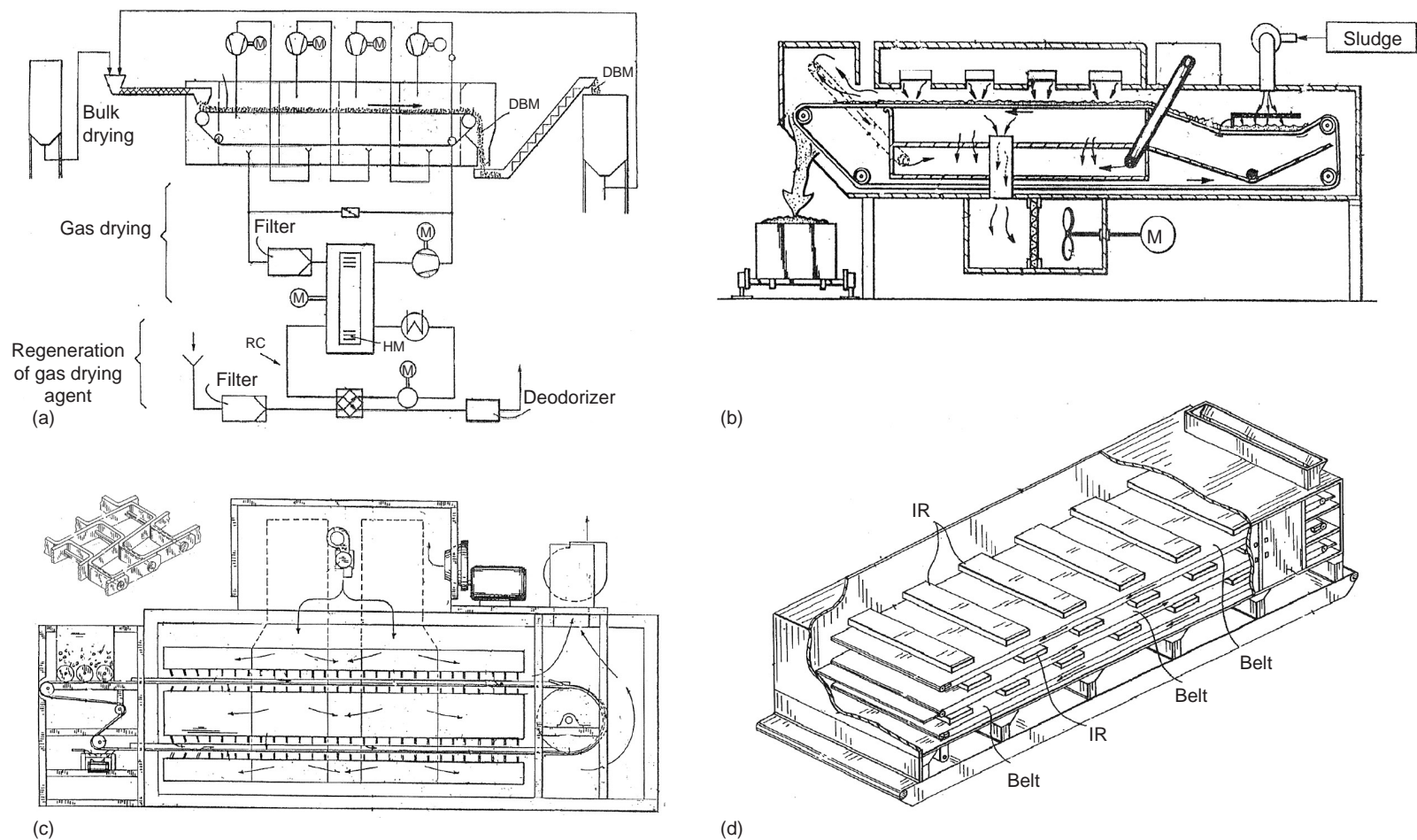


FIGURE 38.10 Different arrangements of belt dryers: (a) moving belt dryer with vapor removal and heating gas recirculation system, DBM-dried biomass; HM-hygroscopic material; M-motor; MBM-moist biomass (adapted from Rutz, A., US Patent 5,428,904, 1995); (b) closed-loop belt dryer system with IR heating; (c) moving belt sludge dryer with direct and indirect heating (adapted from Bein, D.J., US Patent 5,365,676, 1994); (d) Multi-pass moving belt drying with IR heating (adapted from Nugent, J.E., US Patent 5,634,281, 1997).

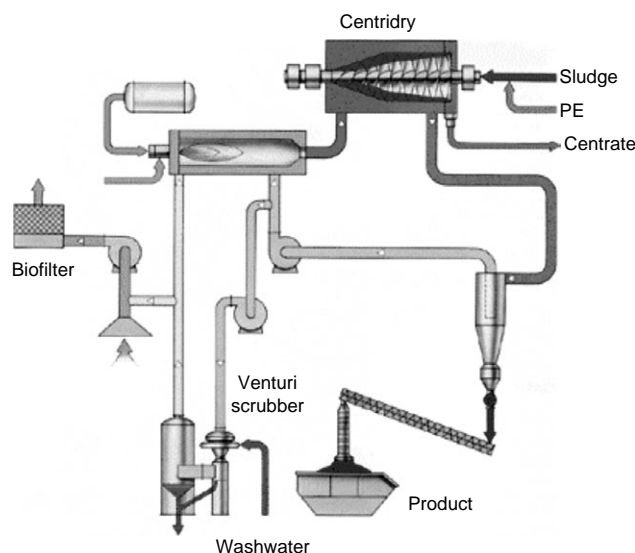


FIGURE 38.11 Centridryer.

Indirect drying with simultaneous sludge drying and pelletizing emerged in the last decade in Europe as a result of environmental and energy conservation concerns. The dryers used in such a practice are vertical multistage tray dryers (Figure 38.14). The sludge is fed via the top inlet and moved by the rotating arms from one heated tray to another in a zigzag path until it exits at the bottom as a dried and pelletized product with up to 95% total solid content. The dryer trays are hollow and are heated by steam or thermal oil [70,71]. The sludge can be uniformly spread on the heated surface with its layer thickness controlled properly. Hence, more uniform drying is achieved in such a dryer.

38.5.5 OTHER SLUDGE DRYERS

In addition to the direct and indirect dryers reviewed above, there are systems available that either use combined conduction and convection heat transfer or use a different drying medium. Examples are the fluidized bed dryer shown in Figure 38.15. The wet sludge can be continuously fed and withdrawn from the fluidized bed system with few moving parts. The heat for water evaporation is supplied mostly from the hot surfaces of thermal oil tubes. The heated air of about 85°C provides some heat and mainly ensures that the sludge granules stay in a fluidized state to enhance the heat transfer between the sludge and the hot tube surface (Figure 38.15a). Sludge can be dried to 5% moisture content in such a dryer [67]. Contrary to the simple design seen in Figure 38.15a, a very sophisticated drying system using superheated steam has also been developed (Figure 38.15b). Wet sludge is fed into one of the cells around the superheater and suspended by the superheated steam blown through the perforated base plate by the impeller. The partially dried sludge moves successively through each cell before it is discharged [72].

The description of a concept of combining sludge drying with sludge incineration is shown in Figure 38.15c. The drying is achieved by mixing heated sand with wet sludge in a rotary kiln. The dried sludge is burned in a fluidized bed with sand. The flue gas heats the sand on top of the burner [73]. This combination is supposed to have higher energy efficiency. In fact, the integrated fluidized bed drying/incinerator is currently a popular choice in Japan because of its compactness and high energy efficiency with no bad odors generated. In this industrialized dryer/

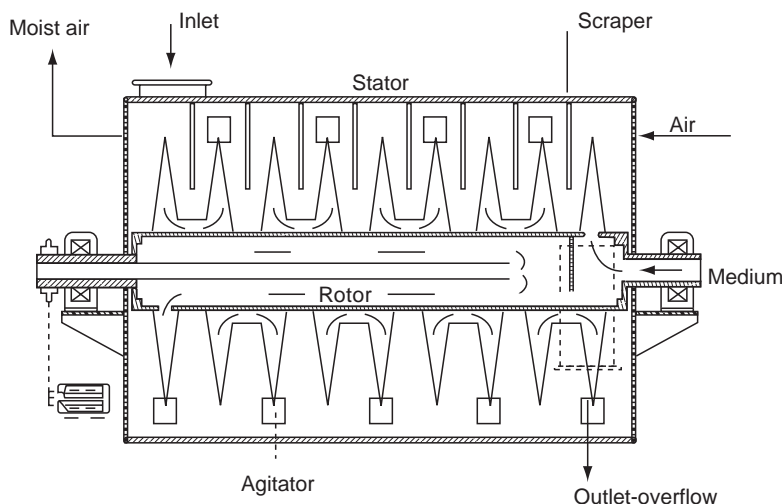


FIGURE 38.12 Design principle of Rotadisc.

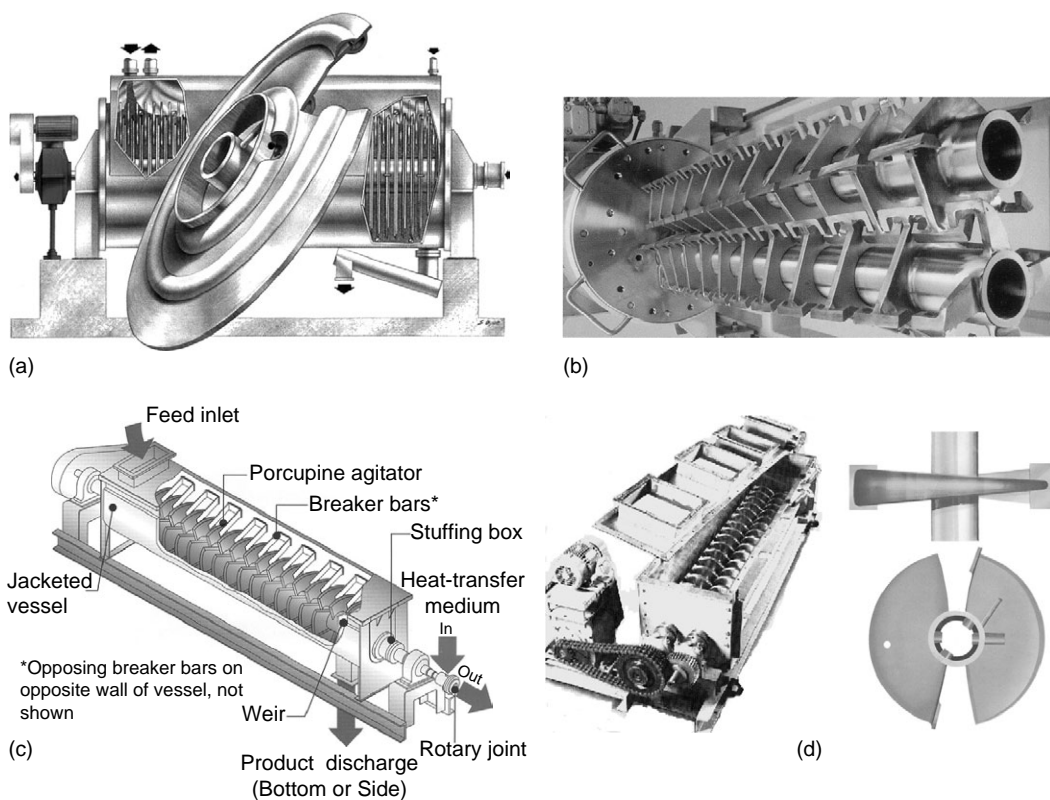


FIGURE 38.13 Various designs of indirect dryers: (a) rotaplate indirect dryer (courtesy of Stord); (b) kneading and self-cleaning disc dryer—twin shafts (courtesy of LIST); (c) the Porcupine processor (courtesy of The Bethlehem Corp.); (d) K.-S. Nara paddle and paddle dryer (courtesy of Komline-Sanderson).

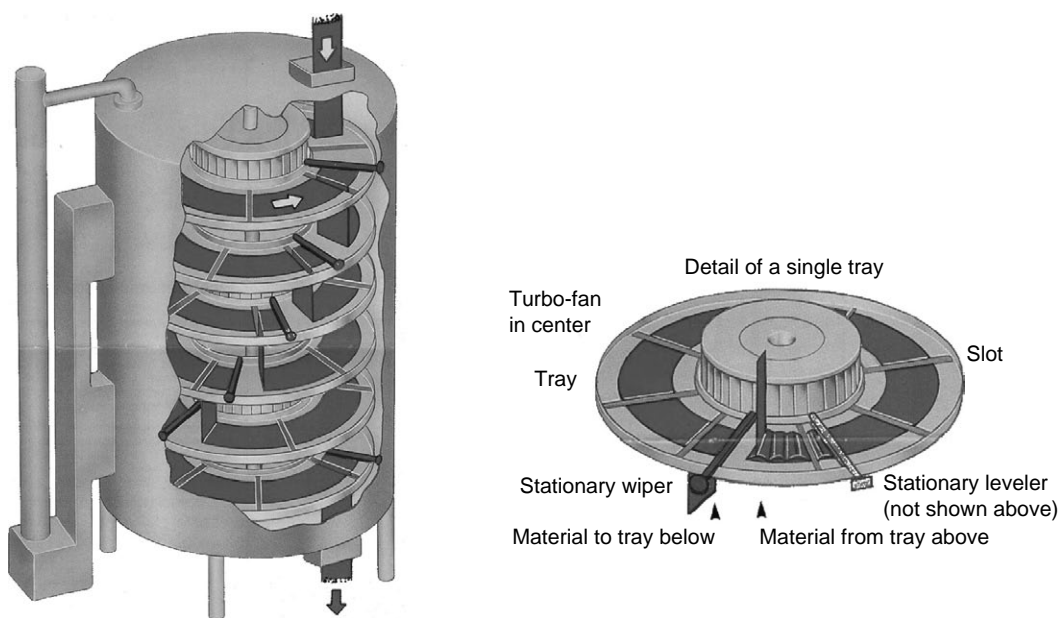


FIGURE 38.14 Vertical multi-tray dryer (courtesy of Wyssmont Co., Inc.).

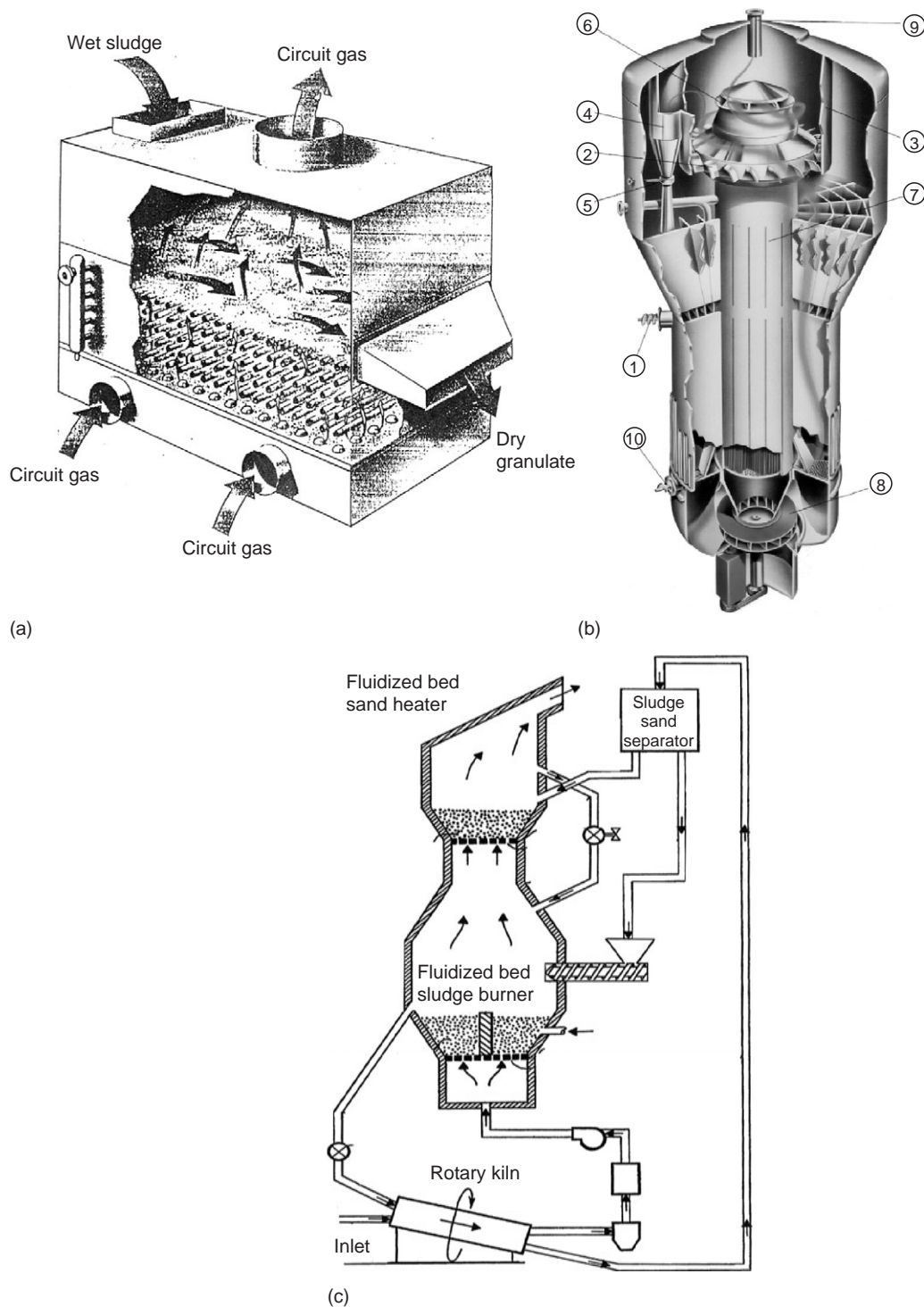


FIGURE 38.15 Fluidized bed dryers: (a) with heating tubes; (b) with superheated steam heating, 1—screw conveyor; 2—stationary blade; 3—cylinder; 4—side cyclone; 5—ejector; 6—stationary vanes; 7—superheater; 8—impeller; 9—top outlet; 10—screw conveyor (courtesy of Niro A/S); (c) rotary kiln dryer using sands heated in a fluidized bed burner (adapted from Isheim, M.C., US Patent 4,248,164, 1981).

incinerator, the dewatered sewage sludge is distributed over a hot fluidized sand bed at about 800°C and subsequently reduced to microsize particles as a result of rapid evaporation and combustion [41].

Multistage Carver-Greenfield evaporation of water from sludge [74] is found to be very energy efficient in addition to offering ease in handling of liquids until the sludge is dried [7]. Impulse drying of sludge was investigated and found to be able to increase the solid content of belt-pressed sludge significantly when the operating temperature exceeded 200°C. This is considered as an inexpensive energy-efficient technology because part of the water is expressed out of the sludge in the liquid form [75]. This investigation has led to two very recent patents under the name of “Apparatus for multinip impulse drying” [76,77].

38.5.6 RELATED ISSUES

The viruses and pathogens contained in sludge must be handled properly. Thermal sterilization is a good by-product of thermal drying. The US EPA requirement of 70°C for 2 h is equivalent to 90°C for 10 to 15 min, which can be easily satisfied by indirect dryers. The operating temperature of direct dryers can usually go above a few hundred degrees. Therefore, thermally dried sludge can be safely used for any purpose. Odor is another concern. In direct drying, exhaust gases may

be introduced to the incinerator or burner. In indirect systems, the limited amount of gas generated can either be burned or biofiltered. In direct drying with high solid contents at high temperatures, prevention of the explosion of the fine powders must be seriously considered. Figure 38.16 shows the typical components seen in a sludge dewatering and drying plant.

In spite of the known nutrients and carbon sources it contains, sludge is presently considered of low commercial value because of its heavy metal contaminants. The cost of sludge dewatering ranges from about US\$34 to 57/t of dry solids at respectively 150 and 60 t/day of dry solids using a high throughput centrifuge. In the case of belt filter presses, the cost for a similar dewatering capacity would be US \$40 and US \$64, respectively [7]. The capital investment for sludge drying is roughly US \$1.8 million/t/h evaporation rate. Variations can be found from the low-cost flash drying systems to the highly sophisticated expensive closed-loop systems. The operating cost depends on the grade, source, and cost of the fuels used for heat exchanger. The final moisture content requirement is a key parameter that influences the capital cost. The predicted range is US\$65 to 80/t of dry solids [55].

Utilizing the waste heat from the burning of sludge is considered to be very important. For example, flue gas drying preceding incineration can

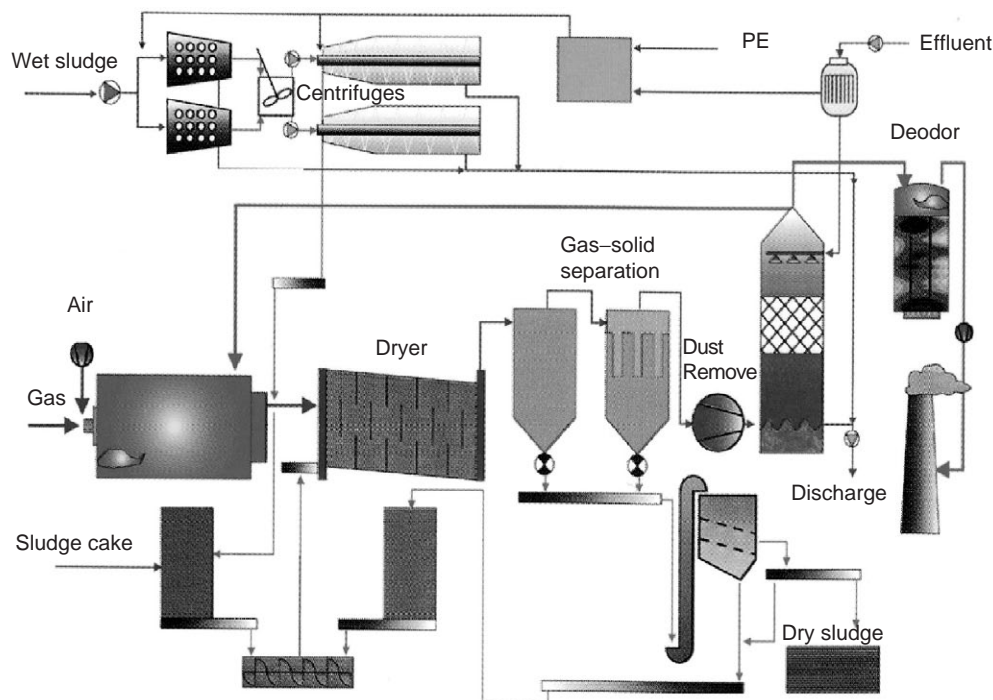


FIGURE 38.16 Typical components seen in a sludge drying plant using a direct dryer. (Courtesy of Vadeb.)

save as much as 60% on the cost compared with direct incineration in large sludge plants [78]. Similarly, mechanical compression of the vapors generated from indirect dryers can improve the energy efficiency considerably [79]. Cogeneration-sludge drying units might be an economically attractive option to consider in large wastewater treatment plants [80].

38.6 CONCLUSION

Wet sludge from wastewater treatment plants has greater than 95% moisture content. Proper utilization or disposal of this sludge has become a global issue because of increasingly stringent environmental regulations. The removal of water from wet sludge is an important step in reducing sludge volume and transportation costs and in increasing the heat value of sludge. Mechanical dewatering is usually the first treatment step after physical, chemical, biological, or thermal conditioning. There are several methods of sludge dewatering such as vacuum filters, belt presses, and centrifuges. Mechanical dewatering can be enhanced by acoustic and/or electro-osmotic effects. Solar drying is a natural choice if the environmental conditions permit.

Thermal drying can be accomplished in direct, indirect, or combined drying systems to produce a higher final solid content. There are numerous dryers available to dry sludge to a solid content of around 55 to 95%. The selection of a particular dryer depends on the initial and final moisture contents, the properties of the product, the ability to operate and maintain the system, and the nature and capacity of the sludge to handle. The cost of sludge dewatering depends on the process selected, with US\$34 to 40/t of dry solids for centrifuge and about US\$60/t of dry solids for a belt filter press. The drying cost varies among different technologies chosen with a range of US\$65 to 80/t of dry solids. Proper use of the waste heat generated from the vent gas can create significant savings in the drying cost.

Sludge drying is not an isolated issue. It has to be addressed along with other economical, environmental, and safety concerns. There may be no obvious and immediate financial incentives from investment in a sludge treatment process, but over the past decade the world has witnessed a significant increase in public investment in environmental protection including for sludge processing. Such a trend will be followed worldwide in the future. Despite being many types of dryers available, the market demands innovative drying technologies with higher thermal efficiencies, lower emissions, less operator involvement, cheaper capital costs, and better final products.

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39 Drying of Biotechnological Products

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39.1 INTRODUCTION

Biotechnology is the action aiming at producing useful products for various branches of the economy by means of biological components and microorganisms, viruses, animal and vegetable cells, as well as extracellular substances found within tissues. The growing scope of these activities includes production of a biological system, a producer strain, using the recombination technique and cell engineering. As a result of processes taking place in the presence of microorganisms, materials of various forms are produced, such as microorganisms similar to the inlet materials, e.g., yeast and bacteria. The product may be a substance with a complex chemical structure in the form of a high-molecular polymer or organic compound (e.g., antibiotics, vitamins, and organic acids).

Biotechnology includes several basic processes, the most important of which is the fermentation process (Figure 39.1). In this process the fundamental transformations of reagents take place in the presence of microorganisms. This is a complex process of

biochemical transformations depending on many microbiological, chemical, physical, and mechanical factors. These factors include microorganism cultures and strains, types and amounts of additives, pH of the medium, temperature, mixing intensity, and so on.

The other process, which is of equal importance, is preservation of the product of microbiological transformation. The preservation process depends, among other things, on product applicability, the way it is utilized, storage possibility, and conditions. Sometimes the product can be used jointly with by-products and unreacted residues of the culture medium and nutrients. Then, it is a raw material for microbiological concentrates (as a result of filtration, concentration, or drying). When the product is used as a purely microbiological material, it should be separated from the fermentation medium, impurities, and residues of reagents, and then the purified substance should be concentrated and preserved by removal of solvent (water) through freezing or drying.

The aim of this chapter is to present methods of preservation of biotechnological products by drying

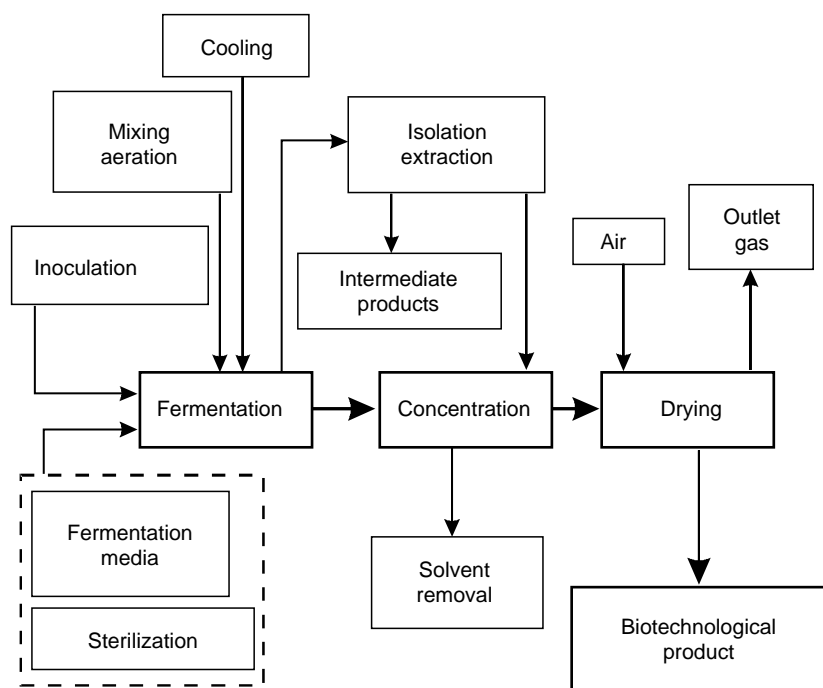


FIGURE 39.1 Schematic diagram of the biotechnological process.

while maintaining required product quality. In the first part, a biotechnological product is presented as an object of drying. Particular attention is given to its sensitivity and qualitative changes due to dewatering and elevated temperature. In the second part, qualitative changes of the product are considered in the mathematical description of the drying process, its modeling, and optimization of process parameters. Finally, many known drying methods are characterized with reference to their advantages and disadvantages as applied to biotechnological products.

This chapter has been updated from the previous version [1]. However, its structure has been preserved and the most important trends and procedures applied in drying of biotechnology products are emphasized. The chapter is supplemented with the results of the latest research, and most recent references are quoted. Worth mentioning are the monograph *Thermal Processing of Bio-materials* [2] and relevant parts of *Advanced Drying Technologies* [3] that have an applicative character. It is also worth referring the monograph on bioproduct granulation [4] that presents problems related to thermal impacts on a material and drying of bioproducts. From among many generally known source materials it is worth to mention the conference proceedings concerning preservation, concentration, and drying of foodstuffs [5–8] and the studies on the role of water in food and biological products [9–12]. Many monographs on drying of biotechnology products were published in Russian

[13–21]. There are still up-to-date general publications presenting the analysis of specific conditions of drying and preservation of biotechnology products used in many European countries [22–30], and also valuable publications that give an account of experimental results [31–35].

39.2 BIOTECHNOLOGICAL PRODUCTS AS MATERIAL BEING DRIED

A typical suspension of liquid fermentation culture is an aqueous mixture of microorganisms or biopolymers, unreacted residues of nutrients, by-products, process-controlling additives, and so on, with dry mass content amounting to several percent. In the process of concentration and drying of these materials, attention cannot be paid only to the removal of water, which is the medium and solvent of a substance, but also to the removal of water, which is a constructional element of the product particle. The presence or absence of such water is of fundamental importance for biotechnological product storage life, its activity, and applicability. In general, dried materials have been classified as drying objects according to their structure and type of water binding [18,20]. A variety of biotechnological products, from the point of view of both their nature and structure, do not allow precise specification of the form of material-moisture binding.

In general, it is assumed that these materials belong to colloidal-capillary-porous substances in which all forms of material-moisture bonds are possible. Apart from water solvent, there are also free water and bound water. A significant part of water occurring in the cells of microorganisms is free water, while that in the macromolecules of biopolymers and products of microbiological transformations is mostly bound water (Figure 39.2). The biosynthesis products are usually hydrophilic colloids that can hold up a significant amount of bound water.

In the microorganism structure, e.g., a typical bacterial cell, there are many important elements (e.g., nucleus, cytoplasm, cell membrane) built of a large number of various inorganic and organic compounds with molecular weights that range from several hundreds to thousands of daltons. Among all components of a live cell, water prevails, amounting to 70 to 98%. In microorganisms only about 15 to 18% bound water was found. A major part of the water present in the cell is free water that is a medium for various reactions and a solvent of many sub-

stances. Fluctuations in water level can cause disturbances in biochemical processes that take place in the cell because many reactions, including those catalyzed by enzymes, depend on hydration of a cell as a whole and of cell proteins. Interactions typical for drying that cause changes in moisture content and temperature of a product include many transformations affecting its quality. Examples of basic changes are given in Table 39.1. Four types of changes are distinguished: (a) biochemical (microbiological); (b) enzymatic; (c) chemical; and (d) physical.

Biochemical changes characteristic of yeast or bacteria are strictly connected with the loss of water in the cells and in their individual structural elements. Enzymatic changes include mainly the changes of activity caused by structural biopolymer decomposition. Chemical changes usually result in a decrease of nutritive values of biotechnological products and in a formation of substances noxious to the environment. The effect of these changes can easily be observed. Biochemical and chemical changes are often revealed as physical changes: the biotechnological product loses its solubility or water-binding capacity; it also loses aromatic compounds due to decomposition or high volatility. Decoloration of the product is often observed.

Another set of quality changes is shown in Table 39.2, which presents examples of such changes in particular groups of biotechnological products and their main components during drying. In general, it may be stated that the basic degrading mechanism is the denaturation of protein, which is the main component of both live microorganisms and products of a microbiological transformation.

Rockland and Beuchat [9] presented a diagram illustrating the effect of moisture content (exactly, water activity) on the reactions of compounds and the preservation of the activity of a biotechnological substance or foodstuff (cf. Figure 39.3).

The level of water removal from the reaction medium depends on the nature of the substance and on the applicability of the product, i.e., whether after drying the substance remains biologically active, is in anabiosis, or becomes an inactive organism with determined structure and biochemical composition. An example can be baker's yeast (with a final moisture content of about 10% by weight—active yeast, after rehydration being able to grow and undergo biochemical transformations) and fodder yeast (with a final moisture content of about 5% by weight—dead microorganisms with no microbiological activity after rehydration). An effect of moisture content on the percentage of living cells is presented in Figure 39.4 [20].

Temperature and the duration of thermal processing are the other factors on which microbiological

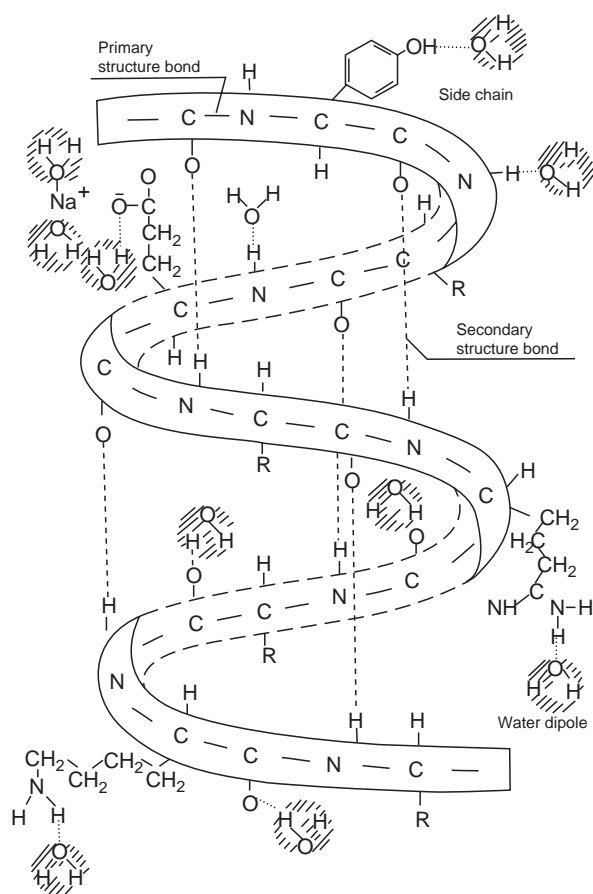


FIGURE 39.2 Schematic diagram of water particle addition to polar protein groups.

TABLE 39.1
Main Changes of Biotechnological Product Properties during Drying

Biochemical (microbiological)	Enzymatic	Chemical	Physical
Atrophy of microorganisms (cells)	Loss of activity	Decrease of nutritive values and activity	Solubility Rehydration Shrinkage Loss of aroma
Yeast Bacteria Molds	Enzymes Vitamins	Protein Carbohydrates Fats Antibiotics Amino acids	Every biotechnological product

properties of preserved biotechnological products depend. In general, a lower process temperature enables a longer processing time. However, as follows from drying theory, under strictly controlled heat conditions it is possible to use elevated temperatures of the drying agent during drying of thermolabile materials. The effect of temperature and time on the percentage of live microorganisms is shown in [Figure 39.5](#) [20].

To quantitatively analyze the biotechnological product resistance and durability, changes of one or several quality factors considered most important are assumed. In general, a qualitative change can be determined by the index of quality degradation A ; the indices written as A_X or A_T refer to quality changes resulting from the changes in moisture content and temperature, respectively. They are the function of time and reflect the number (concentration) of living

TABLE 39.2
Examples of Degradation Processes of Biotechnological Product Components during Thermal Drying

Product	Type of Reaction	Degradation Processes	Result
Live microorganisms	Microbiological changes	Destruction of cell membranes	Denaturation of protein Death of cells
Lipids	Enzymatic reactions	Peroxidation of lipids (discoloration of the product)	Reaction with other components (including proteins and vitamins)
Proteins	Enzymatic and chemical reactions	Total destruction of amino acids Derivation of some individual amino acids Cross-linking reaction between amino acids	Denaturation of proteins and enzymes Partial denaturation, loss of nutritive value Change of protein functionality
Polymer carbohydrates	Chemical reactions	Gelatination of starch Hydrolysis	Enzyme reaction Improved digestibility and energy utilization Fragmentation of molecule
Vitamins	Chemical reactions	Derivation of some amino acids	Partial inactivation
Simple sugars	Physical changes	Caramelization (Maillard-Browning reaction) Melting	Loss of color and flavor

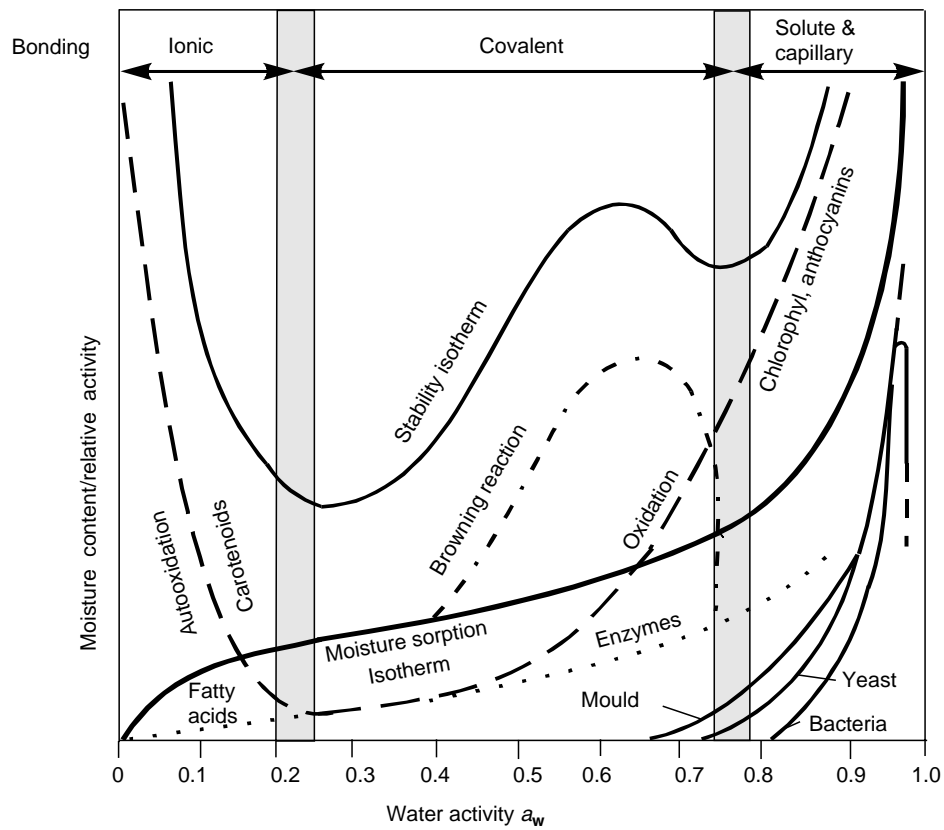


FIGURE 39.3 Water activity–stability diagram. (From L.B. Rockland and L.R. Beuchat (Eds.), *Water Activity: Theory and Applications to Food*, Marcel Dekker, New York, 1987. With permission.)

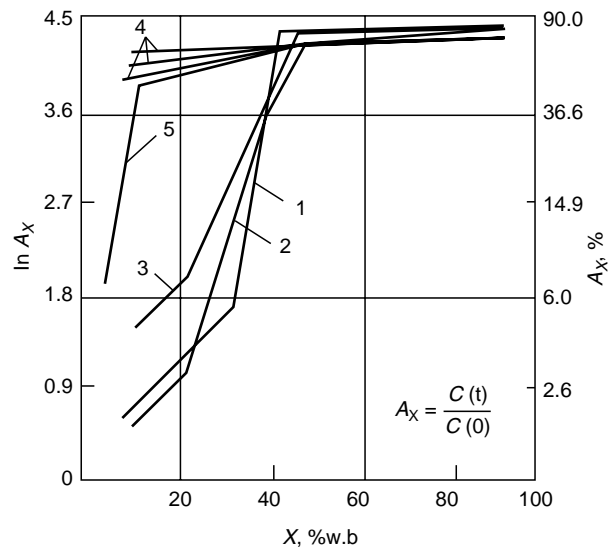


FIGURE 39.4 The effect of moisture content on the percentage of living cells of microorganisms: (1) *Lactobacillus plantarum*; (2) *Rhizobium pisum*; (3) *Beauveria bassiana*; (4) *Rhodotorula* sp., *Candida* sp.; and (5) *Saccharomyces* sp. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow (1987) (in Russian).)

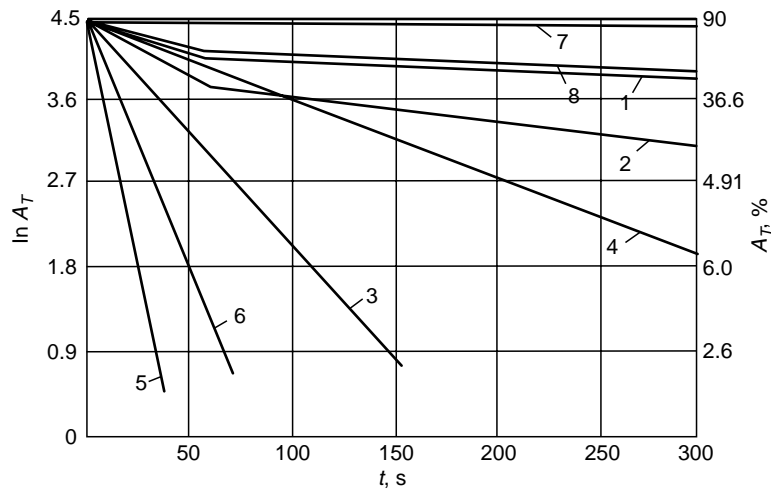


FIGURE 39.5 The effect of temperature and time on the percentage of living microorganisms and active biopolymers: (1, 2, 3) *Bacillus thuringiensis* at 90, 120, and 170°C, respectively; (4, 5) *Beauveria bassiana* at 50 and 60°C, respectively; (6) *Lactobacillus plantarum* at 55°C; and (7, 8) lysine at 150 and 190°C, respectively. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow (1987) (in Russian).)

microorganisms $C(t)$ or the activity of a product $a(t)$ as related to the values characterizing the beginning of the process $C(0)$ or $a(0)$:

$$A_X \text{ or } A_T = \frac{C(t)}{C(0)} \cdot 100, \% \quad (39.1)$$

or

$$A_X \text{ or } A_T = \frac{a(t)}{a(0)} \cdot 100, \%$$

When characterizing the quality of a product of microbiological transformations (and food products as well) a large number of properties are taken into account. In Figure 39.4 and Figure 39.5 only one property is given, namely, the percentage of biologically active biomass cells as a function of moisture content A_X , temperature A_T , and process time t . Literature on product quality refers to different properties and transformations that affect thermal preservation and storage of biotechnological and food products [6–9].

The biotechnological products can be classified as drying objects using special criteria (e.g., resistance to elevated temperature or susceptibility to drying). Tutova and Kuts [20] applied the above criteria and proposed the classification of biotechnological products into two groups:

1. Materials nonresistant thermally, microorganisms of high death rate due to thermal treatment; they are characterized by a relatively

high value of critical, boundary moisture content. So, these are thermo- and xerolabile substances. The assumed boundary values are maximum at temperature, about 60°C and moisture content, about 35 to 40% by weight. In this group, bacterial cultures, enzymes, viruses, yeast, and fungi are included.

2. Biotechnological products of high thermal resistance, with a thermal inactivation rate that is low and critical moisture content reaches several percent. So, these are thermo- and xerostable substances. The assumed boundary values are temperature, about 150°C and moisture content, 5 to 7% by weight. This group contains the products of microbiological synthesis and spores (e.g., amino acids, antibiotics, and selected bacterial strains).

Another classification is proposed, taking as a basis the role of water present in the biotechnological product [31]. Here, two groups are also distinguished:

1. Biotechnological products in which water is one of the decisive elements for the life and activity of microorganisms (bacteria, yeast, and molds) or products of microbiological transformation (enzymes). These substances are characterized by high susceptibility to drying and their qualitative parameters are subjected to significant changes during drying.
2. Biotechnological products in which water is a solvent, a medium for microbiological changes, and not a structural element of biopolymers

that decides their properties (antibiotics, amino acids, and vitamins). These materials are characterized by high resistance to drying and their qualitative parameters undergo only slight changes during drying.

One more classification of biotechnological products can be made, taking as a criterion the applicability of the dry product. The same biotechnological product can be preserved maintaining its microbiological activity and biological structure (e.g., active yeast for promoting fermentation processes) or be devoid of the properties of a live microorganism with its chemical composition maintained (e.g., inactive yeast with high protein content used as fodder yeast).

39.3 DEGRADATION PROCESSES IN THE SCHEME OF DRYING MODELING

In a general scheme of the modeling of drying, the aspect concerning product quality is neglected. In the case of modeling of drying biotechnological products it is necessary to consider the kinetics of qualitative changes of products and relate it to the drying kinetics [29,31–42]. Thus, for one or several selected qualitative parameters being most important for a product, the process of qualitative degradation should be described using mathematical expressions. As stated above, in such an expression for a given index, three most important drying parameters should be considered: (a) moisture content; (b) material temperature; and (c) process duration.

In majority of qualitative changes it is assumed that their kinetics corresponds to the first-order reaction and can be described by the following equation:

$$\frac{dA}{dt} = r_d = -k_d A \quad (39.2)$$

where A is the numerical value of the characteristic index of degradation described by this relationship (e.g., the number of living microorganisms, activity of antibiotics, enzymes). The reaction rate constant k_d , as a function of temperature, is determined by the Arrhenius equation:

$$k_d = f(T) = k_{d,T} = k_{\infty} \exp\left(-\frac{E_a}{R\theta}\right) \quad (39.3)$$

The frequency coefficient k_{∞} (s^{-1}) and activation energy of E_a (J/mol) are characteristic values for a given substance and depend on the temperature, moisture content, and reaction time.

Tutova and Kuts [20] presented a diagram of changes in the reaction rate constant for thermal inactivation $k_{d,T}$ depending on the temperature for several types of biotechnology products (Figure 39.6). As the degradation index, the death rate of microorganisms has been assumed. The first group of lines [1] refer to vegetative forms of microorganisms drying at a relatively low temperature (45 to 50°C). The inactivation temperature of these cultures is in the narrow range (10 to 15K) with $k_{d,T}$ changing 17 times. Lines [2–4] refer to spores and biosynthesis products. They are characterized by higher thermostability and are resistant to the temperature above 100°C. The kinetic constant $k_{d,T}$ changes 6 to 8 times in a wider range of temperatures (60 to 80 K).

On the basis of experimental data, using the calculation method of chemical reaction kinetics, the kinetic parameters of thermal inactivation of many substances were established (Table 39.3) [20]. However, the authors make reference to a high discrepancy of experimental and calculated data, reaching 20 to 30%. Similar kinetic parameters were determined as a result of investigations of selected biotechnological products [43–46].

The qualitative changes induced by a decrease in moisture content in a biotechnological product

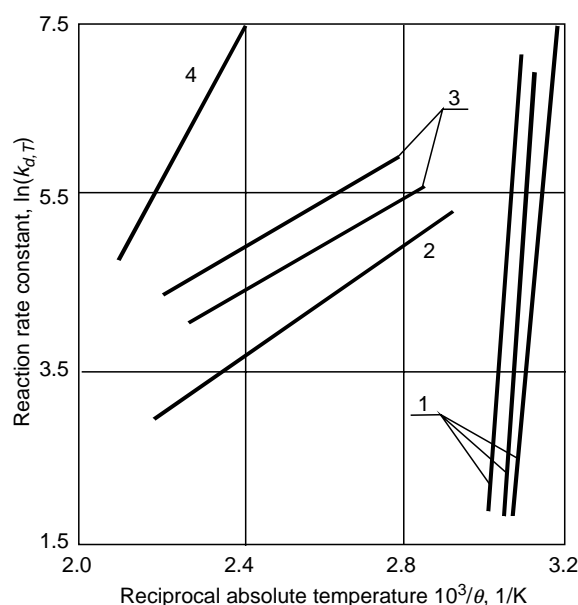


FIGURE 39.6 Dependence of constant reaction rate of thermal inactivation $k_{d,T}$ on temperature θ : (1) nonsporulating microorganisms; (2) bacterial spores; (3) antibiotics (bacitracin and grisin); and (4) lysine. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow (1987) (in Russian).)

TABLE 39.3
Kinetic Parameters of Thermal Inactivation

Substance (temperature range)	E_a (kJ/mol)	k_∞ (s ⁻¹)	Heating Time t (s)
Nonspore bacteria (40–55°C)			
<i>Beauveria bassiana</i>	227	10^{34}	
<i>Rhizobium pisum</i>	248	10^{38}	
<i>Azotobacter chroococcum</i>	341	$10^{52.5}$	
<i>Lactobacillus plantarum</i>	428	$7 \cdot 10^{66}$	
<i>Lactobacillus acidophilus</i>	142	10^{21}	
Fungi			
<i>Penicillium, Trichoderma</i>	299	10^{48}	≤ 60
Enzymes			
Transeliminase	111	$10^{15.5}$	
Polygalaturonase	251	10^{37}	
Spore bacteria (90–190°C)			
<i>Bacillus thuringiensis</i>	33	$2.7 \cdot 10^2$	
Antibiotics (90–200°C)			
Biomycin	64.5	$1.6 \cdot 10^6$	
	58	$1.2 \cdot 10^5$	> 60
Grisin fodder preparation	45	$7.9 \cdot 10^4$	≤ 60
	27	1.75	> 60
Bacitracin	68	$1.5 \cdot 10^6$	≤ 60
	62	$2 \cdot 10^5$	> 60
Lysine (90–200°C)	107	10^{10}	≤ 60
	30	1.9	> 60

Source: From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).

(dehydration degradation) can be described using the reaction rate constant $k_{d,X}$ depending mainly on biomass moisture content X [20]:

$$k_d = f(X) = k_{d,X} = k'_\infty X^{-n} \quad (39.4)$$

where k'_∞ and n are the parameters characteristic for dried material.

Luyben et al. made simulation calculations of the drying process assuming the degradation index to be the change of activity of three enzymes: alkaline phosphatase in skim milk, lipase in rye, and catalase in wheat and catalase from spinach in a buffer solution [33]. They transformed general Equation 39.2 of qualitative changes (i.e., the activity change) to the form:

$$a = a_0 \exp \left(- \int_0^t k_d dt \right) \quad (39.5)$$

The reaction rate constant k_d was calculated according to the Arrhenius equation, Equation 39.3. The activation energy E_a and frequency coefficient k_∞

were made dependent on moisture content X according to the empirical equations:

$$E_a = E_{a,\infty} + (E_{a,0} - E_{a,\infty}) \exp(-pX) \quad (39.6)$$

$$\ln k_\infty = \ln k_{\infty,\infty} + (\ln k_{\infty,0} - \ln k_{\infty,\infty}) \cdot \exp(-pX) \quad (39.7)$$

The numerical value of constant p was given for each enzyme on the basis of experimental data: for catalase, $p = 3.699$; for lipase, $p = 4.880$; and for alkaline phosphatase, $p = 11.366$. The effect of moisture content X on E_a and k_∞ is illustrated in Figure 39.7 and Figure 39.8. From these graphs the values of k_∞ (s⁻¹) and E_a (J/mol) can be read out for a bone-dry product (subscript 0) and for a product with moisture content $X = 1$ and more (subscript infinity). Figure 39.9 presents a change of reaction rate constant k_d of enzyme inactivation as a function of moisture content X calculated from the Arrhenius equation for temperature 60°C.

Zimmermann and Bauer [34,35] investigated the drying of baker's yeast (granulated yeast in a fluidized bed) and developed a mathematical model in which thermal inactivation of a product was based on the

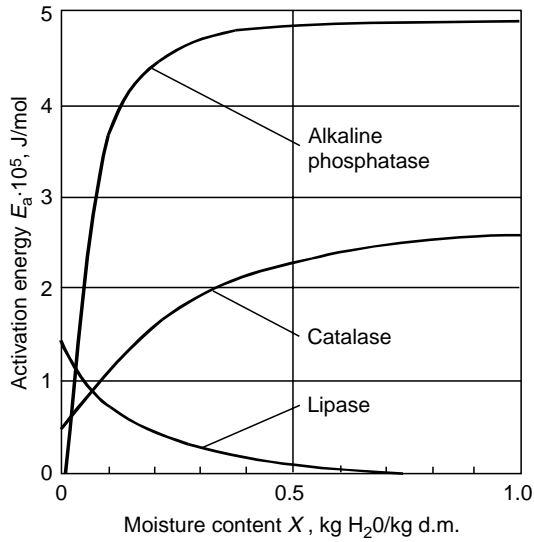


FIGURE 39.7 Activation energy E_a of enzyme inactivation as a function of material moisture content X . (Adapted From K.Ch.A.M. Luyben, J.K. Liou, and S. Bruin, *Enzyme Degradation During Drying*, *Biotechnology and Bioengineering*, 24, 533–552 (1982).)

first-order reaction equation and the Arrhenius equation. Taking into account the form of dried material (pellets of about 0.8 mm or 0.5 mm diameter) and the theory of shrinking during drying, the general

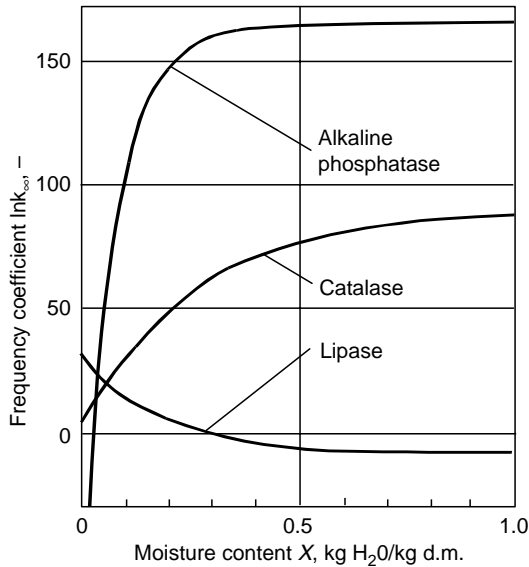


FIGURE 39.8 Frequency coefficient k_∞ of enzyme inactivation as a function of material moisture content X . (Adapted From K.Ch.A.M. Luyben, J.K. Liou, and S. Bruin, *Enzyme Degradation During Drying*, *Biotechnology and Bioengineering*, 24, 533–552 (1982).)

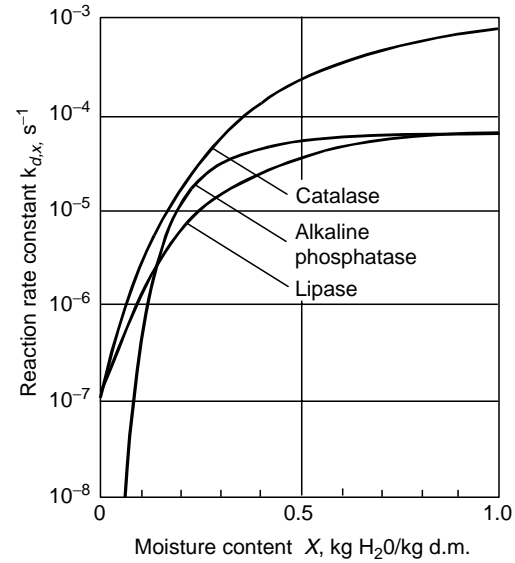


FIGURE 39.9 Specific reaction rate $k_{d,X}$ of enzyme inactivation at 60°C as a function of moisture content X . (From K. Ch. A.M. Luyben, J.K. Liou, and S. Bruin, *Enzyme Degradation During Drying*, *Biotechnology and Bioengineering*, 24, 533–552 (1982), Wiley, New York. With permission.)

inactivation of enzymes was considered to be that of inactivation in the coating layer of pellets at the hygroscopic moisture content X_{hygr} , and medium temperature T_m , and inactivation of enzymes in a wet core of material with critical moisture content X_{crit} and surface temperature T_S . The above-mentioned relations are described by the following equation:

$$\frac{da}{dt} = - \left[k_d(X_{crit}, T_S) a \frac{r^2}{R^2} + k_d(X_{hygr}, T_m) a \left(1 - \frac{r^2}{R^2} \right) \right] \quad (39.8)$$

where r is the radius of the wet core (m) and R is the pellet radius (m).

Kerkhof and Schoeber [36], modeling droplet drying in spray dryers (SDs), used earlier works by Labuza (1972), Verhey (1972), and Ball and Olson (1957). In the case of the first-order reaction of quality degradation of component i , the following equation for a determined volume of the reference system is employed:

$$\frac{\partial C_i}{\partial t} = -k_d C_i - \frac{1}{r^2} \frac{\partial^2}{\partial r^2} (r^2 w_i) \quad (39.9)$$

where w_i is the density of mass stream of component i being degraded ($\text{kg/m}^2\text{s}$) and r is the droplet radius (m).

The effect of temperature on the reaction rate constant $k_{d,T}$ can be expressed by the general formula:

$$k_{d,T} = k_{\infty} \exp(zT) \quad (39.10)$$

which, when used in the calculation of phosphatase inactivation in skim milk, gave better agreement with experimental data than the Arrhenius equation [36]. In model calculations, Kerkhof and Schoeber used the following empirical relation valid for phosphatase inactivation in skim milk and for aqueous maltose solution:

$$k_d = \exp(-34.9 - 1.39/C_w + 0.5183T) \quad (39.11)$$

where C_w is the water concentration inside the droplet (kg/m^3).

Figure 39.10 presents changes of temperature and thermal degradation in the droplet of maltose–water system as a function of reduced time (t/R_0^2) calculated by Kerkhof and Schoeber [36]. It follows from the graph that at temperature 100°C the complete inactivation takes place in a very short time; at temperature 80°C the reaction begins at a slower rate. In both cases the reaction starts when the droplet temperature is about 70°C .

Karel [37] analyzed several recent works on qualitative changes of dehydrated foodstuffs and biotechnological materials. He stressed that when optimizing the process from the point of view of product quality

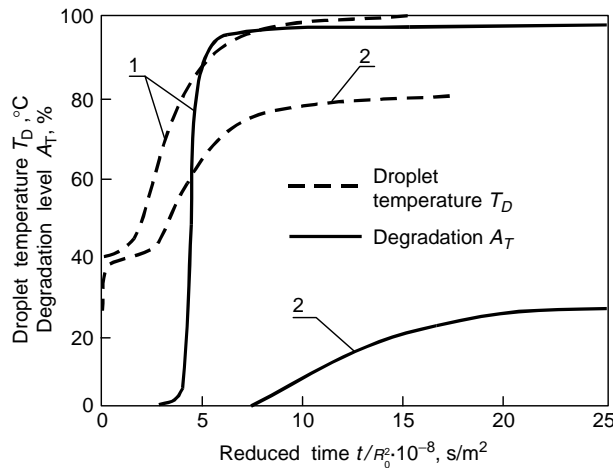


FIGURE 39.10 Droplet temperature T and thermal degradation A_T in maltose–water system vs. reduced time t/R_0^2 for two air temperatures: (1) to 100°C , and (2) to 80°C . (From P.J.A.M. Kerkhof, W.J.A.H. Schoeber, Theoretical Modeling of the Drying Behavior of Droplets in Spray Dryers, in *Advances in Preconcentration and Dehydration of Foods*, A. Spicer (Ed.), Elsevier Applied Science, London, pp. 349–397 (1974). With permission.)

usually the interaction of several qualitative parameters takes place. Their change in the final product depends not only on material temperature and moisture content, but also on changes occurring in other parameters. He stated that numerous processes that cause these changes during water removal are dependent on the “mobility” of components of the material being dehydrated. He stressed that qualitative changes and the effect of process parameters must be analyzed as a function of the time of interaction, which is often neglected in other studies. This statement can be written in the general form of the function of a constant reaction rate of qualitative changes:

$$k_d = f(T(t), X(t)) \quad (39.12)$$

A synthesis of the mathematical modeling and optimization of the process was presented by Strumiłło et al. [29,32] and Kamiński and Strumiłło [39]. A natural starting point for the analysis is that the rate constant k_d depends on the history of the drying process, which should be known if the description of the degradation processes is to be complete. Thus, the following equations are obtained experimentally:

$$T = g_1(t) \quad (39.13)$$

$$X = g_2(t) \quad (39.14)$$

By correcting k_d , the constants in Equation 39.15 are searched for with the use of experimental data:

$$k_d = k_{\infty} f(T, X) \cdot \exp(-E_a/R\theta) \quad (39.15)$$

where k_{∞} and E_a are the experimental constants. To obtain such a relation, a dynamic approximation of the experimental data is required.

For example, on the basis of experimental data presented in Ref. [33], the following correlations of constant k_d of the degradation indices of baker’s yeast *Saccharomyces cerevisiae* were obtained:

Alcohol dehydrogenase (ADH) enzyme activity:

$$k_d = (1 - 0.2555X + 9.529 \cdot 10^{-2}X^2 + 1.859 \cdot 10^{-3}XT) \times \exp\left(-\frac{310.4 + 339.8X}{T + 273.15}\right) \quad (39.16)$$

CO_2 production:

$$k_d = [1.0023 + \exp(2.269 \cdot 10^{-4}(T - 20) + 9.124 \cdot 10^{-4}(T - 20)^2)] \times \exp\left(-\frac{515.27 + 240.72X}{T + 273.15}\right) \quad (39.17)$$

O₂ demand:

$$k_d = [1.0045 + \exp(7.51 \cdot 10^{-3}(T - 20) + 5.276 \cdot 10^{-4}(T - 20)^2)] \times \exp\left(-\frac{525.35 + 200.25X}{T + 273.15}\right) \quad (39.18)$$

In Equation 39.16 through Equation 39.18 material temperature and moisture content are the functions of time.

In general, in model and design calculations optimum process parameters of drying are determined. Taking into account the quality changes, usually several quality indices are considered and thus some hierarchy of the index system can be assumed or one global index A_g comprising all indices of the system may be created [38]. The global index can be written as:

$$\ln(A_g) = \sum_{i=1}^n \ln\left(\frac{a_i}{a_{i0}}\right)^{\lambda_i} = - \int_0^t \sum_{i=1}^n k_{di}(\dots) dt \quad (39.19)$$

where $\lambda_i > 0$, $i = 1, \dots, n$ is the weight system for the assumed system of quality indices a_i [39].

39.4 EXAMPLES OF DRYING METHODS AND EQUIPMENT

A wide choice of biotechnological products indicates that it is possible to use various methods to dry them. However, taking into account thermo- and xerolability and a risk of many qualitative changes in these products, some process conditions should be strictly observed (e.g., temperature and moisture content of a product should be controlled, rapid changes of pressure should be eliminated, and mechanical interactions should be avoided as they may damage the cells of microorganisms). In addition, many processes must be carried out under sterile conditions, using sterile drying agents. In general, the following solutions, which take into account the properties of biotechnological products, can be recommended:

1. Low-intensive conditions
2. Vacuum or freeze drying
3. Application of multistage systems
4. Special techniques based on the decrease of initial material moisture content

Knowing general principles of particular drying methods, we shall discuss their advantages and disadvantages in application to biotechnological products, in particular the one that may endanger high product quality.

39.4.1 SPRAY DRYING

The spray-drying method is widely used, in particular for liquid suspensions of substances resistant to high temperatures, such as biopolymers and products of microbiological transformations. On the other hand, spray drying can also be applied to thermolabile substances provided the sprayed material is highly dispersed and the drying agent has a high temperature (the drying process takes place immediately at a wet bulb temperature). Thus, the product quality is not deteriorated. The installation should operate faultlessly and the control and measuring equipment should be of high quality. Disadvantages of this drying method are scaling up of dryers under mild thermal conditions (at 150 to 200°C), expensive devices for liquid dispersion and dust removal from the product are needed, and instability of the process due to possible deposition of a product on chamber walls and dust-removing gas pipes. The product can often be deposited on the chamber walls because of changing properties of the material being sprayed; this is connected with different compositions of fermentation broth, changing conversion factors during the fermentation process, and varying amounts of additives. The application of special spray dryer constructions can prevent the disadvantageous effects.

An example of the special spray dryer construction to drying of a suspension of antibiotics, fodder yeast, amino acids, and enzymes is shown in Figure 39.11 [20]. Inside the dryer chamber a sweeper rotates along the wall. The cleaning is made by air of proper parameters, such as pressure and temperature, which is introduced by nozzles installed on the sweeper. The air, or an inert gas, prevents the material from sticking to the wall and cools down to a thermolabile substance. A similar result can be achieved in a cocurrent spray dryer when a specially designed air distributor producing an air cushion along the chamber wall is installed [20]. For example, the dryer presented in Figure 39.11 for drying of fodder lysine concentrate, with evaporation capacity of 1000 kg/h, is characterized by the following parameters:

Diameter of the cylindrical section is 5.4 m.
Height of the drying chamber is 9.0 m.
Cone angle is 50°.
Diameter of the spraying disk is 0.25 m.
Frequency of disk rotations is 8,000 to 11,000 rpm.
Drying agent temperature at the inlet is 250 to 290°C and at the outlet is 80 to 125°C.
Fan output is 16,000 m³/h.

For deep drying of a product in spray dryers, it is required to use high drying temperatures. In the case

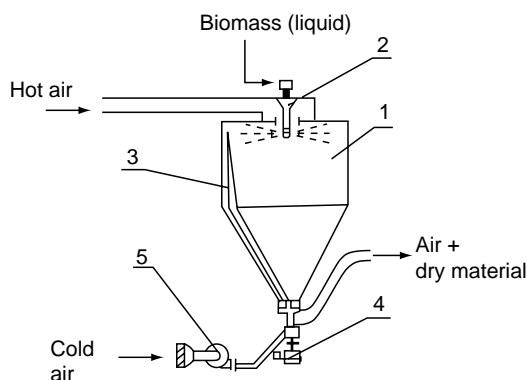


FIGURE 39.11 Scheme of spray dryer with chamber sweeper: (1) drying chamber; (2) biomass distributor; (3) rotating sweeper; (4) sweeper drive; and (5) blower. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

of biotechnological products, under such conditions degradation of product quality may take place easily. Therefore, to preserve the quality of biotechnological products, it is advisable to apply multistage dryers [47]. The first stage is a spray dryer in which material is predried under conditions ensuring a safe temperature and quick product removal from the drying chamber. The second and the third stages are fluidized or vibrofluidized bed dryers in which slow drying or cooling of the product takes place. In such a system a detailed control of product temperature and moisture content is possible in an appropriate residence time at the second or the third stage, which makes it possible to obtain a product of preserved biotechnological properties. Spray drying for drying of biotechnological products is a method still most frequently applied in practice and research [44,48–53].

39.4.2 JET-SPOUTED BED DRYING

For drying of suspensions at a low solid-phase concentration, a solution called a jet-spouted bed (JSB) is applied (Figure 39.12) [54–56]. This is a conical-cylindrical chamber containing inert material spouted by a high-velocity airstream. The material being dried is sprayed by nozzles onto an inert material. Among many advantages of the presence of inert material, usually in the form of cubes or spheres of diameter $5 \cdot 10^{-3}$ to $10 \cdot 10^{-3}$ m, two are worth mentioning:

1. A heated material surface causes an intensified drying.
2. An intensive motion of particles leads to a mechanical cleaning of the chamber and ensures high disintegration efficiency of dry particles.

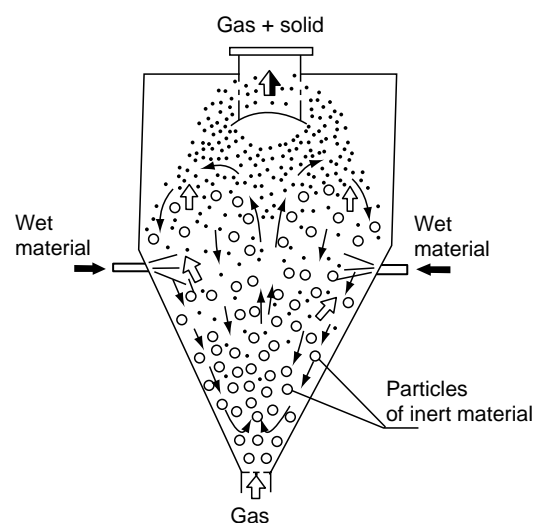


FIGURE 39.12 Scheme of jet-spouted bed dryer.

This method is useful in drying of thermally resistant products (e.g., amino acids and antibiotics). A JSB dryer is a competitive construction for a spray dryer, particularly for technologies of low and medium capacity, due to a higher volumetric evaporation rate and lower capital costs (a significant reduction of drying chamber volume). Table 39.4 presents a comparison of parameters and results of drying of a fodder antibiotic (Zn–bacitracin complex) in both dryer types [55]. A several times higher volumetric evaporation rate W_v is observed in the JSB dryer and about 20% higher heat utilization coefficient k is reported for a spray dryer.

TABLE 39.4
Comparison of Optimal Drying Parameters for Zn–Bacitracin in a Jet-Spouted Bed Dryer and Spray Dryer

Parameter	JSB Dryer	SD (lab)	SD (full scale)
T_{G0} , °C	170	200	248
T_{Gk} , °C	90.5	85	80
X_0 , %	80.2	83.4	83
X_k , %	2.64	6.55	5.24
W_v , kgH ₂ O/m ³ ·s	$4.37 \cdot 10^{-2}$	$0.26 \cdot 10^{-2}$	$0.14 \cdot 10^{-2}$
k	0.45	0.53	0.58
V_G , kg air/kgH ₂ O	36.5	25.4	18.9
d_p , m	$24.7 \cdot 10^{-6}$	$49.7 \cdot 10^{-6}$	$30.9 \cdot 10^{-6}$
a , units/mg	4.2	4.3	4.4
W_{H_2O} , kgH ₂ O/h	8	12	1200

Source: From A.S. Markowski, *Drying Technology*, 11(2), 369–387 (1993).

39.4.3 IMPINGEMENT STREAM DRYING

One of the latest drying methods for suspensions and solutions called, in general, impingement stream drying (ISD) is a method specially recommended for drying of liquid biotechnological products [57]. The method consists of multiple mutual collisions of drying agent streams flowing at high velocities into which fermentation liquid is sprayed by nozzles. Depending on a specific design solution, the following dryers are distinguished: (a) counterrotating countercurrent; (b) corotating countercurrent; and (c) coaxial with a moving impingement zone. An intensive evaporation and the motion of suspension and product particles ensure short residence time of the material in the drying zone, which has an effect on high quality of products. A schematic diagram of ISD in the counterrotating countercurrent version is presented in Figure 39.13.

39.4.4 DRYING IN A FLUIDIZED AND VIBROFLUIDIZED BED

The advantage of the fluidized and vibrofluidized system is that it is isothermal due to intensive mixing of the solid state, which prevents local overheating of the particles. It is possible to build multistage dryers and apply independent temperature control in particular sections. Thus, the product can be dried uniformly and changes in material moisture content can be controlled precisely in every zone of the dryer. This is particularly important for thermo- and xerolabile products. Advantageous also are high intensity of transfer processes, relatively simple dryer construction, and easy control of process parameters.

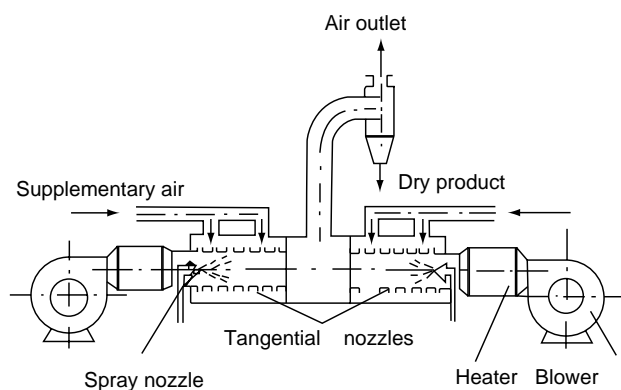


FIGURE 39.13 Impingement stream dryer for biotechnological products. (From T. Kudra and A.S. Mujumdar Impingement Stream Dryers for Particles and Pastes. *Drying Technology*, 7(2), 219–266 (1989).

Disadvantages of this drying method are the risk of mechanical damaging of cells (abrasion) or formation of material agglomerates and deposits on the walls, which disturb the uniform bed fluidization and cause local overheating of material and thermal inactivation of the product. A disadvantage of drying in a fluidized bed, particularly of a polydisperse biotechnological product, is the necessity of limiting the velocity of a fluidizing agent below the velocity of material particle entrainment, with a related risk of overheating a thermolabile material near the grid baffle of the dryer. This, however, can be overcome if a vibrofluidized bed is applied. During drying of granulated material there is a temperature drop in the granule cross section and thus there is a risk of overheating the surface layer and inactivation of this part of the product. Because process parameters must be controlled and, due to product quality changes, drying is carried out in batch or continuous fluidized bed multistage dryers with parameters adjusted in particular sections (heating and cooling) (Figure 39.14).

As already mentioned, before drying the biotechnological products are usually liquids or paste-like suspensions. Despite such a form they can be dried in fluidized or vibrofluidized beds. Three characteristic variants of process schemes are used: spraying of liquid onto the bed of inert material (sprayed fluidized bed) [59], which is similar to the JSB dryer presented above; use of a bed of recirculating product (Figure 39.15) [58]; and drying of granulated biotechnological product [34]. However, these methods have limited applicability.

The application of thermosensitive microbiological materials in product recirculation may influence a risk of thermal inactivation if part of the substance being dried stays too long in the heating zone during recirculation. In the case of spraying the material onto the bed of inert material, its particles are covered with a thin layer of wet material. With

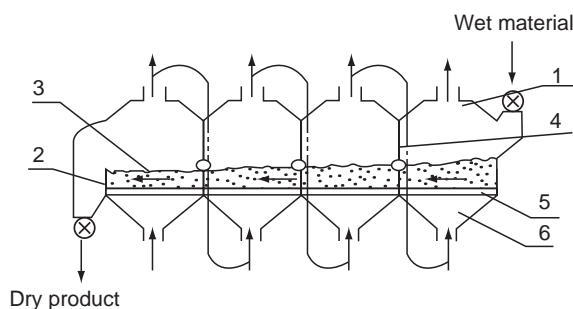


FIGURE 39.14 Scheme of the fluidized bed horizontal multisectional dryer: (1) drying chambers; (2) threshold; (3) bed; (4) baffles; (5) grid; and (6) air collectors.

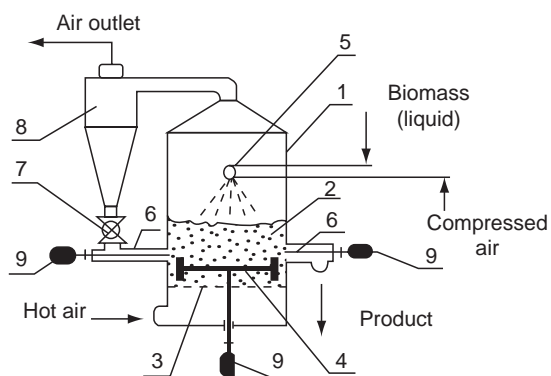


FIGURE 39.15 Wetted bed fluidized bed dryer: (1) fluidization chamber; (2) fluidized bed; (3) air distributor; (4) disintegrator; (5) nozzle; (6) screw feeder; (7) rotary feeder; (8) cyclone; and (9) motor. (From B. Dencs and Z. Ormos, *Recovery of Solid Content from Ferment Liquor Concentrates in Fluidized Bed Spray Granulator*, *Proceedings of Fifth Conference on Applied Chemistry, Unit Operations and Processes*, 3–7 September, Balatonfured, Hungary, 320–325 (1989). With permission.)

drying in a turbulent bed the dry product is abraded and possibly damage the microorganism cells or mechanically disrupt the biopolymer chains. Hence, vegetative and spore cultures must not be dried. It should be emphasized that for a correctly chosen inert material and at a properly conducted drying process the output expressed as the amount of moisture evaporated in unit volume of the drying chamber is 15 to 17 times higher than the output of a spray dryer at the same temperature.

39.4.5 DRYING IN A FIXED BED

A characteristic feature of drying in a fixed bed is that the material in the bed is fixed (e.g., in band, chamber, and tunnel dryers). The advantage of this method is the application of multisectional or multilevel dryers that facilitate the control of products and adjustment of drying conditions in particular drying zones. Also advantageous are the separate feeding of hot or cold air at a constant or varying flow rate, multiple application of an air portion if it flows through several sections or several band layers, possibility of cocurrent, countercurrent or mixed flow, and recirculation of a drying agent.

The method has some disadvantages, namely, the dried product is not homogeneous as far as moisture content and microbiological quality are concerned, process control is difficult in the case of a one-section dryer, the product can agglomerate and get stuck to trays in drying chambers or onto transporting bands. Long drying time and low dryer output, large size of

the drying installation, and bad sanitary conditions for drying of such products are additional disadvantages of this method.

Drying in a fixed bed is applied when wet material is formed and concentrated on filters or concentrated by adding a filling material and when low drying temperature is required. Chamber and shelf dryers are often applied in low-output technologies, in pilot-scale situations, and in drying of highly thermo- and xerolabile products.

39.4.6 DRYING IN DRUM DRYERS

In drum dryers, material can be dried convectively, inside the drum. There are also cylindrical dryers with contact drying of material on the external surface of the cylinder. The first type of drum dryers are widely used for bulk products. The material should be in the form of granules or dense paste. The efficiency of such drying is not high. The drum dryers of the second type are more frequently used for drying suspensions. A thin layer of the material is retained on the hot cylinder surface, then dried and removed by special scrapers. Such a drying method is suitable for products that do not require special control of drying or a specific temperature regime (e.g., fodder concentrates, waste products).

39.4.7 FREEZE AND VACUUM DRYING

In freeze and vacuum drying, drying takes place at low temperatures, below 0°C. In the case of drying of biotechnological, food, and pharmaceutical products, the temperature ranging from –5 to –40°C and average vacuum are applied. The process has the following advantages:

- Moisture is removed at low temperature, which in fact excludes thermal inactivation of the product.

- The material structure is maintained (particles and cells are not damaged).

- It is easy to obtain a sterile dried product in commercial unit packages (ampules and bottles).

The main disadvantages of freeze drying include long process duration, high energy consumption, and complicated equipment.

In freeze drying, a very important stage is freezing, which directly precedes sublimation. Depending on the method of drying and physicochemical properties of substances, an initial freezing called self-freezing is applied. The initial freezing takes place in the dryer chamber as a result of a slow decrease of temperature and pressure. A very slow cooling unfavorably affects

cells because electrolyte concentration increases in the ambient medium and cells are dehydrated excessively. At a slow and average cooling rate (10 to 100°/min) large ice crystals are formed that can mechanically damage cell structures such as cytoplasmic membranes. The self-freezing takes place at ultraswift cooling (e.g., at the rate of 10,000°/min) as a result of rapid pressure drop. The process takes place in the whole volume of the material. Small ice crystals are distributed uniformly, and in favorable conditions amorphous ice is formed (transformation into a vitreous form, termed vitrification), which does not cause mechanical damage of cells. The self-freezing not only facilitates the technological process, but also intensifies dehydration. It is estimated that the evaporation rate is 3 to 5 times higher than in the process of sublimation. However, the self-freezing is recommended particularly in drying of materials with relatively low moisture content as substances of high moisture content are easily foamed and require initial freezing.

The optimum cooling or freezing rate is different for various microorganisms and biopolymers. For instance, in bacteria, slow freezing (1°/min) to –35°C at the beginning and then rapid cooling to the desired low final temperature is recommended.

In Table 39.5, initial freezing temperatures and percentage of frozen moisture as a function of temperature of several materials are presented. It follows that in most substances the freeze-drying temperature should be lower than –40°C because at this temperature a significant part of moisture remains unfrozen.

In freeze drying, the sublimation temperature and the kinetics of changes of the sublimation surface (or

a displacement of a wet surface) within the material layer are very important (Figure 39.16) [60–62]. The optimal temperature is the one below the eutectic temperature. In practice, however, it is difficult to attain because at that temperature the drying time must be prolonged and so more energy is required. The presence of protective substances in dried material, such as glycerin and sugars (fructose, saccharose, lactose, or specially prepared compositions known only under trademarks), has influence on the eutectic temperature and the applicability of elevated temperatures during freeze drying. The substances protect microorganisms against too high salt concentration and provide more favorable conditions for ice crystallization during freezing. For example, the role and mechanism of glycerin interaction are as follows:

In the temperature range from 0 to –20°C glycerin decreases the crystallization rate, promotes the formation of small crystals, and sustains the state of cooling; in the process of water crystallization, it dissolves salts and, penetrating the cells, it contributes to the restoration of osmotic balance.

In the temperature range from –20 to –80°C glycerin controls the crystallization process and makes it reversible, thus eliminating eutectic solutions caused by the presence of salt solutions. It also shifts the coordinates of eutectic borders and contributes to the formation of vitreous structures.

The effect of various protective substances on the stability of microbiological preparations during

TABLE 39.5
Frozen Moisture as a Function of Temperature, %

Substance	Minimum Freezing Temperature (°C)	Mean Ambient Temperature (°C)				
		Frozen Moisture, %				
		–5	–10	–20	–30	–40
Water						
Distilled	0	100	100	100	100	100
Tap	0.3	94	97	98.5	99	99.2
Virus of aphthous fever	–2.2	56	78	89	92.7	94.5
Fowl cholera strains	–2.4	52	76	88	92	94
Swine erysipelas strains	–2.1	59	79	89.5	93	94.5
Anthrax vaccine	–1.0	80	90	95	96.7	97.5
Strains of comma bacillus	–1.0	80	90	95	96.7	97.5
Lactic acid bacteria	–1.4	72	86	93	95.7	96.5
Skim milk	–0.5	90	95	97.5	98.3	98.8

Source: From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).

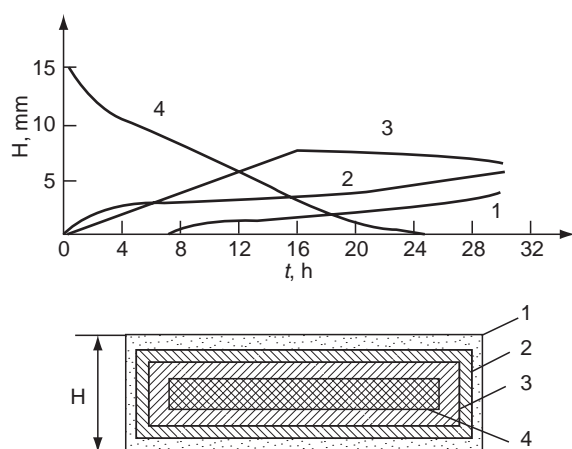


FIGURE 39.16 Moisture content distribution in a slab of freeze-dried material. (From V.G. Popovski, L.A. Bantysh, N.T. Ivasiuk, N.H. Grinberg, and G.B. Gorshunova, *Sublimation Drying of Food Products of Vegetable Origin, Pishchevaya Promyshlennost*, Moscow, (1975) (in Russian).)

freeze- and other drying techniques has been discussed in many publications [63–66] and exemplary results are presented in Figure 39.17 [65].

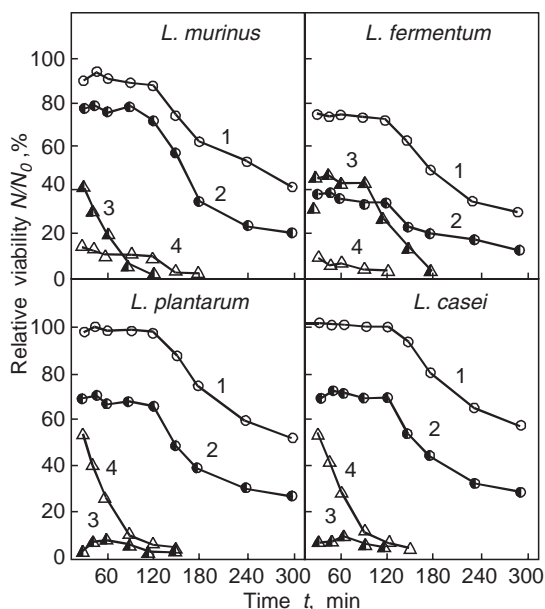


FIGURE 39.17 Effect of the protective substance's medium on the survival rate of *Lactobacilli* at various drying times: (1) adonitol; (2) glutamate; (3) PEG 1000; and (4) nonfat skim milk. (From C.F. De Valdez, G.S. De Giori, and A.P. De Ruiz Holgado, *Effect of Drying Medium on Residual Moisture Content and Viability of Freeze-Dried Lactic Acid Bacteria*, *Applied and Environmental Microbiology*, **49**, 413–415 (1985). With permission.)

While analyzing the sublimation kinetics (Figure 39.16), one can observe that initially, while heating frozen material, the sublimation of ice crystals starts at the material surface and successively penetrates the interior, leaving a solid matrix and void pores through which water vapor must be released to the material surface. In the case of biotechnological products this matrix is in fact a highly concentrated liquid. So, as sublimation proceeds, conduction of heat to the sublimation front results in further dehydration by diffusion. Hence, the consecutive zones of different moisture content appear in a freeze-dried material until all the ice crystals are sublimed.

Since the sublimation rate can be increased by a higher heating rate only to a certain extent, the rate of water vapor removal from the sublimation front becomes the key point in freeze drying. Adsorption of water vapor by sorbents placed inside a drying chamber or by direct contact of the sorbent and drying material (as a mixture) is one of the most promising methods in this area. Molecular sieves, zeolites, or ceramics can be recommended here because their sorption activity under rarefied conditions does not drop with temperature rise due to the heat of sorption. The use of particulate sorbents permits sublimation to take place in a fluidized state under atmospheric pressure [67,68].

A typical freeze dryer consists of a drying chamber, vacuum system, and vapor condenser, which can either be separated or built within the drying chamber. The drying chamber construction and the system of heat supply for sublimation can be solved in many ways. A schematic diagram of a continuous scraper-type freeze dryer especially suitable for biotechnological materials is presented in Figure 39.18 [69]. A liquid product of low concentration of the solid phase (up to 40%) is sprayed by nozzles placed on the rotating central pipe and then freezes on the chamber walls. The dryer chamber is a cylinder equipped with a cooling jacket. The heat of sublimation is supplied by radiators also placed on the central pipe. Dry product is scraped from the chamber walls by adjustable brushes or scrapers.

Figure 39.19 presents a schematic diagram of a continuous vibrogravitational freeze dryer for liquids and pastes [70]. In the upper chamber, the vacuum-frozen liquid is dispersed. The granulated material is transported to the lower chamber equipped with a system of vibrating screens with holes of diameters that decrease with the direction of material transfer. The heat of sublimation is supplied by radiators installed above each screen.

The idea of the intensification of freeze drying by introducing a moisture-adsorbing substance into the drying chamber is shown schematically in Figure 39.20

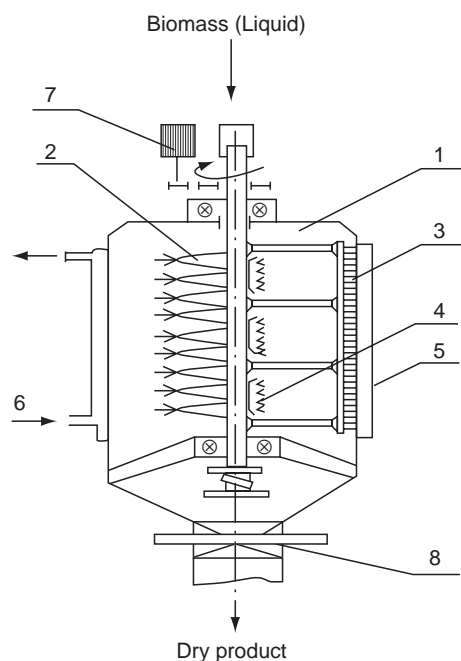


FIGURE 39.18 Continuous scraper-type freeze dryer: (1) vacuum chamber; (2) spray nozzles; (3) scraper; (4) radiators; (5) cooling jacket; (6) freezant; (7) motor; and (8) vacuum lock. (From K.P. Shumski, *Vacuum Apparatus and Equipment in Chemical Industry*, Mashinostroyenie, Moscow, (1974) (in Russian).)

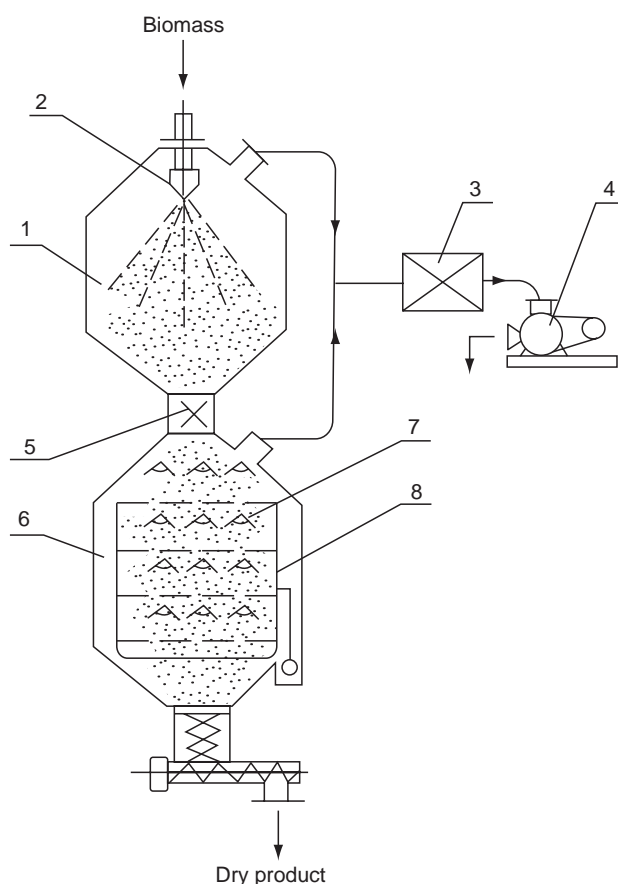


FIGURE 39.19 Continuous vibrogravitational freeze dryer for liquid and pastes: (1) vacuum granulator; (2) material feed; (3) condenser; (4) vacuum pump; (5) feeding valve; (6) drying chamber; (7) radiators; and (8) vibrating trays. (From P.A. Novikov, I.F. Pikus, and E.G. Tutova, *Continuous Freeze-Dryer for Liquid Materials*, Russian Patent No. 27,3734 (1970).)

and Figure 39.21 [20]. Figure 39.20 presents a multi-band continuous dryer in which dried material and sorbent are transferred countercurrently on adjacent bands. In Figure 39.21 the sorbent is transported along the grid baffle placed over the connecting pipes to which containers (bottles or ampules) with dried ma-

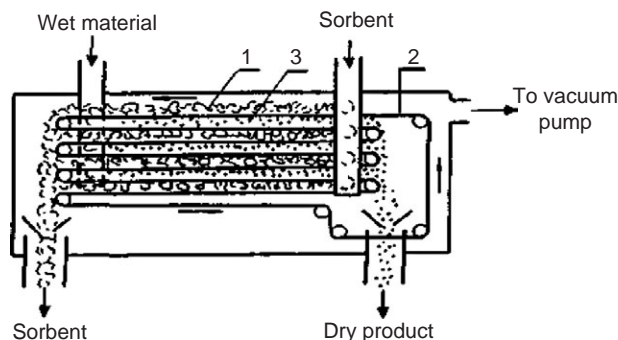


FIGURE 39.20 Scheme of the multiband continuous freeze dryer of countercurrent dislocated material and sorbent layers: (1) sorbent; (2) transporting band; and (3) dried material. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

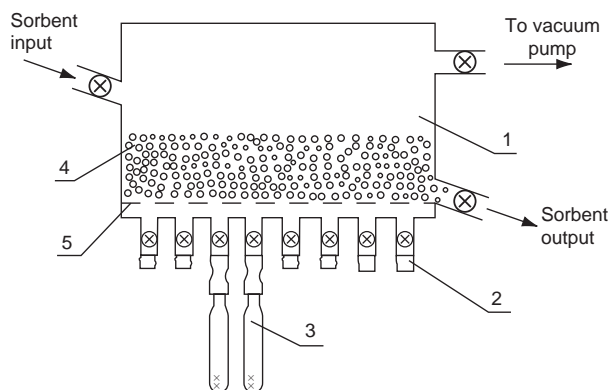


FIGURE 39.21 Scheme of the device for ampuled material dehydration by freeze/vacuum drying: (1) sorbent reservoir; (2) connecting pipes for ampule coupling; (3) ampules containing dried product; (4) sorbent material; and (5) grid for sorbent charging. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

terial are attached. In both design solutions the transfer of sorbent layers takes place near the dried substances, from which the sorbent sublimates moisture, enhancing the evaporation rate significantly. The costs of sorbents and the energy consumed during their recovery justify the application of such solutions because the process of sublimation is intensified and product quality is improved as a result of decrease of the process temperature and duration.

39.4.8 DRYING OF MICROORGANISMS AND BIOTECHNOLOGICAL PRODUCTS ON CARRIERS

This is a method of drying suitable for substances used as mixtures. One of the components of the mixture is a carrier. In the literature, this drying method is known under various labels: drying on carrier material, drying with immobilized cells, or contact-sorption drying. Material that participates in the process is called a carrier, filling material, or sorbent. The results of experimental investigations and theoretical assumptions concerning this drying method are discussed elsewhere [20,24,71–77] and in a monograph on microbiological concentrates [19].

Let us analyze the role of material called a carrier and its properties:

The carrier introduced into the biotechnological product suspension changes the form of a substance being dried (from liquid or suspension, a granulated product is obtained).

Hygroscopicity of the material changes.

A biotechnological product in the mixture changes its xerolability.

As a result, the technology and method of drying the mixture can be changed, and so a product of high quality is obtained. The carrier is a material that protects cells or biopolymers against mechanical or thermal destruction during drying. The carrier as a sorbent takes over part of moisture from biomass and thus the amount of water to be evaporated is smaller, which causes energy savings. Another source of energy savings is that the process can be carried out at a medium temperature of the drying agent, which contributes to the diminishing of heat losses. Two basic features of the carrier include:

1. Its suitability for application and use in the mixture together with biotechnological products.
2. Neutral interactions with product. (The material must not destroy or deteriorate product quality; also, it cannot be a medium on which

an uncontrolled growth and development of product takes place.)

The carrier should have lower hygroscopicity, which facilitates long-term storage of the product without the use of specific methods and conditions. It should also be easily available and cheap. The carrier is often a by-product of another technology, and then its application in preserving biotechnology products contributes to the protection of the environment against solid wastes.

Tutova and Kuts [20] propose to evaluate the sorbent suitability from the point of view of its sorptive abilities. They characterize material comparing sorptive activities (i.e., relative sorptivity K or hygroscopicity) that under thermodynamic equilibrium of the system is written as follows:

$$K = \frac{a_s}{a_r} = \frac{X_{s,e}}{X_{r,e}} \quad (39.20)$$

where a_s and $X_{s,e}$ are the absorptivity and equilibrium moisture content of the sorbent, respectively, and a_r and $X_{r,e}$ are the absorptivity and equilibrium moisture content of a reference material, respectively. A *reference material*, or a *matrix*, is a filter paper with constant absorptivity for the whole range of relative air humidities. In calculations and the diagram presented in Figure 39.22, the authors took the equilibrium

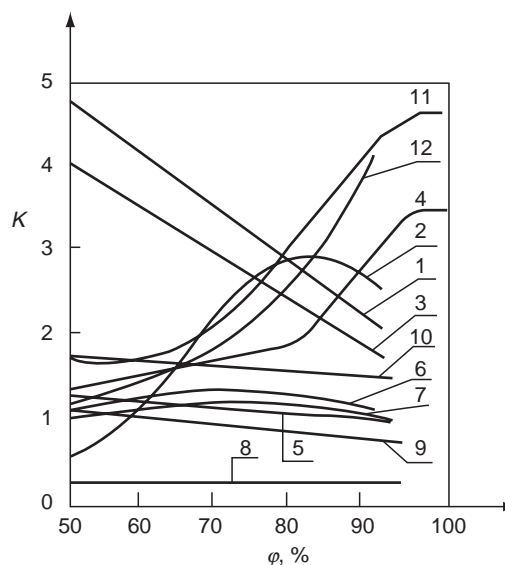


FIGURE 39.22 Relative absorptivity K of some sorbents: (1) Zeolite CaA; (2, 3) activated carbon; (4), wheat bran; (5) maize meal; (6, 7) peat; (8) kaolin; (9, 10) potato flour; and (11, 12) silica gel. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

moisture content $X_{r,e} = 0.01 \cdot X_{e,max}$, where $X_{e,max}$ is the maximum equilibrium moisture content of the matrix at 25°C. From the diagram in Figure 39.22, materials of mean sorptive activity ($K = 1$ to 1.5, e.g., maize meal, potato flour, and peat) with an absorptivity K that does not depend on air humidity, φ , and materials of high sorptive activity ($K > 2$, e.g., zeolite, silica gel, and activated carbon) with absorptivity that depends to a large extent on the relative air humidity, φ , can be selected.

The method for drying biotechnological products on carriers can be used in every technique discussed above. The most interesting is spray and fluidized/vibrofluidized bed drying. These methods differ not only in the technique of drying but also in the way of preparing substrates. In the spray dryer a material and a carrier get in contact directly in the dryer chamber. Both materials are dispersed and at the same time the product is adsorbed on the surface of carrier particles, moisture transfer between the materials takes place, and the process of evaporation proceeds. During fluidized bed drying (and also vibrofluidized bed, drum, and band drying), granulated material should be supplied to the dryer chamber. In this case the contact of the biotechnological product and carrier takes place in a separate mixing chamber (this also can be in a fluidized bed or in a screw conveyor), from which the mixture passes to a batch feeder or a granulator and next to the drying chamber.

Examples of the design solutions are shown schematically in Figure 39.23 through Figure 39.26 [20]. Figure 39.23 presents dispersion of a carrier in the zone of drying agent feeding and biomass suspension spraying in a cocurrent spray dryer. A mixture that forms in the contact zone is transported with a stream of drying agent and removed from the dryer to a cyclone. Figure 39.24a and Figure 39.24b show schematically a chamber in the spray-fluidized bed dryer. It presents also a special three-channel nozzle. Biomass is sprayed in the drying chamber by two nozzles, the central nozzle (Figure 39.24b) is used to spray and contact the two materials (i.e., a biomass suspension and carrier). In the lower part of the chamber the mixture is dried in a fluidized bed by means of a separate air stream.

Figure 39.25 presents a schematic diagram of a two-stage fluidized bed dryer with a separate feed for each stage with a drying agent, and a granulated biomass and carrier mixture fed to the drying chamber. The mixture of both materials is prepared in a screw conveyor during the transport of material to the feeder-granulator. The drying material is transported along perforated plates and gets in contact counter-currently with the drying agent of variable parameters.

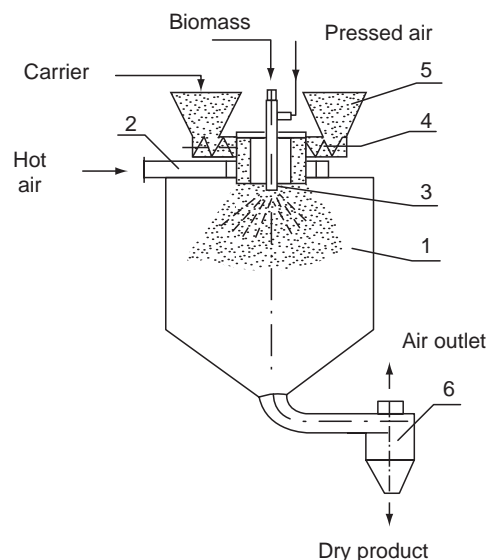


FIGURE 39.23 Scheme of the cocurrent spray dryer with a carrier dispersion in the zone of biomass suspension spraying and drying agent feeding: (1) drying chamber; (2) air duct; (3) spray nozzle; (4) carrier feed; (5) carrier tank; and (6) cyclone. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

Figure 39.26 presents an interesting two-stage dryer with a fluidized bed for granulation and drying of paste-like biotechnological products. In the upper chamber, material is prepared for drying, that is, in the fluidized bed granules of biomass are mixed with carrier particles that stick to granules, adsorb part of the moisture, and protect granules from agglomeration and sticking to the chamber walls. The granulated material prepared in such a way is transported to the lower chamber where, in the fluidized bed, the mixture is dried.

39.4.9 DRYING OF ENCAPSULATED BIOTECHNOLOGICAL PRODUCTS

Recently, special attention has been paid to the up-to-date technology of encapsulation of valuable substances and next drying of these systems so that they could be stored in a stable state. So far, the technology has been most often applied in cosmetic, pharmaceutical, and food industries, mainly for the production of aromatic and coloring substances. Lately, there are many references to the studies and applications of this technology in biotechnological products [78–86].

This technology is related to the immobilization of biotechnological substances mentioned previously. It consists in a durable combination and often a

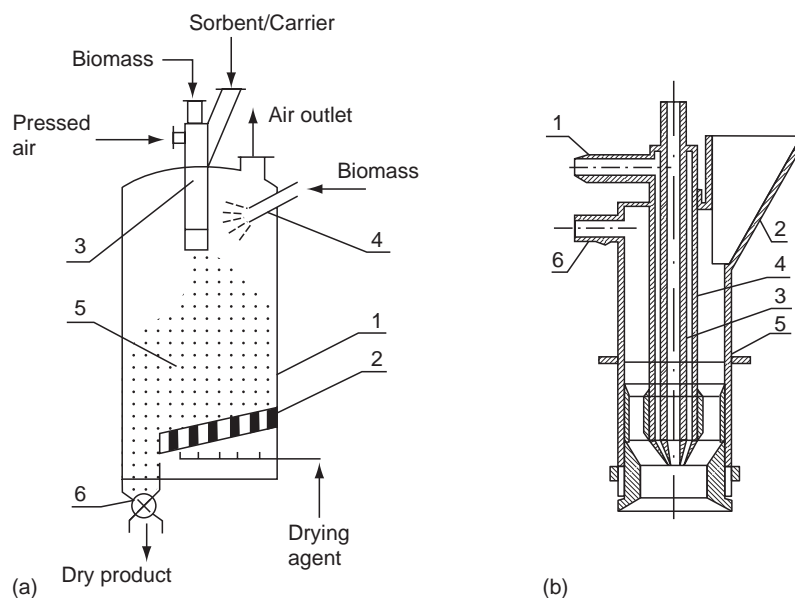


FIGURE 39.24 Scheme of the spray-fluidized bed dryer: (a) components mixing in a dryer chamber: (1) Mixing-drying chamber; (2) grid; (3) spraying device (three-channel pneumatic sprayer); (4) sprayer; (5) fluidized bed; and (6) valve; (b) scheme of the spraying device (three-channel pneumatic sprayer): (1) Pressed air feeding valve; (2) sorbent vessel; (3) biomass feeding pipe; (4) air feeding pipe; (5) sorbent feeding pipe; and (6) blow-through valve of the external duct. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

capsulation of a sensitive component in the stable structure that protects it against external factors. Additionally, in the systems characterized by such properties the protected component is released in a

way controlled by external factors and time of exposure. So, complex systems of medicines (antibiotics and vitamins) are obtained which are slowly released in the organism. Due to this, the medicine is not

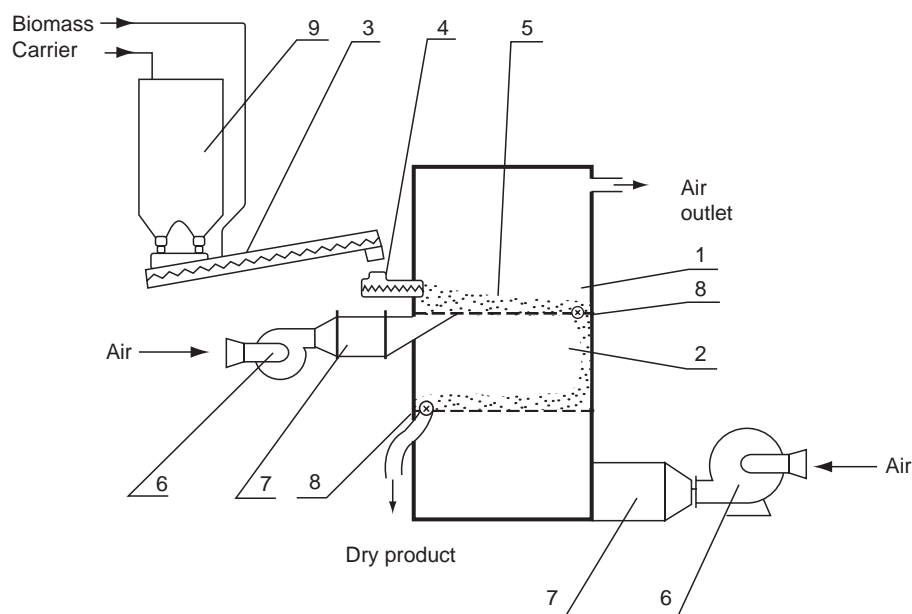


FIGURE 39.25 Scheme of the two-stage fluidized bed dryer with carrier-porous material: (1) drying chamber I; (2) drying chamber II; (3) screw conveyor; (4) feeder-granulator; (5) fluidized bed; (6) blower; (7) heater; (8) valve; and (9) carrier tank. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

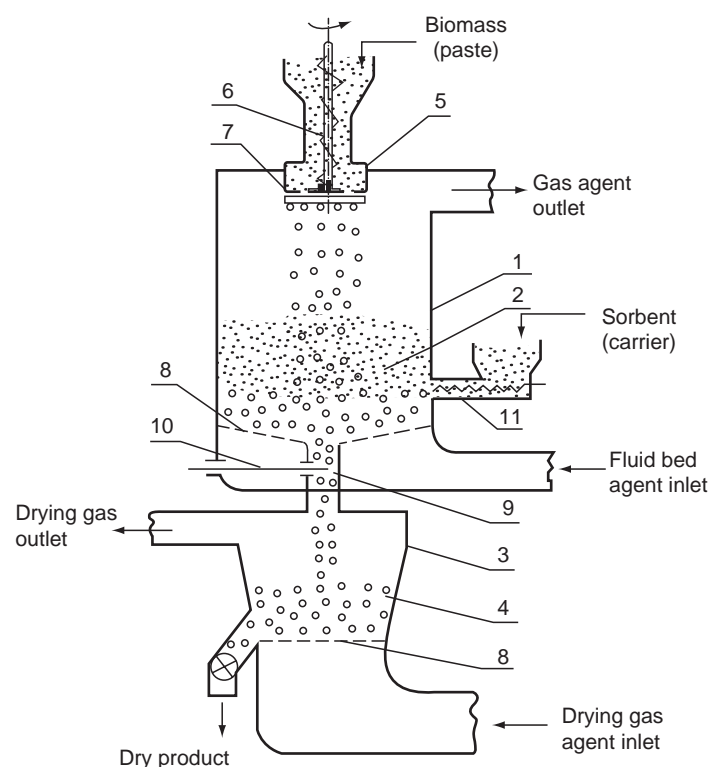


FIGURE 39.26 Scheme of the apparatus for paste-like material drying in a sorbent fluidized bed: (1) chamber for granulation and predrying of granules in a falling bed; (2) sorbent-fluidized bed; (3) drying chamber; (4) granule bed; (5) granulator; (6) screw; (7) moist granules; (8) gas distributors; (9) granule overflow channel; (10) gate and batcher; and (11) drying device for sorbent supply. (From E.G. Tutova and P.S. Kuts, *Drying of Microbiological Products*, Agropromizdat, Moscow, (1987) (in Russian).)

excreted after a short time and its action is prolonged to many hours. The aim of encapsulation is similar in the case of enzymes, bacteria, protein structures, and other biotechnological products.

Drying of such systems is not a widely applied technique and the results of investigations published in the literature most often refer to laboratory scale. As dry product is required in the form of very fine spheres with the diameter around 10 μm , the process of drying is usually carried out in spray or freeze dryers. In the encapsulation process, one of the fundamental questions is a proper choice of material that forms a structure carrying or encapsulating a biotechnological substance. Therefore, significant part of studies and publications are concerned with different behaviors of such materials as starch with different degrees of modification and origin (e.g., potato and maize), gelling substances (rubber of different origin), biopolymers (e.g., chitosan and polyglycols). Frequently, a proper choice of the material for microcapsule production is most important for reaching this aim.

It is worth mentioning that in order to solve correctly the problems of encapsulation (in other words

micro- or bioencapsulation) adequate knowledge is required, and industry and academia people representing different branches should cooperate. So, very popular are joint conferences such as the World Congress on Encapsulation and international research teams and projects (e.g., the Bioencapsulation Research Group in Nantes, France [87,88]). The contribution of drying specialists to this technology is indispensable. Only a proper selection of drying methods and the most precise designing of drying and control systems can guarantee that such a specific final dry product will have the desired properties.

Summing up the above review of drying techniques and dryer constructions for biotechnological products, the following factors are worth stressing:

1. Shortening of high temperature interaction of a product (e.g., cooling of the spray-drying chamber walls, drying at a lower temperature in multistage systems).
2. Possibility of a frequent control, adjustment, and changes of drying parameters in many points of the dryer connected with the control of product quality changes.

3. Possibility of a complete change of the drying technique by the application of carriers/sorbents in multicomponent biotechnological concentrates.

39.5 CONCLUSION

The subject of biotechnological products covers a variety of complex problems concerning production, preservation, and storage, maintaining at the same time high product quality. One of the very important stages in the technology of biotechnological products is the drying process, which has the decisive influence on product quality and preservation. The process of drying for the purpose of product preservation should be strictly adjusted and controlled because of high sensitivity of substances to changes of temperature and moisture content as well as to the process duration. The program of design of process parameters, choice of a drying method, and dryer construction should cover the following tasks:

1. The influence of drying parameters (e.g., range of water removal, change of temperature, time) on changes of product quality and marking of the critical values of these parameters.
2. A mathematical description of the kinetics of degrading changes and choice of the most important quality indices.
3. Many drying methods can be used with respect to individual properties of biotechnological products, e.g., changes of thermal and hydrodynamic conditions in multistage drying installations or change of technique and operating parameters in the case of drying with carrier/sorbent materials.

It should be stressed that the quality of a biotechnological product is also decided by instantaneous fermentation conditions of a unit charge of a product; thus a product being dried can have properties that change in time. Therefore, preservation of specific product quality can be achieved by adding special protective substances to the product of fermentation that preserve product quality during drying.

39.6 NOMENCLATURE

a	activity, units/mg
a	moisture absorptivity, % by weight
A	index of quality degradation, %
A_g	global index of quality degradation, %
C	concentration, number of cells/cm ³ , kg/m ³
d_p	Sauter mean particle diameter, m (in Table 39.4)

E_a	activation energy, J/mol
f, g	function
i	successive component or quality index
k	heat utilization coefficient
k_d	reaction rate constant, s ⁻¹
k_∞	frequency coefficient, s ⁻¹
K	relative sorptivity
n	number (e.g., function and indices)
p	empirical constant in Equation 39.6 and Equation 39.7
r	reaction rate, units/(mg·s) or number of cells/(cm ³ ·s)
r	variable pellet (droplet) radius, m
R	pellet radius, m
R	gas constant, J/(mol·K)
t	time, s
T	temperature, °C
w_i	density of mass stream, kg/(m ² ·s)
W_{H_2O}	evaporative capacity, kg H ₂ O/h
W_v	volumetric evaporation rate, kg H ₂ O/(m ³ ·s)
V_G	air flow ratio, kg of air/kg of H ₂ O
z	empirical constant in Equation 39.10
X	material moisture content, % by weight

Greek Symbols

λ	weight systems
θ	temperature, K

Subscripts

d	degradation
G	gas
i	component
k	final
T	temperature
t	time
w	water
X	moisture
0	beginning
∞	infinite

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40 Drying of Coated Webs

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40.1 INTRODUCTION

The objective of this chapter is to review briefly the drying process, drying equipment, drying strategies, and web handling available for coated webs. Based on the substrate materials, coated webs can be divided into three types: (1) coated paper and paperboard; (2) coated plastic films (e.g., photographic films) and tapes (e.g., adhesive tapes, magnetic tapes, pressure-sensitive tapes, and photosensitive tapes); and (3) coated metallic sheets. Paper and paperboard are coated on machine or off machine, while plastic films, tapes, or metallic sheets are generally coated off machine. (*On machine* indicates the coating operation that is done on the web before it is removed from the original manufacturing machine, whereas *off machine* implies the coating operations done on a free-standing machine remote from the original machine.)

During the coating process, some coated webs require a single coating; other webs require more

than one coating layer either by passing a web of material through a single coating station more than one time or by coating a web with a multiple-station coating machine. In the converting industry, paper, films, and foils can be combined together to form multiple-layer structures in a process called *laminating*. In the graphic arts industry, the coated papers are further coated with ink to generate the desired images through a single printing station or multiple-color-printing units. Figure 40.1 shows a finished Polaroid instant color picture containing polyester supports on the top and bottom with active layers sensitized to the three primary colors (blue, green, and red), timing layers, and spacing layers to display the image between the supports.

Coatings, inks, and adhesives contain more than one component: (1) a binder that may include particles to give it a useful function (e.g., pigments for color and opacity, silver halide particles for photographic activity, or iron and chromium particles for

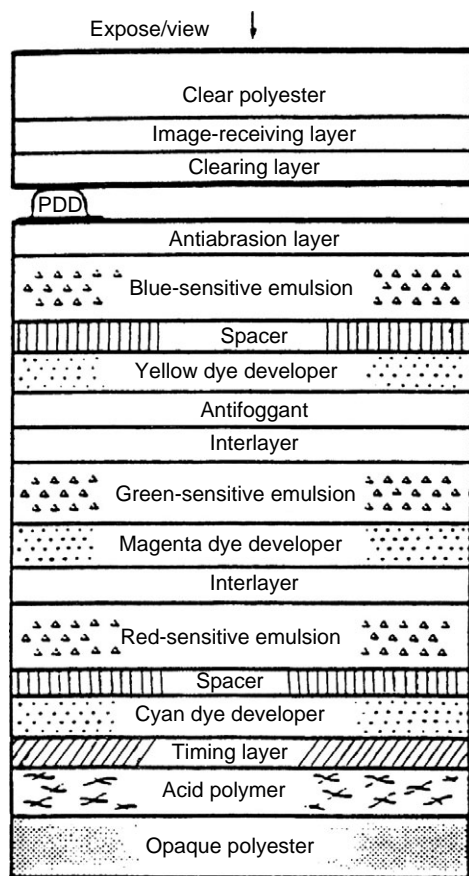


FIGURE 40.1 Makeup of a typical Polaroid film.

magnetic activity); (2) a variety of additives (e.g., surfactants that aid the coating process, plasticizers for flexibility, biocides that prevent bacterial growth, cross-linking agents for toughness, and other additives to minimize static buildup); (3) surface particles to control reflectivity and transport; and (4) a liquid solvent to dissolve or suspend all particles. These components can be reduced to two basic elements: (1) coating liquids and (2) coating solids. The solids can be concentrated on the web surface, such as in coating and printing, or can be distributed throughout a fibrous web, as in saturating or encapsulation, or can be located at the interface between two webs, providing adhesion, as in laminating.

Coatings can be applied in three forms: (1a) solvent borne; (2) water based; and (3) 100% solids. Both solvent and water types must pass through an evaporation dryer to remove the diluent (e.g., solvent or water) so that a dry film remains on the web or sheet. Organic solvents require use of a closed-cycle operation and special care in design and operation owing to the potential fire. A 100% solids coating is a liquid that does not contain any solvent or water. It changes into a solid through a chemical action (i.e., exposure

to a catalyst). The catalysts used in the printing and converting industries are moisture, heat, and ultraviolet (UV) and electron beam (EB) energy. The web coated with solventless liquids usually passes through a hot-air dryer for heat-active liquids, or a UV or an EB emitter for UV and EB liquids. The antipollution regulations have caused the publication, commercial, and packaging printers to seek a replacement for solvent types of printing inks. The three most commercially acceptable replacement printing inks are: (1) aqueous; (2) high-solids (heat-curable); and (3) solventless (100% solids) inks.

Most coating systems involve drying as well as curing of the coating. The drying of the coated webs involves a combined transfer of heat and mass. During the transfer process, water or solvent vapor is removed from the web, leaving nonvolatile solids behind. Heat transfer resulting from temperature difference between a coated web and its surrounding media can be accomplished by conduction, convection, radiation such as infrared (IR), or a combination of these methods. Mass transfer occurs within the coated web primarily through capillary force and vapor diffusion, and from coating surface to its surrounding air by diffusion or forced ventilation. While conductive and radiative heating are useful techniques for some applications, convective heating is by far the most common means of supplying the energy needed to evaporate water or solvent, and is the only form of heating that also provides a means of enhancing the transport of water or solvent vapor away from the coating surface. Curing involves the cross-linking mechanism of UV- or EB-curable materials, or drying and hardening of any film coating by chemical reaction, especially polymerization.

Air-knife coating poses a special consideration in drying system design because of the low solids contents (14 to 20% solids) of the coating applied, e.g., in carbonless copy coating, in which coating applied is in the 40 to 50 g/m² range and the coat weight (dried product) is in the 2 to 5 g/m² range. The high water content causes extension of the web between the coater and the dryer and contraction during the drying. In on-machine coating of board substrates, this problem is not encountered. Air-knife coating systems operate with web speeds up to 600 m/min; in most cases 450 m/min is the maximum speed. Other coating techniques, e.g., blade or roll coating, may operate at speeds up to 1200 m/min.

40.2 DRYING CURVES

In order to design an efficient drying system for coatings, the individual drying curves showing their drying mechanism must be taken into consideration.

Because of the product quality problems associated with the drying process, the proper quantity of heat and mass transfer has to be applied to the coating at the proper time in the drying curve. There are two basic types of drying curves that can be easily constructed from drying data. The first type of drying curve is shown in Figure 40.2, representing percent solvent remaining in the coating vs. drying time for a typical solvent-based coating of about 0.1 mm wet thickness. Although this curve shows that the drying rate decreases with drying time, there are no drying rate values at various points in time for the effective drying equipment or optimal drying process control.

More information is given in the second type of drying curve (Figure 40.3) constructed from the same drying data. Here, the drying curve is a plot of drying rate as a function of drying time and reveals three distinctive drying rate periods: (1) the adjustment stage or the cooling-down period as represented by the segment A–B; (2) the constant drying rate period represented by the segment B–C; and (3) the falling drying rate period by the segment C–D.

Figure 40.3 also represents many solvent-based coatings. A rather short duration of cooling down and constant drying rate period is then followed by a very long period of falling drying rate. In contrast, the drying curve of water-based latex coatings has a longer constant drying rate than solvent-based coatings. Paper has a drying curve with a short warm-up period and a constant drying rate period even longer than latex coatings (Figure 40.4). Once the drying mechanism or drying characteristics of each drying rate stage are identified for each coated web, the proper design and control of the drying system can

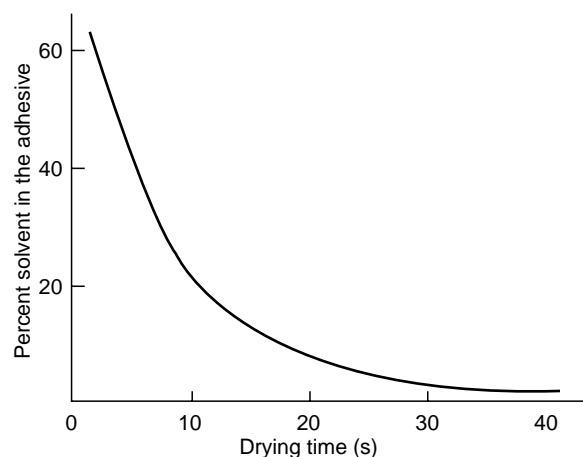


FIGURE 40.2 Drying curve (solvent remaining vs. drying time).

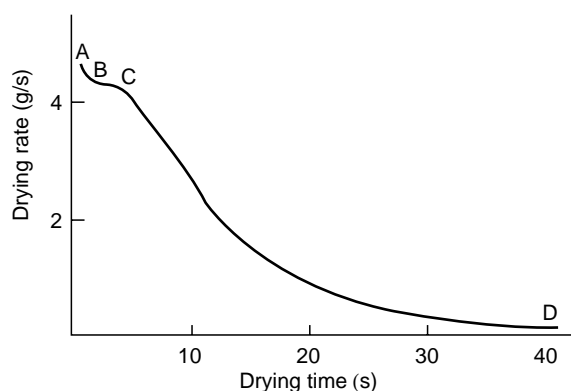


FIGURE 40.3 Typical drying rate curve (drying rate vs. drying time).

be offered to obtain the best quality products as each drying rate stage has its special drying mechanism to affect the finished product quality. For instance, Figure 40.3 shows a large amount of drying solvent evaporates during the constant drying rate period; it is desirable to design a drying system to extend this constant drying rate period for optimal drying operation.

The success of a drying system for coatings requires not only a means to remove water and solvent vapor from the coated webs, but also proper drying operation to accomplish the quality requirements necessary for further application. For example, most coated papers are intended for printing, and binder migration causing printing mottles should be eliminated or minimized during drying operation. Binder migration can be affected by the rate in the preheating stage. The time in the constant drying rate zone and the temperature in the falling drying rate zone also affect many of the final sheet properties. Over-drying or rapid drying can cause brittleness, blistering, curl, and wrinkles. Brightness, ink receptivity,

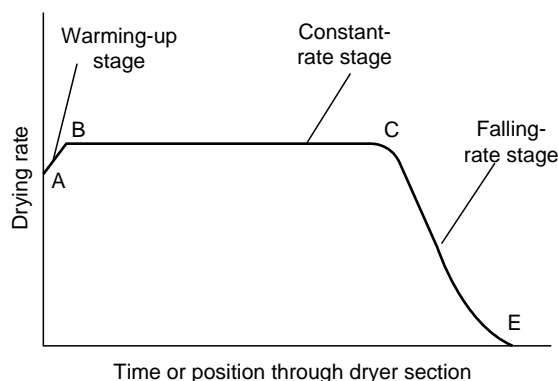


FIGURE 40.4 Typical drying rate curve for paper.

and varnishability can be reduced. Nonuniform moisture profile, which is normally caused by raw stock or coating, can sometimes also be attributed to drying conditions.

40.3 DRYING TECHNIQUES FOR COATED WEBS

The following are the principal hardware available for drying of coated webs:

1. Steam-heated cylinders
2. High-velocity air cap
3. Impinging jet tunnel dryers
4. Air flotation dryers
5. Air turns
6. IR dryers
7. Ultraviolet curers/dryers
8. EB curers/dryers

40.3.1 STEAM-HEATED CYLINDERS

In the papermaking process, the steam-heated cylinders have been traditionally placed in staggered positions for drying of paper web because of their ease and economics in operation. However, coated webs cannot be dried on the same steam-heated cylinder configuration without special precautions because of picking of the coatings. To prevent picking, the drying cylinder surface needs a special finish, particularly on the first few cylinders. The CIS (coated one-sided) web is made to wrap around a series of steam-heated cylinders (1 to 2.5 m diameter, steam pressures of 3 to 5 kg/cm²), yielding a rather low evaporation capacity of 3 to 6 kg/m²·h. One of the new solutions to this picking problem includes the use of circular flotation dryers' jet foil cylinder in the location of the first and second steam cylinders (Figure 40.5). This initial con-

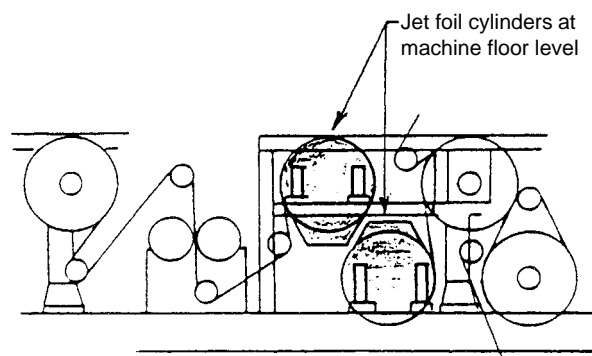


FIGURE 40.5 Steam-heated dryer configuration.

vection air drying would allow enough evaporation to prevent picking on downstream cylinders. (The jet foil cylinder is a unique contactless drying system designed in a circular configuration. Contactless drying is achieved using combination of airfoil and pressure pad air flotation techniques described in Section 40.3.4 and Section 40.3.5. High-temperature circulating air is the heat and mass transfer medium, while total circulation with a controlled exhaust ensures economic operation.)

Many times a steam-heated cylinder has a felt on the opposite side of the web from the steam-heated cylinder to insure good contact with the cylinder at elevated web speeds and low web tensions. Drying rates would range from 2.4 (assuming full wrap) to 7.3 kg/m²·h, depending upon whether the steam-heated cylinder is operated with or without a felt and whether or not the coated side is toward the cylinder or away from it. Also, the drying would need to vary for coating quality reasons, depending on which point in the drying curve is being addressed. Drying rate enhancement can be accomplished through pocket ventilation systems that evacuate evaporated compounds between steam-heated cylinders. Also, heat transfer rates will be increased through enhanced design of steam delivery and removal within the cylinder.

40.3.2 HIGH-VELOCITY AIR CAP

Air cap dryers resemble the Yankee dryer, wrapping around a steam-heated cylinder as shown in Figure 40.6. The Yankee dryer was designed for drying of tissue and towel, whereas the air cap was intended for drying of paper coating when the blade coaters were introduced. Air caps consist of a series of circular or slot nozzles with a pressure and exhaust plenum hood, providing high-velocity hot air for rapid drying by penetration of the vapor barrier on the coated side of the web. Both circular and slot nozzle spacing typically range from 10 to 25 mm and the nozzle clearance from the steam cylinder surface varies from 6.5 to 50 mm. The air cap must be retractable to permit easy threading and broke clearance. One air cap is generally adequate to immobilize the coating (75% dryness) if the web speed is no more than 500 m/min. Air cap systems can be used as long as the drying or curing load can be handled by a 1.5 to 2.4 m diameter cylinder or air cap assembly.

Air cap drying rates range from 35 to 95 kg/m²·h with jet air at 50 to 60 m/s at temperatures from 150 to 315°C. Higher drying rates are attainable with higher jet velocities (above 60 m/s). However, very high jet velocities can cause flow of the coating; the typical operating range is 40 to 60 m/s. Lower velocities, 20 to 40 m/s, are needed when the solid content

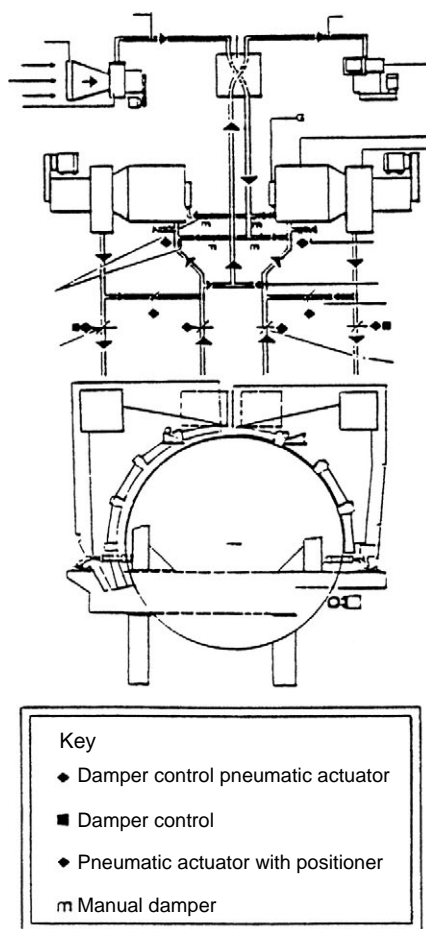


FIGURE 40.6 Air cap dryer layout.

of the coating formulation is low. For acceptable thermal efficiency, significant recirculation of the air (70 to 80%) is needed.

Binder migration is a common problem that depends on the coating formulation, substrate characteristics, and coating rheology. Upper drying rates are imposed by the onset of binder migration. The use of natural starch binders tends to impose a lower drying

rate limit to avoid binder migration, and synthetic binders (e.g., latex) permit drying rates of up to $35 \text{ kg/m}^2 \cdot \text{h}$. Little migration occurs once the coating is about 75% dryness. Further, higher drying rates can be accomplished with little migration problem if the thermal energy is supplied from both sides of the web. However, the high-velocity air impinging directly on the coating surface tends to dry the surface coating and accelerates the binding migration in the coating. “Railroad tracking” will take place when the dryer temperature exceeds the wet bulb temperature of the constant drying rate period.

40.3.3 IMPINGING JET TUNNEL DRYERS

If the impingement length of 4 m is inadequate in an air cap dryer, longer dwell times (or lengths) can be provided in a tunnel dryer (Figure 40.7). In the 200 to 250 m/min range, the web is supported on bars carried through a tunnel on conveyor chains while hot air is impinged on the coated surface. In improved versions of this impingement tunnel dryer, the web is supported on rolls (driven, undriven, or tendency driven). To avoid sag and flutter, slot nozzle impingement is typically applied above each web support roll. For lightweight webs running at speeds up to 450 m/min, a fabric support is commonly employed.

Both round jet and slot jet arrays are used. In the early drying period, slots may be preferred to minimize disturbance to the web coating. Jet velocities up to 75 m/s are used. Air-recycling rates up to 92% may be needed for optimal thermal efficiency. To avoid the possibility of condensation on the cold web as it enters the dryer, the web may be preheated using IR lamps. The air jet humidity is typically in the range of 0.016 to 0.25 kg water/kg dry air.

40.3.4 AIR FLOTATION DRYERS

There are two types of air flotation dryers: (1) the single-slot airfoil dryer and (2) the double-slot air-bar dryer. Figure 40.8 shows the pressure distribution

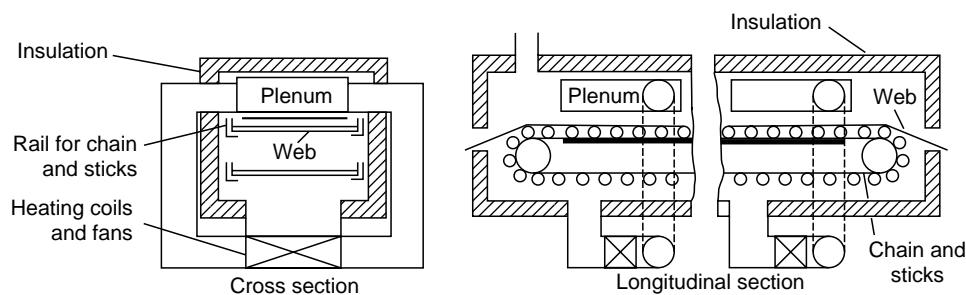


FIGURE 40.7 Conveyor tunnel dryer schematic.

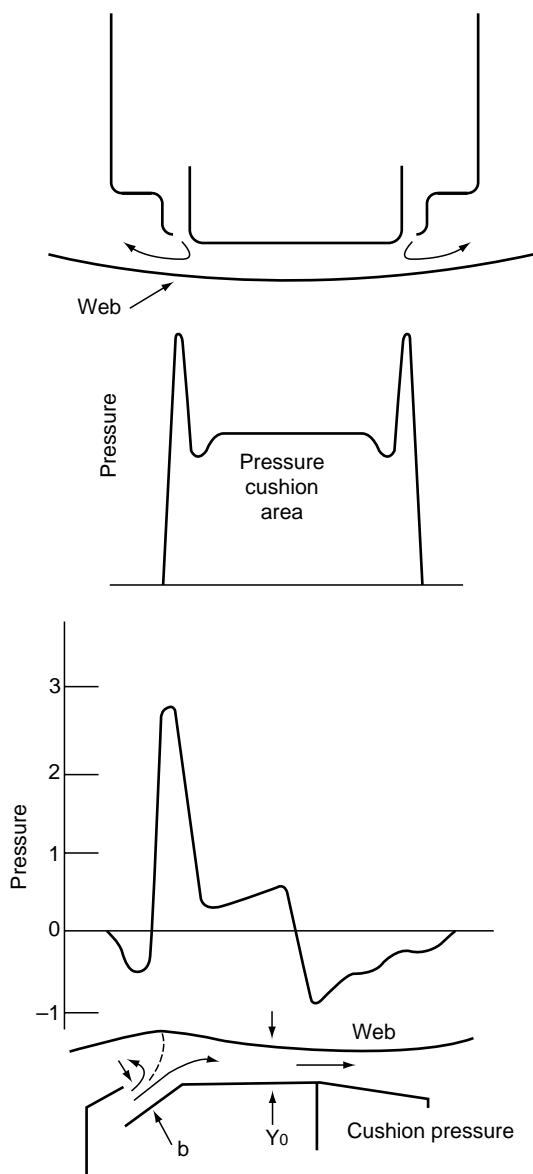


FIGURE 40.8 Pressure distribution of an air bar/airfoil.

on the flat face of an air bar or airfoil. The triple-slot air bar, intended for higher heat transfer, has positive pressure distribution similar to that of the double-slot air bar.

The single-slot airfoils (which have both a positive and negative pressure on the face of the airfoil) are mounted above the web, while the double-slot air bars, having only positive pressure, are placed on both sides of the web in staggered positions. The moving coated web in the flotation dryer is supported on a cushion of heated air issued from the slots (Figure 40.9). The key requirements in slot design are high heat and mass transfer rates and web stability. Heated air emerges out of the slots at velocities up to 80 m/s

and impinges on the moving web. The double-slot air bars have a more stable flotation than the single-slot airfoils partly because the positive pressure pad of the double-slot air bars reacts against the vertical components of web tension and creates an equilibrium, and partly because the staggered configuration forms a sine wave in the web that eliminates edge curl and provides the best possible web transport without contact. Besides, the double-slot air bars can be operated over a wider range of pressure and clearances.

The standard size of conventional airfoils and air bars ranges from 6 to 8 cm in width. Typical flotation clearances of a single-slot, standard-size airfoil would be 2 to 3 mm as compared with a double-slot air bar having a typical flotation height of 6 to 7 mm. The double-size air bar would create twice the clearance of the single-size air bar. Air bars are typically spaced 25 to 30 cm apart but can be as close as 18 cm for single-size air bars or as far as 1 m for air bars of four times the size, depending on web weight, porosity, tension, needed clearance from the air bar, and air-bar pressure.

The basic airflow system for flotation dryers is shown in Figure 40.10. It consists of three basic components: (1) supply fan; (2) exhaust fan; and (3) heater. The supply fan is sized for the air volume required by the open area of the dryer nozzles and the maximum nozzle outlet velocity. It blows hot air through the air-bar nozzles onto the coated web surface. The spent air is exhausted by an exhaust fan. The heater can be a direct-fired gas burner for air temperatures up to 400°C or high-steam coils for air temperatures up to 200°C. When drying aqueous coatings, the exhaust is led to a recirculation system where part is bled to the atmosphere and the remainder reheated for recirculation. When solvents are involved, this spent air may be incinerated or led to a solvent-recovery unit.

Air floaters used in the graphic arts industry operate at web tensions on the order of 5 to 6 kg/cm; in the coating application, tensions as low as 0.03 kg/cm may be needed. Higher tension stabilizes the sheet. Drying rates up to 60 kg/m² · h are attained in commercial installations. Web speeds of up to 1100 m/min can be handled with web widths of up to 8 m.

An important aspect of any flotation system is the stability of the web as it passes over an air bar. Airflow instabilities near the web can induce web flutter and subsequent web contact with mechanical parts of the dryer, resulting in coating disturbance. Web flutter can come in a multitude of forms, ranging from a violent flapping of the web to a high-frequency drumming. Some designs will allow stable flotation near the center of the web but cannot accommodate proper flotation at the web edges.

In order to avoid these types of instabilities, an air bar should be manufactured to tight tolerances, contain

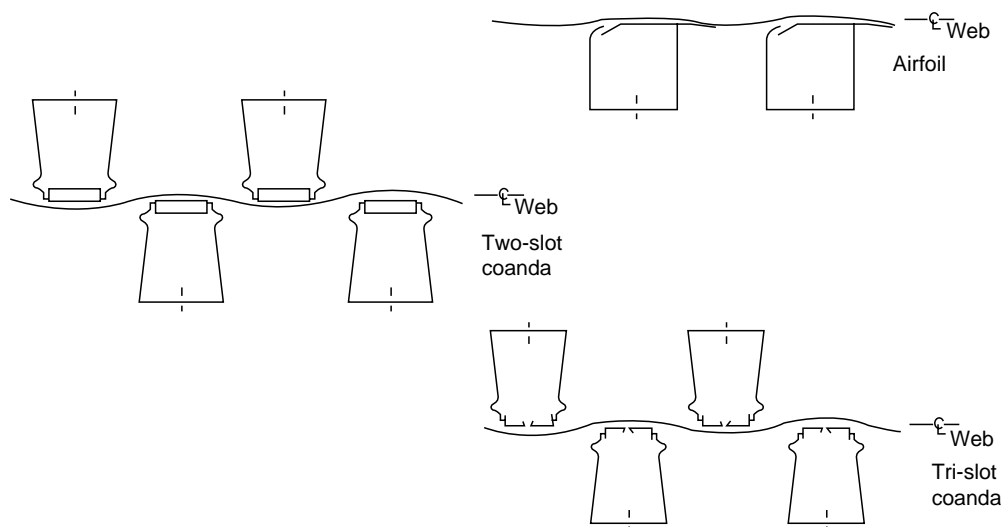


FIGURE 40.9 Flotation air-bar/airfoil arrangement.

“Coander” radius, and have a cushion pressure suppression plate placed at a higher elevation than the impingement nozzles. Also, distance between impingement nozzles and air-bar alignment should be optimized. By following these basic rules and delivering the supply air evenly to the air bar, web flutter can be avoided. Also, problems with web shift, such as movement of the web toward the drive or operator side, and air-bar noise can be eliminated.

Now flotation air bars come in different sizes. The historical size would place the air bars on approximately 25- or 30-cm centerlines, while one of the large sizes would have air-bar center distances in the 50- to 60-cm range. Even higher sizes are available but are mostly used on nonwoven or coil-coating operations. When adequate heat transfer is maintained by the air-bar design, the need for a longer dryer is avoided and the advantages of the large-scale air bars can be

realized. A double-size air bar would require only one half of the number of air bars of a single-size air-bar dryer. This benefits the user during cleaning and maintenance. Double-size air bars also have wider slots and are therefore less likely to become plugged from debris or foreign materials. The double-size air bar also provides twice the flotation clearance while maintaining the same air horsepower as the smaller air bar. Flotation clearance and stability of flotation are important considerations when producing quality products that require no contact and avoidance of markings on the coating during drying. Double-size air bars make lot of sense when considering all operating conditions. Equipment maintenance, flotation clearance, and capital costs seem to be benefited.

The drying capability of a flotation dryer is often measured in terms of heat transfer coefficient h (kcal/h/m²/°C or Btu/h/ft²/°F). Every air bar or foil has an h value that should have been measured as well as calculated under special conditions. Figure 40.11 shows heat transfer values for the single-, double-, and triple-slot air bars. Air velocity, air temperature, web temperature, orifice coefficient, open area, and spacing play an important role. In some air-bar types, clearance is also a major consideration. For example, single-slot airfoils have a dramatic decrease in both heat transfer and flotation stability as clearance is increased. This compares with double- and triple-slot air bars that have fairly stable heat transfers up to a clearance of 25 mm on the single-size air bar and 50 mm on the double-size air bar.

When comparing air bars, there should be a logical basis. Equal power consumption is one such basis for comparison. Under equal air horsepower, single-, double-, and triple-slot air bars have very distinctive

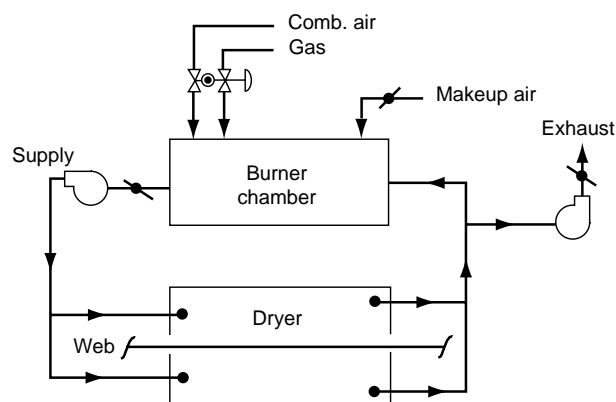


FIGURE 40.10 Flotation dryer airflow system.

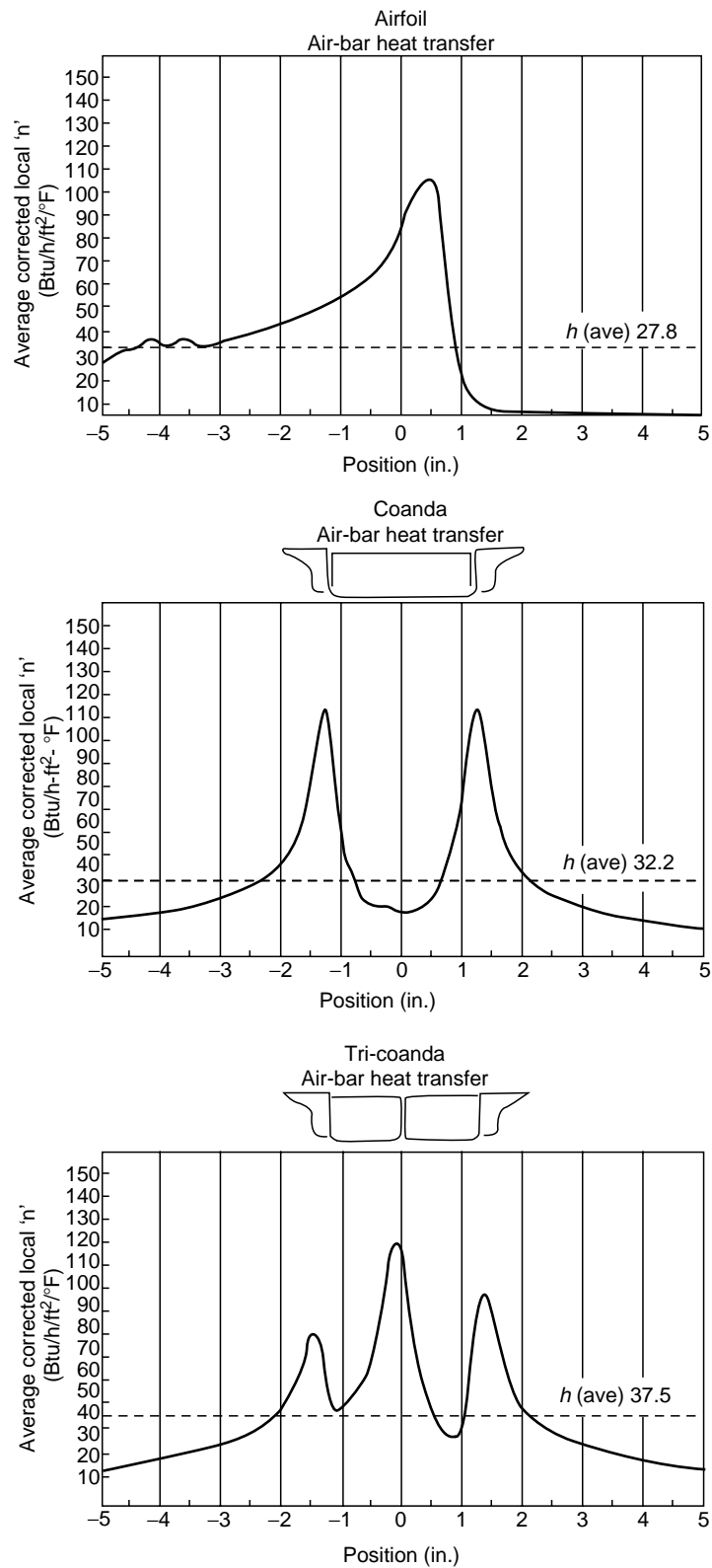


FIGURE 40.11 Heat transfer coefficient comparison.

heat transfer profiles. As seen in Figure 40.11, this TRI-FLOAT has the highest average heat transfer coefficient. The air-bar type uses multiple small orifices to accomplish the same task as the double-slot air bar. Total open area is the same on both.

Heat transfer with double- or triple-slot air bars can be controlled by air velocity over a very broad range since these air-bar types are not nearly as sensitive to flotation clearance as single-slotted air bar. Whenever air can be tolerated on both sides of the web, either the double- or triple-slot air bar is by far the best choice. When flotation must be accomplished with air only on one side, the single-slot airfoil is most efficient.

The impact of spent air on flotation and heat transfer is often overlooked. Delivering the air in a useful fashion to the web is of paramount importance but, if the air after impingement to the web is not removed properly from the web and air-bar area, it can significantly erode an otherwise excellent heat transfer. Drying streaks, flotation problems, or difficulty with web tracking also might occur.

Spent air should be removed from the web and air-bar interface area quickly, evenly, and at substantially reduced velocity as compared with the air impingement velocity. Spent air velocities usually should not exceed 10% of the air-bar outlet velocities. On narrow webs, spent air is exhausted at the web edges. When dealing with wide webs, spent air is removed at the web center to avoid overdrying the web edges.

Air-bar performance is not only dependent on the air-bar design but also is significantly affected by aerodynamically correct supply and return air between air bars and within a given air bar. Air velocity entering the air bar should not exceed 15 to 25% of the air-bar outlet velocity. As with spent air, supply air is often delivered correctly only in one portion of the dryer and incorrectly delivered in another portion.

All successful air bars have a cushion pressure between the face of the air bar and the web. The

cushion pressure profile in the web direction often looks very similar to the heat transfer profile. The profile of a successful positive-pressure air bar would have most, if not all, points between the web and air-bar face at a positive pressure as shown in Figure 40.8. This sounds simple enough but, too often, air-bar designs have pressure profiles with negative pressure just outside the impingement slots. Worse yet are the differing pressure profiles at the web centerline vs. the web edges. The profile at the web edge is of great concern if web flutter is to be avoided at that point. Swings of positive or negative pressure at the web edge will cause the web to act like a flag in heavy wind.

Often overlooked in cushion pressure evaluations is the angular flow from the air-bar slot. If the air emanates from the slots at a very uniform angle toward the drive side or operator side of the web, a steering effect will take place that will hamper good web transport. Usually, proper internal construction of the air bar will eliminate this possibility and allow angularity confined within $\pm 5^\circ$ and in a random fashion. Finally, the amount of air translated from supply pressure to cushion pressure is a sign of an air bar that is efficient. Therefore, the ratio of cushion pressure to supply pressure P_c/P_s is important and worth study at the desired flotation clearance. Figure 40.12 shows a typical curve of this type.

40.3.5 AIR TURNS

Air turns combine the features of a circular web path with a flotation dryer. They use single-size air bars and support a web without contact on a cushion of air while the web follows a circular path and is simultaneously heated and dried. When using an air turn to float and dry, recirculation air is required. Usually, the jet foil cylinder approach provides this option. Air turns were initially proposed for retrofitting paper machines by replacing steam-heated cylinders located

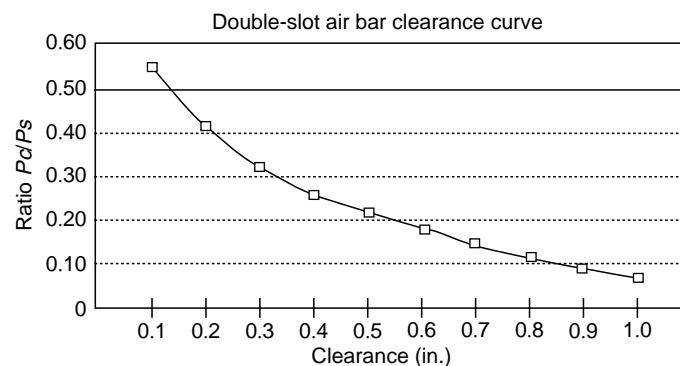


FIGURE 40.12 Clearance curve.

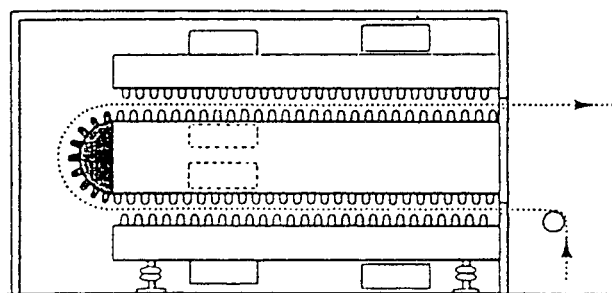


FIGURE 40.13 Air turns in a U shape.

immediately after coating or sizing stations. Quality problems frequently occur in these applications because wet coating is sensitive and sometimes subjected to picking when contacted by a steam dryer surface. Today, air turns are often combined with straight path flotation systems allowing web paths that have a U shape as well as an S shape (Figure 40.13 and Figure 40.14, respectively).

The web-supporting cushion pressure on an air turn is proportional to the web tension and inversely proportional to the turning radius. Assuming a uniform cushion pressure, then

$$P_c(\text{cushion pressure}) = T(\text{web tension})/R(\text{radius})$$

In air turns with double-slot air bars, the cushion pressure is not uniform, but is made up of two separate and distinct pressures. The first is developed over the face of the air bar and is referred to as the pressure pad, while the second is a back pressure component that occurs as a function of the spent air condition, if any, between air bars. The two pressure fractions are weighted by the area according to the proportion of wrap area they each represent; the “weighted” pressure, in total, must be equal to the equivalent uniform cushion pressure required. Pressure combinations as well as pressure and area combinations suitable for any given tension are virtually unlimited.

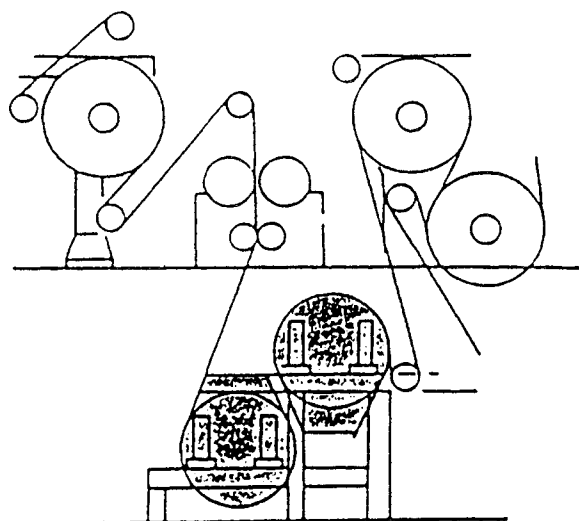


FIGURE 40.14 Air turns in an S shape.

Air turn configurations are typically available in 0.5 to 2.0 m diameters. Volumetric flow rates, pressure requirements, and physical layouts would be considered when sizing an air turn for a particular process. The interrelationship among cushion pressure, web tension, and clearance results in an operational system that is forgiving and self-adjusting. Close attention should be given to the design of web entry and leaving air bars so that touching of the web can be avoided at these points. Heat transfer of the selected air bar will not change when utilized in a circular configuration as compared with a straight path configuration.

40.3.6 INFRARED DRYERS

Figure 40.15 shows the electromagnetic energy spectrum ranging from the gamma ray of short wavelength to the radio waves of long wavelength. The IR waveband falls between the wavelengths of 0.76 and 100 μm , which can be divided into three regions: (1) short-wave

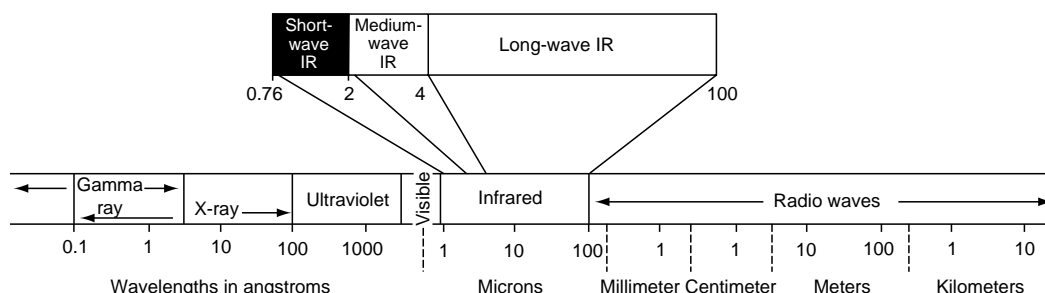


FIGURE 40.15 The electromagnetic energy spectrum.

IR (0.76 to 2 μm); (2) medium-wave IR (2.1 to 4 μm); and (3) long-wave IR (4.1 to 100 μm).

There are two basic types of IR heaters: electric IR and gas-fired IR (Figure 40.16). The most important element of the IR drying system is the IR emitter. These two types of IR heaters fit into three temperature ranges: (1) 343 to 538°C (e.g., gas and electric IR); (2) 538 to 1100°C (e.g., gas and electric IR); and (3) 1100 to 2200°C (e.g., electric IR only). Electrically heated or natural-gas-heated units are used mainly to preheat the web, although these can be used in principle to carry out the entire drying process. IR temperatures are typically in the 650 to 1200°C range. Improved drying rates can be obtained by a combined radiation–convection system.

Most high-temperature units have a cooldown time of 1 to 1.5 min. Occurrence of a break means a significant fire risk if the hot surface is below the web. Since air-knife and blade coaters usually coat the surface, this is not a problem.

In general, the operating efficiency of an electric IR heater ranges from 40 to 70%, while that of gas-fired IR heater ranges from 30 to 50%. These figures depend upon the design of the IR units and moisture content in the web, and represent conversion efficiencies for drying of thin films, coatings, and paper.

The penetration of IR radiation into a material is a function of the wavelength. Higher emitter source

temperatures produce a shorter wavelength, which has the ability to penetrate more than lower wave temperatures, which produce a longer wavelength. The maximum depth of penetration is approximately 1.5 mm. Good penetration into the substrate of coating, film, or ink heats the substrate rather than skinning and blistering it.

Color is sensitive to the IR source temperature. The darker the color, the more heat absorption there is.

The short-wave IR radiation can only be generated by an electric IR heater employing a tungsten heating element. A tungsten filament has small mass as well as low electrical resistance, which allows heavy current flow and very rapid heatup and cooldown. The rate of response of various types of radiant heat sources can be an important criterion in the selection of a proper source for coating drying applications.

The useful IR wavelengths for industrial applications range from 1 to 10 μm (Figure 40.17). For this reason, the IR radiant heating produced by short-wave or medium-wave IR heaters is often used in drying thin coating, films, and webs. The primary criterion for determining which type of IR system and wavelength is most effective for drying of inks, coatings, and webs is the IR absorptivity characteristics of various materials.

Radiation that strikes a material surface must be reflected, absorbed, or transmitted. Only the absorbed radiation raises the temperature of the material. The absorption of incident radiation into a homogenous material increases with the thickness of the material, while the transmission of incident radiation exponentially decreases with the thickness of the material. If the coating or substrate is very thick or opaque to IR radiation, then there is no transmission and the incident radiation is fully absorbed by the coating or substrate.

Figure 40.18 shows the IR absorption spectrum of a 4-mil water film, which is a common ingredient of paper and latex coating. The first peak at 3 μm matches the radiation of medium-wave IR heaters. The second peak at 6 μm and the longer wavelength corresponds to the radiation spectrum of long-wave IR heaters. In order to be effectively heated, the water film requires radiation energy concentrated between 2.6 and 3.3 μm and between 6 and 8 μm . Thus, the selection of an IR unit with properly tuned characteristics becomes important in order to raise the temperature of the target material quickly and effectively. The absorption and radiation curves should be matched as closely as possible, and the absorption rate of the material must be examined. (The absorption of a material is a function of the material's physical, chemical, and color properties, and its moisture content.)

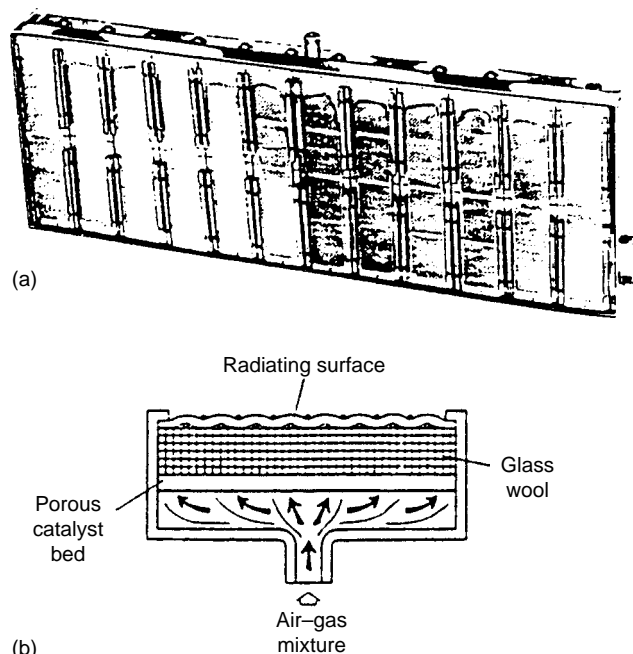


FIGURE 40.16 Infrared radiation heaters: (a) electric and (b) gas fired.

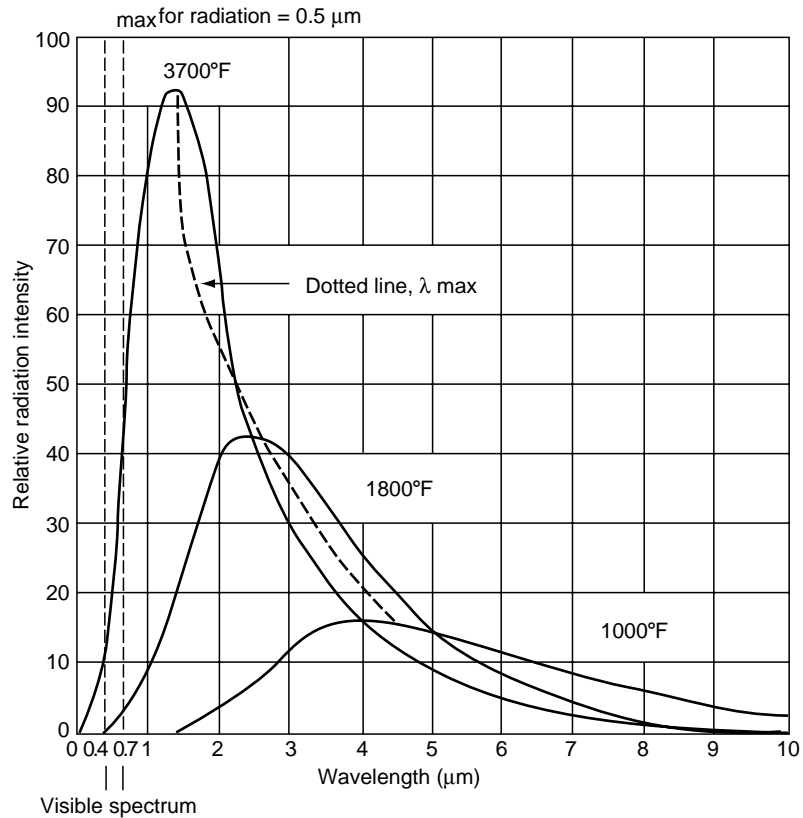


FIGURE 40.17 Infrared radiation energy distribution.

Many times, IR drying is combined with convection or conduction dryers to provide special surface or product properties. Electric IR drying has been very successfully employed as a moisture

profiling system in the paper industry. The IR bulbs are oriented in the web direction and are cycled on or off locally to correct moisture profile differences resulting from the primary drying process and/or

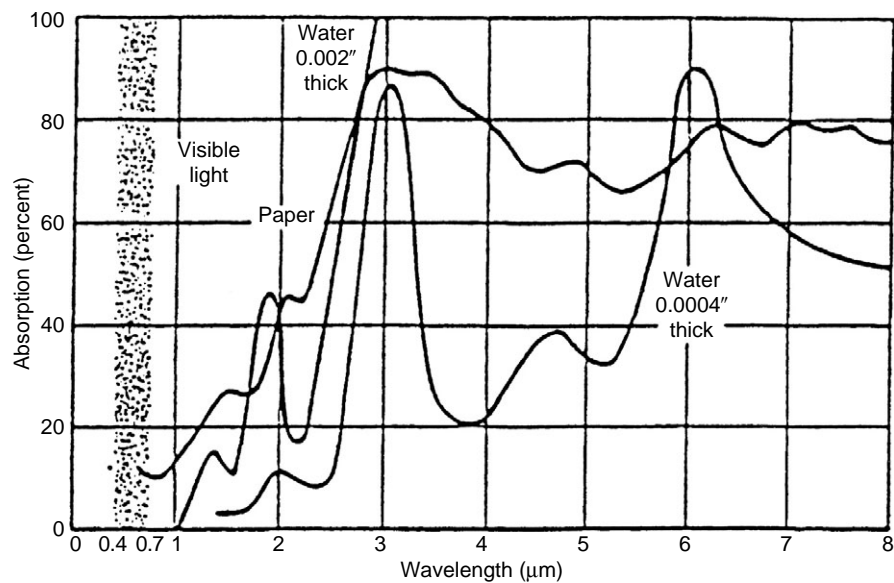


FIGURE 40.18 Absorption of paper and water.

uneven application of water. These systems are almost always connected to sophisticated moisture measurement and electronic devices. Some success has been found in the drying of coated webs, e.g., paper coating, capsulated coating, and thermal papers.

40.3.7 ULTRAVIOLET CURERS/DRYERS

UV dryers are intended for curing of coatings. UV-curable coatings consist of a blend of reactive monomers or oligomers capable of free-radical-initiated polymerization. Photoinitiators are used with UV-curable coatings and they are the source of free radicals produced on irradiation.

UV radiation, which comprises less than 1% of the sun's incident energy, can be selectively generated to promote the electronic excitation of molecules, resulting in chemical changes. UV wavelength ranges from 100 to 400 nm (nanometers) and can be divided into four regions: (1) UVA (315 to 400 nm); (2) UVB (280 to 315 nm); (3) UVC (200 to 280 nm); and (4) vacuum UV (100 to 200 nm). The basic tool used for curing UV-sensitive materials is the mercury vapor lamp, which provides the UV flux activating the photoinitiators in the curable materials. Since different photoinitiators require different UV wavelengths, the radiation from the source and the photoinitiators should be matched to effect polymerization.

Five basic lamp systems are available to produce UV radiation: (1) medium-pressure mercury vapor lamps; (2) electrodeless lamps; (3) pulsed xenon lamps; (4) hybrid xenon/mercury vapor lamps; and (5) low-pressure germicidal lamps. Medium-pressure mercury lamps that emit a wide range of wavelengths are by far the most important radiation sources for curing of coatings.

The UV dryer consists of lamp housing and reflector assembly. The reflector itself may be of either parabolic or elliptical geometry. Elliptical reflectors are most often used with medium-pressure mercury lamps. Parabolic reflectors provide a parallel beam of radiation, while the elliptical reflectors produce a focused beam of radiation on the substrate (Figure 40.19). UV-curable coatings are 100% solids and solvent removal equipment is not required.

While this type of UV-curing process does indeed have unique application, there are also hazards introduced, including biological effects on the skin and eyes, along with the generation of ozone, oxides of nitrogen (NO_x), and other by-products from the UV-curing process. In order to employ this type of light energy, all UV equipment must be designed to meet all safety requirements of personnel and environment.

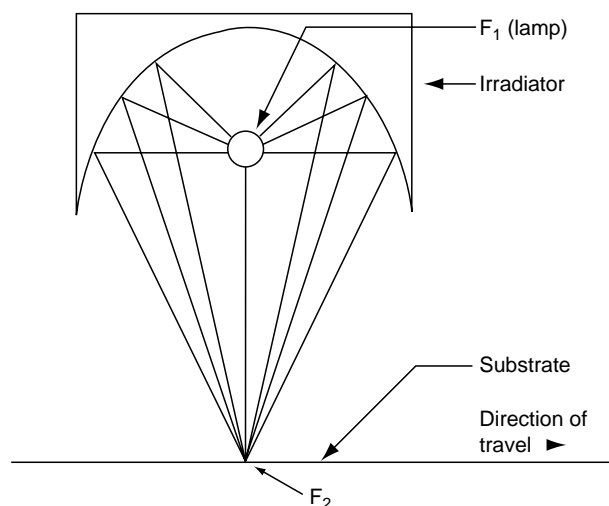


FIGURE 40.19 Elliptical reflector assembly.

40.3.8 ELECTRON BEAM CURERS/DRYERS

Earlier work in EB curing was done with relatively high-energy (150 to 1000 kV), scanned-beam-type equipment. The advent of the modern linear cathode unscanned accelerator has revolutionized the radiation curing/drying industry. This equipment generates a continuous “waterfall” of high-energy electrons. As a result, the greatly simplified design provides excellent reliability. Beam energy generated is in the range of 100 to 300 kV, allowing compact shielding, which is usually built integral to the accelerator itself.

Figure 40.20 shows a schematic of an electron processor. The electron processor is a stainless steel vacuum tube in which one or more longitudinal heated-filament electron “guns” are raised to a potential of several kilovolts. The electrons are then accelerated as they move from the filaments to the perimeter of the tube. A slot in the tube defining the width of the processing zone permits the accelerated electrons to reach a metallic foil window that is thin at ground potential. This foil window is thin enough to permit efficient electron transmission. The electrons emerge into a controlled processing environment, where they are absorbed by the products to cause curing of the coating. The accelerated or energetic electrons as carriers of energy are somewhat like bullets. The higher the voltage through which they are accelerated, the deeper they can penetrate into the products.

EB curing/drying systems have been successfully integrated into multicolor web offset printing presses for printing of liquid packaging and folding carton stock. Other commercial applications include

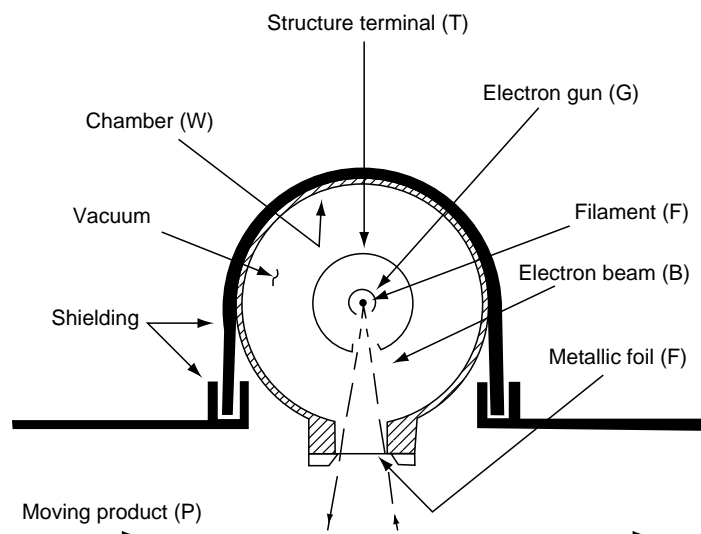


FIGURE 40.20 Schematic of electron processor.

EB-curing intaglio inks and EB-curing high-gloss overprint varnishes over heat-set inks in-line on roto-gravure presses.

EB processing is a very practical and widely used commercial process for carrying out chemical reactions such as cross-linking and curing sterilization on web materials. In most cases, it is more energy efficient than normal processing methods such as: (1) high-temperature chemical cross-linking; (2) hot air drying in the case of laminating or printing; and (3) high-temperature or chemical sterilization. The use of radiation curing for laminating adhesives and printing inks eliminates or minimizes the costly solvents. This in turn reduces or eliminates the need for an emission control system to meet environmental standards set by government agencies. In the case of curing or cross-linking, it generally provides a clear-cut reaction with no residual promoters or cross-linking agents that might cause degradation of the product or problems with governmental agency regulations on toxic residues in products for food and drug applications.

UV and EB curing may be used as practical, efficient, and safe methods to cure coatings and printing inks in the converting and graphic arts industries. The most apparent advantage is the cutting of production time by eliminating steps and the reduction of labor cost, making the end product more competitive. A second advantage is the special curing of unique products that cannot be made any other way as efficiently. Typical examples include static control of electric components, textured release surface coatings, graphics on food packaging (because of the very low extractables achievable), and curing of magnetic media binders (process control and durability).

With the increasing popularity of higher-intensity (120 W/cm or more) UV lamps and the development of curing formulations, especially in the high-speed wide-web applications, UV equipment, lower in cost is much easier to justify than EB equipment.

40.4 APPLICATION EXAMPLES

This section presents a few typical cases of flotation drying systems applied to coated papers and films because flotation dryers have been widely used in the converting and graphic arts industries. They have numerous advantages over other competing techniques.

Within the last 10 y, the use of IR heaters has been widely accepted for the drying of coated paper and paperboard. For example, when drying coating was applied to commercial- or publication-grade papers during the papermaking process, the technique used IR, flotation air, and steam-heated cylinders. The degree of drying with each mechanism varies depending on coating formulation and use. [Figure 40.21](#) shows a typical system that uses electric or gas-fired IR to heat the web to 66°C while continuing the drying to an 80% solid level by air flotation. The balance of the drying is conducted through the use of 4 to 6 steam-heated cylinders with an evaporation rate restriction of 20 kg/h/m². There are some differences in the opinion regarding the amount of IR, air, and contact drying necessary and whether or not gas-fired or electric IR is most beneficial. Often, IR is placed as close as possible to the coating source to enhance product quality as it relates to solids migration and adhesion of the coating to the paper web.

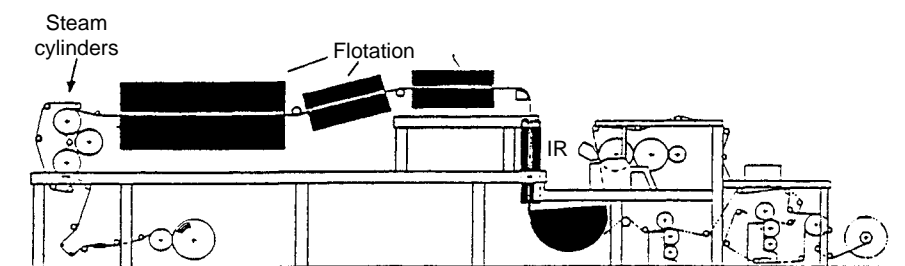


FIGURE 40.21 Paper coating with infrared radiation, flotation, and steam cylinders.

Drying of publication- and commercial-grade printed products is usually done with a system having an air velocity between 61 and 76 m/s at web speeds from 6 to 18 m/s. Web temperatures range from 93 to 177°C with multizone drying systems employed. The printing process typically includes both evaporation of hydrocarbon and solidification of a thermal plastic solid material; therefore, postevaporation conduction cooling is usually employed. Some special problems in this drying area include excessive shrinkage and damage to the paper product during the tortuous drying curve, environmental problems associated with the evaporated hydrocarbon, and special quality demands imposed by the consumers.

When drying adhesives, including pressure-sensitive adhesives, a dry coat weight of 20 g/m²/mil (1 mil = 40 μm) of coat weight thickness usually is employed. Solid contents range from 20 to 50%. Solvent removal rate limitations are at 18 kg/h/m². Drying systems are usually multizone with outlet velocity ranging from 20 to 40 m/s and air temperatures ranging from 100 to 160°C. Typically, a two- or three-zone drying system is employed.

Magnetic media webs typically range from 15 to 66 cm in width and employ a multizone (generally three) drying system. Impingement velocities are usually limited to 20 m/s in zone 1, 30 m/s in zone 2, and 40 m/s in zone 3, with operating temperatures ranging from 60 to 120°C. Evaporation rates are generally in the 64 to 80 kg/h/m² range. Airstream components in the dryers should be stainless steel with filters for maintaining a clean airstream. Varying substrate and coat weight thicknesses are employed depending on whether the magnetic media is audio, video, computer, or floppy disk. Solvents generally are tetrahydrofuran (THF), methyl ethyl ketone (MEK), toluene, methyl isobutyl ketone (MIBK), or cyclohexanone.

Carbonless coatings or encapsulated coatings would typically be used for carbonless carbon paper, cosmetic or fragrance samples, or other special applications. Coatings could range from 1.5 to 5 g/m² dry at a solid range of 20 to 40%. Solvent removal rates, assuming no damage to the capsules, would range

from 30 to 50 kg/h/m². Outlet velocities typically are 40 to 60 m/s with operating temperatures of 204 to 260°C.

Polyvinylidene chloride (PVDC) is typically used as a vapor barrier in food packaging. In these applications, evaporation rates are usually 10 to 15 kg/h/m² with air velocities of 20 to 40 m/s when utilizing paper substrates and 20 to 40 m/s when using plastic films. Dryer construction is usually of stainless steel because of the high corrosivity of the coatings and airstream.

Other applications include drying of aluminum printing plates, 100% solid silicone coatings, steel coils, and publication-grade paper products.

40.5 SAFETY ASPECTS

In the converting and graphic arts industries, the drying of coated webs involves the evaporation of solvents, water, and oils from the solvent-based coatings, adhesives, and printing inks into the air. These evaporated solvent vapors and ink oil mixing with dryer air will cause a fire and explosion hazard if an excessive amount of solvent vapors is accumulated in the dryer without monitoring. In addition, the exhaust gases combined with traces of carbon monoxide from the dryer burner due to incomplete combustion are toxic to humans, wildlife, and vegetation if emitted to the atmosphere without treatment. Furthermore, odor of the exhaust stream is objectionable.

The role of government agencies and insurance companies is becoming greater in safety, product formation, raw material selection, and all environmental affairs of all industries. In the United States, the federal Environmental Protection Agency (EPA) will gradually tighten up air pollution control by enforcing existing and new environmental regulations under the Clean Air Act, including the regulations of emissions from all adhesive, coating, and ink processes. Also, the Factory Mutual (FM) and Fire Insurance Association (FIA) will require more safety plants, equipment, and operations, e.g., all machines using a solvent-based or a water-based liquid with approximately 20% alcohol mixed with water must

be built to conform to explosion-proof regulations. Therefore, the design and operation of dryers for the solvent-based adhesive, coating, and ink printing processes are confined by two factors: (1) the explosion-proof standard required by MF and FIA and (2) the dryer emission control enforced by the EPA. These two factors are briefly discussed below.

40.5.1 THE EXPLOSION-PROOF STANDARD

It is of advantage to recirculate as much heated air inside the dryer as possible for the purpose of energy conservation. However, accumulation of an excessive amount of flammable solvent vapor will not only retard the evaporation rate of solvent but also create a dryer fire or explosion hazard. To avoid the potential fire or explosion hazard, the dryers must be designed to meet the explosion-proof standard that requires that the mixture of solvent vapors and air cannot burn if exposed to an open flame.

Solvent vapors cannot burn if the amount of air in the mixture is above or below the correct combustion ratio. If the amount of oxygen in the mixture is too small to support combustion due to a surplus of the fuel vapors, it is at the upper explosive limit (UEL) of the mixture. When the volume of the fuel in the mixture is too low to support combustion due to a lack of burnable vapors, it is at the lower explosive limit (LEL) of the mixture. It is necessary to ensure that these vapor–air mixtures are always below LEL or above UEL ratios.

40.5.1.1 Lower Explosive Limit Regulations

Conventionally, LELs are always used as the basis for making a dryer system safe. In the United States, LEL regulations stipulate an exhaust volume of 10,000 standard cubic feet (SCF) per U.S. gallon (gal) are required, while in Europe the regulations specify 90 standard cubic meters (SCM) per kilogram (kg) of solvent evaporated, if proper instrumentation such as the combustible gas analyzer or the IR analyzer for measuring percentage of the flammable solvent vapors in the dryer or exhausting air streams is used. All agencies permit a maximum solvent vapor concentration of up to 25% of LEL of the solvents being evaporated, providing the operator with a 300% safety factor before the mixture will reach combustible levels. However, if one specific solvent is used on a machine that does not have the LEL measuring system, the legal limit is 25% of LEL for that particular solvent rather than the standard of 10,000 SFM/U.S. gal or 90 SCM/kg.

The LEL regulations also allow the flammable solvent vapor concentration up to 50% LEL when

the LEL measuring system is provided. This effectively doubles the amount of solvent that can be evaporated in the same amount of air. However, the LEL measuring system must be properly calibrated for each specific solvent or combination of solvents, must be fitted with audible and visible alarms to warn the operators when the mixture approaches 50% LEL ratio, and most importantly, must automatically shut down the machine if the limit is exceeded.

40.5.1.2 Inert Gas Drying Process

Since the LEL regulations permit the conventional dryers to maintain the maximum flammable solvent concentration of 25% LEL, or up to 50% LEL if proper instrumentation for monitoring the solvent vapor concentration is applied, the LEL dryers allow quite limited recirculation, emit a large volume of heated air to the atmosphere, and introduce an equally large volume of makeup air. Alternately, if the oxygen concentration in the drying gases is decreased below 12%, which is too small to support combustion, then the solvent vapor concentration and the recirculation can be largely increased without maintaining LEL for avoiding explosion or fires. Instead of air, which contains approximately 21% oxygen, the inert gas (e.g., carbon dioxide or nitrogen), containing only a small amount of oxygen, can be used as the drying medium. This inert gas drying process allows an increase in recirculation because a flammable mixture cannot be developed in a low-oxygen atmosphere regardless of how high the solvent concentration is. [Table 40.1](#) shows maximum permissible oxygen percentage to prevent ignition of flammable gases and vapors using nitrogen and carbon dioxide for inerting.

[Figure 40.22](#) shows an example of a low-oxygen dryer with a solvent-recovery system, meeting National Fire Protection Association (NFPA) standards. As in the conventional case, the coated web enters and leaves the dryer chamber through the web openings. The dryer atmosphere consists of an inert carrier gas that is simultaneously recirculated through the dryer enclosure (line 1). A gas seal is provided around web openings to restrict air from entering the drying chamber and to prevent the inert gas from escaping. In actual operation, oxygen levels are monitored and controlled within the dryer to maintain <5% oxygen. This allows a higher solvent vapor concentration in the dryer environment without risking explosion and fires. A bleed stream (line 2) is processed by a solvent-recovery system. A coolant (line 3) acts to condense the solvent, which is discharged to storage (line 4). With solvent removed, the gas stream is returned to the dryer instead of being discharged to the atmosphere.

TABLE 40.1
Oxygen Percentages Relating to Ignition

	N ₂ –Air		CO ₂ –Air	
	O ₂ Percent above Which Ignition Can Take Place	Maximum Recommended O ₂ Percent	O ₂ Percent above Which Ignition Can Take Place	Maximum Recommended O ₂ Percent
Acetone	13.5	11	15.5	12.5
Benzene (Benzol)	11	9	14	11
Butadiene	10	8	13	10.5
Butane	12	9.5	14.5	11.5
Butene-1	11.5	9	14	11
Carbon disulfide	5	4	8	6.5
Carbon monoxide	5.5	4.5	6	5
Cyclopropane	11.5	9	14	11
Dimethylbutane	12	9.5	14.5	11.5
Ethane	11	9	13.5	11.0
Ether	—	—	13	10.5
Ether (diethyl)	10.5	8.5	13	10.5
Ethyl alcohol	10.5	8.5	13	10.5
Ethylene	10	8	11.5	9
Gasoline	11.5	9	14	11
Gasoline				
73–100 Octane	12	9.5	15	12
100–130 Octane	12	9.5	15	12
115–145 Octane	12	9.5	14.5	11.5
Hexane	12	9.5	14.5	11.5
Hydrogen	5	4	6	5
Hydrogen sulfide	7.5	6	11.5	9
Isobutane	12	9.5	15	12
Isopentane	12	9.5	14.5	11.5
JP-1 fuel	10.5	8.5	14	11
JP-3 fuel	12	9.5	14	11
JP-4 fuel	11.5	9	14	11
Kerosene	11	9	14	11
Methane	12	9	14.5	11.5
Methyl alcohol	10	8	13.5	11
Natural gas (Pittsburgh)	12	9.5	14	11
Neopentane	12.5	10	15	12
<i>n</i> -Heptane	11.5	9	14	11
Pentane	11.5	9	14.5	11.5
Propane	11.5	9	14	11
Propylene	11.5	9	14	11

40.5.2 THE DRYER EMISSION CONTROL

The evaporated solvents, ink oils, and toxic gases such as carbon monoxide in the dryer exhaust are mixtures of hydrocarbons, i.e., carbon bonded to each other and/or to hydrogen in various lengths and configurations. The general term for these materials is volatile organic compounds (VOCs). Regulations for the total reduction of VOC emissions is a combination of destruction efficiency and capture efficiency. The United States is the only country

following a path to cleaner air. Most often, 90, 95, or 99% destruction efficiency will be seen, depending on whether it is the local, state, or federal level. Germany and the Scandinavian countries already have regulations limiting emissions from a source to 50 mg/Nm³ (normal cubic meter) in many areas. Some require as low as 20 mg/Nm³ depending on the type of VOCs. Many other industrialized countries are following this lead.

Basically, there are two methods of solvent removal from the dryer exhaust: (1) solvent recovery

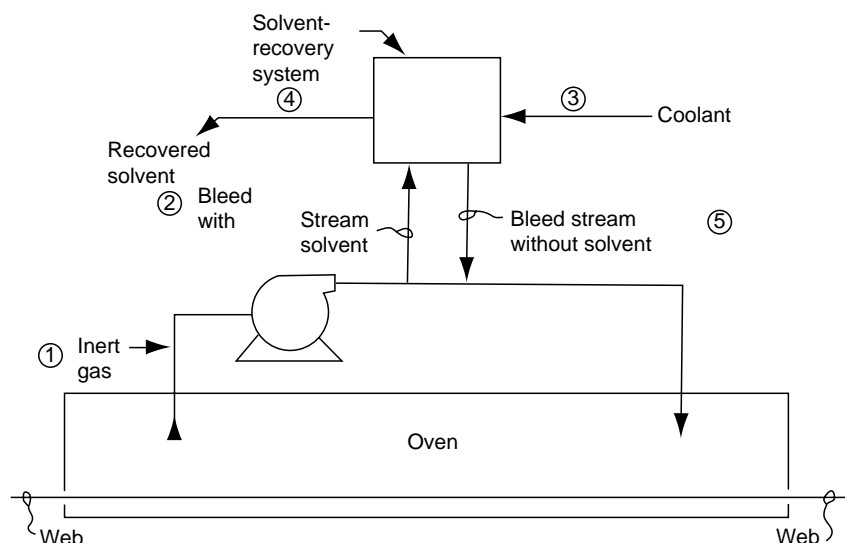


FIGURE 40.22 Inert gas dryer layout.

and (2) solvent incineration. Solvent recovery requires a high solvent concentration operation or the use of a very expensive solvent to be economically worthwhile, while solvent incineration is mostly used for the dryer effluent around or below 25% LEL.

40.5.2.1 Solvent Recovery

Solvent vapor can be recovered by adsorption on activated carbon. The exhaust stream is passed through an activated carbon bed and the hydrocarbon vapors are adsorbed. When the carbon bed becomes saturated with vapor, the exhaust stream is directed to another fresh bed and the saturated bed is recovered by stripping the solvent by either steam or vacuum. Steam stripping is the often-used method for most coating operations. The solvent is removed along with the condensate and it is separated by decantation if the solvent is water insoluble or by distillation if it is soluble.

The efficiency of activated carbon adsorption units is 90 to 99%, depending on the specific solvent vapor used and the design of the carbon bed. If the gas adsorption system is equipped with an IR analyzer for monitoring the solvent vapor concentration, the discharge effluent can be automatically switched from one saturated bed to another fresh bed by the IR sensing device, and the recovery process is started.

Solvent vapor can also be recovered by using an inert gas (e.g., nitrogen) dryer atmosphere and gas condensation. Essentially, the recovery unit attached to an inert gas dryer (Figure 40.22) contains a heat exchanger, a condenser, and a separator. The dryer exhaust stream is passed through a heat exchanger

and a condenser that cools the exhaust, causing the solvent vapors to condense. The liquid solvents flow to a solvent separator and are pumped into solvent collection tanks. The cleaned nitrogen then returns to the dryer to pick up more solvent vapors. Heat removed from the exhaust stream by the heat exchanger preheats the cleaned nitrogen before it returns to the dryer.

40.5.2.2 Incineration

Incinerators are designed to heat the dryer exhaust for proper time and temperature required to decompose hydrocarbons into carbon dioxide and water through the process of oxidation. Temperature, time, and turbulence are the three factors contributing to the conversion efficiency. Thermal and catalytic incinerators are the two major incinerators in use. Thermal incinerations are used for the solvent vapor concentration around or below 25% LEL, while catalytic incinerators should be considered for applications in which there is a low concentration (below 15% LEL).

Thermal incinerators process dryer discharges by burning them at high temperatures (between 760 and 800°C). They are designed with 0.5 s or greater total residence time, although hydrocarbons generally oxidize in 0.3 s. Residence time that the exhaust stream is maintained at the oxidation temperature is critical not only for proper oxidation but also for proper mixing. Turbulence assures proper mixing of the airstream. Improper or unsatisfactory mixing will require higher temperatures and/or longer residence time to achieve proper oxidation.

Figure 40.23 shows a schematic diagram of a thermal incinerator containing two basic elements, a heat exchanger and a burner (combustion chamber). A heat exchanger is designed to transfer thermal energy from one airstream to the other for the purpose of reducing the burner input needed for reaching oxidation temperature. Typical VOC reduction of thermal incinerators is in the range of 99%, but the cleanup rates can be greater than 99%, depending on the design and the operating conditions (e.g., operating temperature, residence time, and turbulence) to obtain good mixing.

Carbon monoxide can be either created by partial incineration of ink oils or destroyed by complete incineration of carbon monoxide to carbon dioxide in a thermal incinerator, depending on the operating temperature. Carbon monoxide production tends to rise with increasing temperature until it reaches a maximum at about 650°C. Then, the carbon monoxide content tends to decrease with increasing temperature until it reaches a negligible level at about 760°C.

Burners create NO_x (e.g., NO and NO_2) at high temperatures. As the temperature of the incinerator rises, NO_x concentration in its exhaust will increase. Unfortunately, VOCs and NO_x in the atmosphere, in the presence of sunlight, enter into a chemical reaction that produces ozone (O_3). Ozone is the substance measured as the criterion for “smog alerts” in cities. Even extremely low levels of ozone in the air can cause significant respiratory difficulties in a sizable portion of the population.

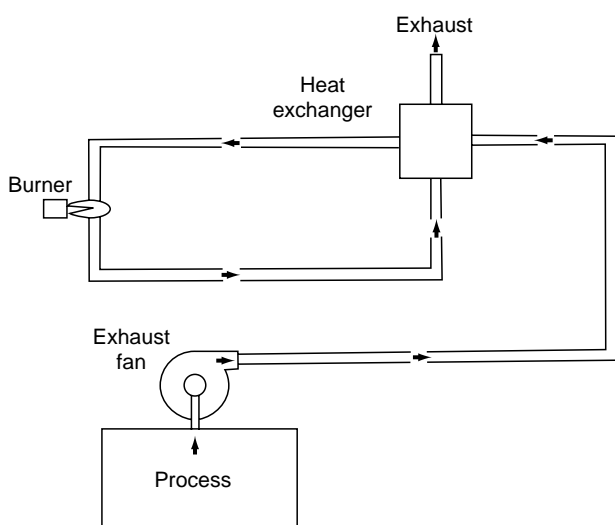


FIGURE 40.23 Thermal incinerator schematic.

Thermal incinerators typically use two major types of heat exchangers: (1) envelope type and (2) shell and tube exchangers (Figure 40.24). A regenerative system is a special type of thermal incinerator that utilizes a ceramic stoneware as the heat exchanger medium and operates at very high temperatures (above 800°C). The ceramic stoneware is mounted in at least three columns for better automatic control purposes. Figure 40.25 shows a two-column regenerative system for discussion purposes.

The process airstream is passed through the stoneware as it enters and exits the central incineration chamber. By constantly cycling the airstream between two columns, the incoming airstream is heated by stoneware, which in the previous cycle was absorbing heat from the airstream exiting the central chamber. As this column loses heat to the incoming airstream, it cycles and becomes the receptor of heat, repeating the cycle. The cleanup rates are greater than 90%. Regenerative systems require high capital/installation costs, but their operating cost is low if run continuously due to large stoneware mass.

Catalytic incinerators are an alternative to thermal incinerators as a means of oxidizing exhaust hydrocarbons into carbon dioxide and water at lower temperatures (315 to 480°C). The contact time with the catalyst bed is about 0.3 s. Catalytic incinerators typically destroy 90 to 95% of VOC, depending on the design of the unit and catalyst activity. Figure 40.26 shows a schematic diagram of a catalytic unit. The basic elements of the catalytic units are a preheating/mixing section, designed to achieve a uniformly preheated and distributed exhaust stream flow, and the catalyst bed and catalytic matrix, where the major portion of the oxidation reactions take place.

Two types of catalyst systems are commonly used: (1) packed beds and (2) monolith blocks. The first type is a perforated container containing spherical beads coated with catalytic ingredients; it allows the reactants to flow through the catalytic beds. It has high external surface area and high mass transfer efficiency between reactants and catalyst. The second type has the walls of monolithic honeycomb wash coated with the catalyst ingredients and allows the reactants to flow through the channels in the monolith.

Catalytic systems are limited to applications in which the exhaust stream has the lower particulate loading and/or the exhaust stream has negligible “poisons.” The poisons are primarily silicon and phosphorus that coat the catalyst. Halogens such as chlorine and sulfur can decrease the effectiveness of the catalyst.

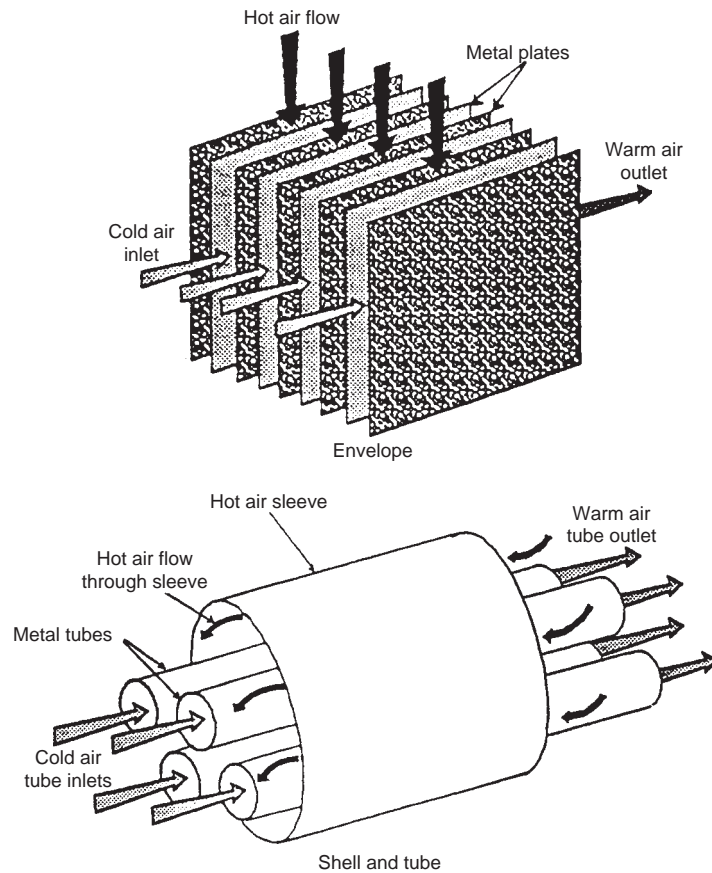


FIGURE 40.24 Heat exchanger types.

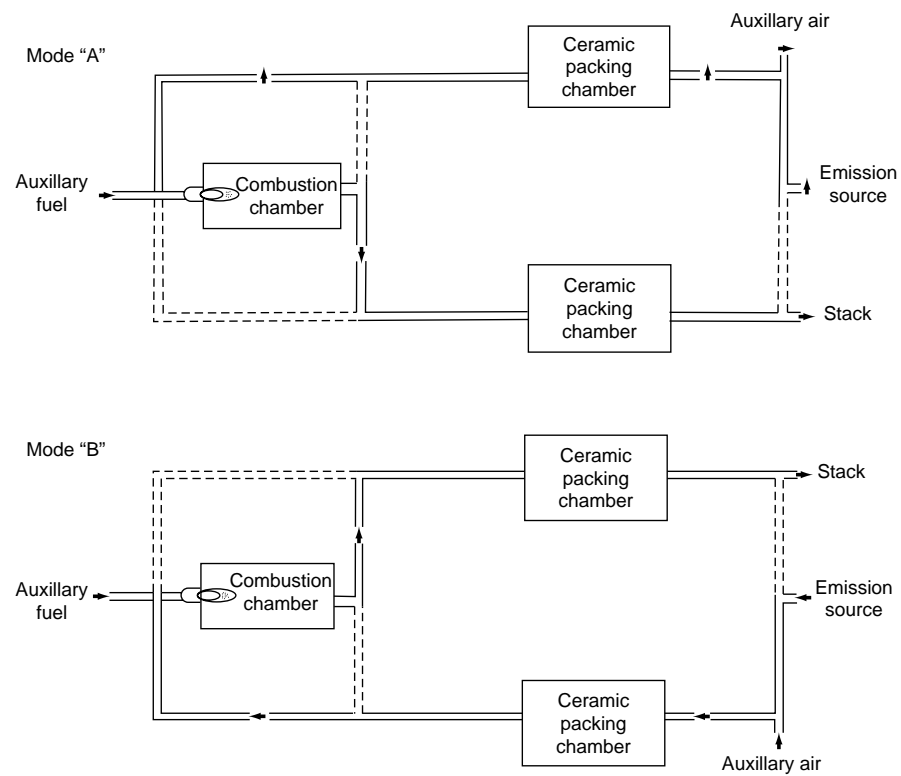


FIGURE 40.25 Regenerative system of thermal incinerator.

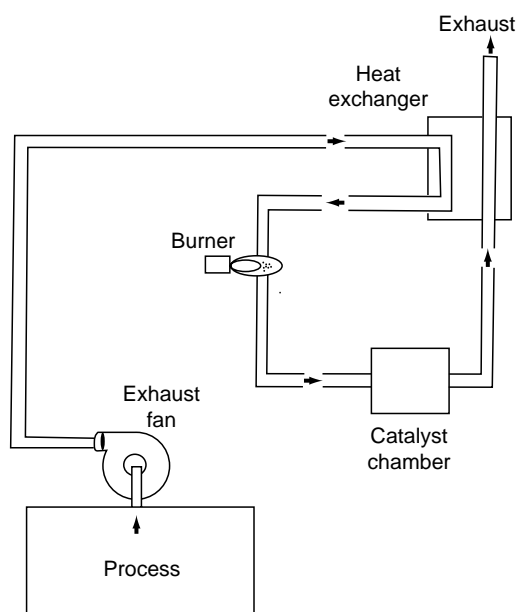


FIGURE 40.26 Catalytic incinerator.

40.6 CONCLUSION

Various commercially employed drying systems for coated webs are summarized. The reader is referred to the cited references in the bibliography for further details.

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41 Drying of Polymers

Arun S. Mujumdar and Mainul Hasan

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41.1 INTRODUCTION

Spurred by continually escalating energy costs, along with the advent of new competitive polymers accompanied by new and extended applications of polymers and plastics, interest in the energy-intensive operation of drying of polymers has been on the rise in recent years.

Drying is one of the prime polymer recovery operations performed before transfer to the compounding plant or packaging for direct use. It is the part of the process in which the polymer is handled essentially as a solid and liquid and gas streams become relatively minor. In polymer production, other recovery operations include salvation of the unreacted monomer and solvent, coagulation and precipitation, concentration and devolatilization, and liquid–solid separation [1].

Although drying is the oldest and most commonly encountered of all unit operations of chemical engineering, it is one of the most complex and least understood operations. One of the prime reasons for this state of affairs is the enormous diversity of drying equipment; over 100 clearly identifiable different types of dryers are in commercial use around the world. Depending upon the nature of the processing mode, physical state of the feed, mode of heat and mass transport, operating temperature and pressure, and other factors, one can classify existing dryers into so many different types that it is impossible to develop or hope to develop generalized procedures for analysis of all types of dryers [2].

This chapter provides a few guidelines for the selection of polymer dryers and discusses the alternatives available. Picking the best dryer for a specific polymer application is beyond the scope of this chapter because of many variables involved in such selection. Since most of the work in this area is proprietary in nature with very little information available in the open literature, it is therefore believed that this chapter will help to overcome the initial agonies of a polymer engineer in selecting dryers for water- or solvent-wet granular polymer particles.

For better understanding of the subject for a nonspecialist in polymer technology, a brief survey

is given about the various polymerization techniques employed commercially before discussion of the drying equipment.

41.2 COMMON POLYMERIZATION PROCESSES

The selection of dryers for polymer drying depends to a large extent on the upstream operations, e.g., polymerization, since the behavior of a polymerization reaction and the properties of the resulting polymer can vary greatly according to the nature of the physical system in which the polymerization reaction is carried out. Several processes are used commercially to prepare polymers. Each process has its advantages, usually depending on the type and final use of the polymer. The following types of polymerization processes based on physical systems are considered briefly in this section: (1) bulk; (2) solution; (3) suspension; (4) emulsion; and (5) gas-phase polymerization [3].

41.2.1 BULK POLYMERIZATION

The polymerization of the pure monomer without diluent is called *bulk polymerization* or *mass polymerization*. The monomer (e.g., styrene, vinyl chloride (VC), vinyl acetate (VA), acrylic esters, butadiene, or acrylonitrile) is first purified to remove oxygen or other inhibitors (by bubbling nitrogen through it, by distillation, or by evacuation) and then the polymerization is started through heating, ultraviolet (UV) radiation, or the addition of an initiator (e.g., peroxides, azo compounds, and others). Usually, after a short period of heating, the reaction mixture continues to heat by itself, and therefore it is necessary to remove the heat by cooling. With increasing conversion, because of the rapidly increasing viscosity of the polymer–monomer mixture, this becomes more and more difficult. With large amounts of monomer, bulk polymerization often takes very turbulent and even explosive form as a result of the rapidly increasing temperature. The violence of the reaction is even further increased by the increase in the radical concentration that occurs with increasing viscosity.

Because of the difficulty in heat removal, bulk polymerization is only carried out in a few cases. However, in the cases in which it is used, it is done on a very large-scale, e.g., the bulk polymerization of styrene or ethylene (high-pressure process). Since in this type of polymerization the possibility of chain transfer is relatively small and because of self-acceleration, other polymers with high molecular weights are found. One of the characteristics of this process that is always a technical advantage is the great purity of the polymer resulting from the lack of additives during polymerization.

41.2.2 SOLUTION POLYMERIZATION

For solution polymerization, a solvent inert to the monomer is used to control the polymerization. High exothermicity is limited by dilution, causing the reaction rate to be slowed owing to solvent addition. The solvent is recycled after cooling and is sent back to the polymerization reactor. The concentration of the solvent is chosen in such a way that the polymerization mixture can still be stirred after complete conversion.

Solution polymerization has been employed almost exclusively in cases in which the polymer is then used in the form of solutions (50 to 60%) for lacquers, adhesives, impregnation materials, and other products. Obtaining the pure polymer by distilling off the solvent is complicated because the hard polymer cannot be taken out of the vessel after evaporation of the solvent. Through construction of extruders with vacuum distillation zones and by using other special evaporators, it is possible to separate the polymer from the solvent.

In this process the choice of the solvent's chain transfer constant is very important because this influences the molecular weight to a considerable extent. Because of chain transfer with the solvent and because of the lower monomer concentration, the molecular weight of polymers prepared by solution polymerization is usually lower than that of the corresponding bulk polymers.

Commercially, solution polymerizations are not carried out to high conversions (near 100%) but continuously at a constant monomer concentration. The unreacted and evaporated monomer is recycled together with the solvent. This type of production process has two advantages. The reactor always works in a range of high polymerization rates, and the molecular weight distribution curve is not so broad as it is with polymers produced in a discontinuous process with high conversions.

41.2.3 SUSPENSION POLYMERIZATION

In the suspension polymerization process, water is used to control heat generation. A catalyst is dis-

solved in the monomer, which is dispersed in water. A dispersing agent is incorporated to stabilize the suspension formed. For any nonpolar monomers, this method offers a method of eliminating many of the problems encountered in bulk and in solution polymerization, especially the heat dissipation problem in the former and solvent reactivity and removal of the latter.

Another attractive feature of large batch preparations is that the polymeric products obtained from a suspension polymerization, if correctly carried out, are in the form of finely granulated beads that are easily filtered and dried.

On a technical scale, suspension polymerization is used in the production of polyvinylchloride, polystyrene, polymethyl methacrylate, and others. For the production of rubbery, sticky, polymers (e.g., the polyacrylates), this is less suitable.

41.2.4 EMULSION POLYMERIZATION

The emulsion polymerization process is similar to suspension polymerization. This process is also carried out in a water medium. An emulsifier, either anionic soap or cationic soap, is added to break the monomer into very small particles. The initiator is in solution in the water. After polymerization, the polymer can be precipitated, washed, and dried, or the mixture can be used directly (e.g., latex paint).

Emulsion polymerization is superficially related to suspension polymerization, but the kinetic relationships are entirely different. The major causes of the differences are: first, the monomer droplets in the latter system are approximately 0.1 to 1 mm in size and the particles in the former are approximately 10^{-7} to 10^{-6} mm in size; and second, the catalyst is dissolved in the aqueous phase in the latter but is incorporated directly into the droplets in the former.

In all cases in which the presence of an emulsifier is not disturbing, emulsion polymerization is advantageous. In comparison with other polymerization techniques, it has the following advantages: (1) the polymerization heat can be removed very easily and (2) the viscosity of the lattices, even with high concentration (up to 60%), is low in comparison with corresponding solutions.

One large-scale use of this process is in the production of synthetic rubber and, on a small-scale, in the production of polyvinyl chloride (PVC) and polystyrene. The other large-scale use is in the production of plastics dispersions used as such (without first coagulating them) for the production of paints, pigments, inks, coatings, and adhesive paste (e.g., polyvinyl acetate, polyvinyl propionate, and polyacrylic ester dispersions).

41.2.5 GAS-PHASE POLYMERIZATION

The term *gas-phase polymerization* is a misnomer in that it refers only to a polymerization reaction initiated on monomer vapors, generally by photochemical means. High-molecular weight polymer particles are not volatile, so a fog of polymer particles containing growing chains quickly form and the major portion of the polymerization reaction occurs in the condensed state.

High-uniformity polyethylene (PE) can be manufactured by passing gaseous ethylene through an active chromium-containing catalyst bed. Other monomers that have been polymerized successfully in the gas phase include methyl methacrylate, VA, and methyl vinyl ketone [4].

41.3 DRYER CLASSIFICATION [5–7]

41.3.1 CLASSIFICATION BY MODE OF HEAT TRANSFER

41.3.1.1 Indirect Dryers

Indirect dryers, also called nonadiabatic units, separate the heat transfer medium from the product to be dried by a metal wall. These dryers are subdivided on the basis of heat applied by radiation or through heat transfer surface and also by the method in which volatile vapors are removed.

Heat transfer fluids may be of either the condensing type (e.g., steam and diphenyl fluids, such as Dowtherm A) or the liquid type (e.g., hot water and glycol solutions). Because of low film coefficients of the noncondensing gaseous system, it is seldom used as the heating medium.

Indirect dryers have several distinctive operating features: (1) the risk of cross-contamination is avoided since the product does not contact the heating medium; (2) since a limited amount of gas is encountered, solvent recovery is easier than with an adiabatic dryer; (3) dusting is minimized because of the small volume of vapors involved in indirect drying; (4) dryers allow operation under vacuum or in closely controlled atmospheres that can avoid product degradation; and (5) explosion hazards are easier to control.

Typically, indirect dryers are used for small- or medium-size production. The product from such a unit has a higher bulk density than the same material processed in direct dryers. Particle size degradation usually can be minimized by proper selection of agitator speed or design. The common indirect heated dryers are tubular dryers (with or without vacuum), drum dryers (atmospheric, vacuum, horizontal or rotary vacuum, and others), hollow disk dryers, paddle

dryers, mechanically fluidized bed dryers, pneumatically conveyed dryers, cone or twin-shell dryers, and others.

41.3.1.2 Direct Dryers

Direct dryers or adiabatic or convective dryers transfer heat by direct contact of the product with the hot gases. The gases transfer sensible heat to provide the heat of vaporization of the liquid present in the solid.

Direct dryers may use air, inert gas, superheated vapor, or products of combustion as the heating medium. Combustion gases are seldom used in polymer drying because of possible product contamination. Inert gas eliminates the explosion and fire hazard and may be desirable to prevent oxidation of polymers prior to the introduction of stabilizers. Use of superheated vapor as a heat carrier is highly desirable when solvent is vaporized in the dryer and has to be recovered.

Commonly used direct dryers in polymer plants are rotary warm air, fluidized bed, flash, spray, tunnel, and various vibrating and spouted bed (SB) types. All these have a common disadvantage. The amount of air or hot gas required is fairly large, which causes the auxiliary equipment needed (e.g., air heaters, blowers, and dust collectors) to be sized accordingly; the thermal efficiency is also lower than that of indirect dryers.

Although this classification of dryers has some importance, it is quite difficult to apply it in more than a general way. Both types of dryers are commonly used in polymer-drying processes. Often a combination of direct and indirect drying is economically the most efficient solution to some polymer-drying problems.

41.3.2 CLASSIFICATION BY RESIDENCE TIME

The pressing need of product quality in the plastics industry also forces one to consider residence time distribution of the product when comparing dryers.

41.3.2.1 Short Residence Time

The short residence time category comprises spray dryers, pneumatic dryers, and thin-film dryers in which the residence time may be of the order of several seconds.

41.3.2.2 Medium Residence Time

Continuous fluid-bed dryers, steam tube rotary dryers, and rotary dryers can be designed to provide medium residence time (of the order of minutes).

41.3.2.3 Long Residence Time

Rotary dryers, batch fluid dryers, continuous or batch tray dryers, hopper dryers, multispouted bed, and vacuum tumble dryers are typical long-residence units used in polymer drying.

41.3.3 OTHER CONSIDERATIONS

On the basis of the polymerization alone, it is difficult to specify definite dryer selection rules since typically polymer properties differ over a wide range. The choice of dryer is also limited by the physical properties of the polymers, e.g., polymer-handling characteristics, individual or closely related drying curves, properties of the emitted volatiles, limitations on temperature, and particle size and distribution requirements. Other factors include equipment space limitations, production rates, pollution control requirements, solvent recovery, thermal sensitivity, and product quality specifications.

The primary step in specifying a dryer is to define the physical, thermal, and chemical properties of the product and the volatiles present. Often the consistency of the feed reduces the choice of dryer. A few guidelines are always helpful in selecting polymer dryers. For example, if a solvent must be evaporated and then recovered, it is usually not desirable to choose a convection dryer. Since solvent must be condensed from a large carrier gas flow, the condenser and other equipment become rather large. If the maximum product temperature is lower than about 30°C, it is possible to specify a vacuum dryer. If the average particle size is about 0.1 mm or larger, a fluidized bed dryer may be considered, or if the feed is a slurry or paste a spray dryer may be a judicious choice. Scaling is another important factor that can dictate dryer selection. For example, if the requirements are to produce high tonnage of a polymer in one line, it probably would be advantageous to consider a fluid-bed dryer rather than a mechanical rotating type.

The reader is referred to other chapters of this handbook for details concerning specific dryers discussed.

41.3.4 COMMON POLYMER DRYERS

41.3.4.1 Rotary Dryers

Historically, rotary dryers (RDs) have been the most popular in polymer-drying operations. A rotary dryer consists of a slowly revolving drum (often fitted with internal flights or lifters) through which both the material and the gas pass. Gas, cocurrent or counter-current with the granular polymer, can be introduced at either end of the cylindrical shell.

Solids move through the dryer by the effect of gravity, the rotation of the cylinder, and gas flow (in the case of cocurrent units). Internal scoops, blades, and lifters, which give the solids a showering pattern, are provided for better gas–solid contact. Baffles and dam rings are also available to retard the forward motion of the solids and to increase residence time (5 to 20 min is common; much larger times are also found in drying of certain polymer pellets).

An improvement over the standard rotary dryer is the steam tube rotary dryer. Here, two or three rows of steam tubes are located in concentric circles within the shell, which extend the full length of the cylinder. The tubes together with a series of small radial flights serve to agitate the material for uniform drying. These types of dryers were used in the polymer industries for heat-sensitive polymers requiring indirect heating.

With the advent of the new and energy-efficient dryers, rotary dryers nowadays are seldom used in polymer drying in new polymer plants. Modified fluid-bed dryers as well as novel spouted bed dryers can replace rotary dryers in many applications.

41.3.4.2 Flash Dryers

The flash dryer (FD) is a direct-type, cocurrent unit that is essentially a long vertical tube with no moving parts. In polymer drying this is mostly used as a predryer to remove surface moisture.

In FD units, hot inlet gases contact the wet product, which may be powdery, granular, crystalline, or pasty material, as discharged from a centrifuge or filter press. Providing a short residence time of several seconds, FD is well suited for high evaporative loads. Drying is nearly adiabatic, an advantage with heat-sensitive polymers. High mass and heat transfer rates are obtained because of the high relative velocity between feed and inlet gas and a large exposed product surface area.

The method of feeding wet polymer to FD is very important. Granular products are relatively free flowing when wet polymers are fed with devices such as screw and rotary star feeders; sticky polymers may be best handled with a table feeder. Lumpy or pasty polymer must be broken up or mixed with dry product recycle to produce a more uniform and free-flowing feedstock.

Among the developments in flash drying, the first and the simplest is the “thermo venturi” drying concept in which a vertical drying column expands so that coarse particles remain suspended while drying and finer particles travel straight through with the drying air. This is quite effective as long as the particles are relatively spherical and the size spread is not

too great. Similar designs feature “bicones” in which the drying column expands and contracts, possibly with the addition of supplementary hot air, often injected tangentially.

Recent improvements in flash drying include the ring dryer. The heart of this dryer is a centrifugal separator. It combines renewal of the drying air with centrifugal classification. The lightest and finest fractions of the product are passed with the spent drying medium into the product collection system; oversize, partially dried material is held in circulation. The split is varied by adjusting the positions of suitable deflectioners, introduced in the flow loop. This type of FD is available in both multistage and closed-circuit designs with both direct and indirect heating options for removing both surface and bound moistures, as well as solvent removal and recovery.

41.3.4.3 Spray Dryers

In spray dryers, the feed material, in the form of a solution, suspension, slurry, or paste, is sprayed in a high-temperature gas zone by centrifugal disks or pressure nozzles. Such dryers are used in polymer industries in which the polymers cannot be separated mechanically from the carrier liquid, e.g., emulsion-polymerized PVC.

In polymer industries, wherever spray dryers are used they are primarily used as predryers of a multistage system. Final drying is normally done in a fluid bed, which is either stationary or vibrated type. Stationary fluid beds are used when spray-dried powder leaving the drying chamber is directly fluidizable. The vibrated type of fluid bed is used for products that, on leaving the spray dryer, are not readily in a fluidizable state owing to their particle form, size distribution, or wetness.

In such a multistage system, the higher moisture content powder leaving the spray-drying chamber is transferred to the second stage, which is a fluid bed for completion of drying. The higher inlet temperature and lower outlet temperature operation in such a system give improved dryer thermal efficiency and increased dryer capacity without product quality degradation.

41.3.4.4 Fluidized Bed Dryers

Fluidized bed dryers (FBDs) involve the suspension of solid particles in an upwardly moving stream of gas, which is introduced through a distribution plate that may be cooled for heat-sensitive polymers. Such a dryer may operate batchwise.

The advantages offered by FBDs are: (1) the even flow of fluidized particles permits continuous,

automatically controlled, large-scale operation with easy handling of feed and product; (2) no mechanical moving parts, i.e., low maintenance; (3) high heat and mass transfer rates between gas and particles—this is well mixed, which also avoids overheating of the particles; (4) heat transfer rates between fluidized bed and immersed objects, e.g., heating panels, are high; and (5) mixing of solids is rapid and causes nearly isothermal conditions throughout the bed, thereby facilitating easy and reliable control of the drying process.

Using the solvent being removed as the heat carrier and fluidizing medium (i.e., a superheated vapor) has proved a feasible and beneficial design. Its advantages include: (1) reduction in size of condensing and recovery equipment; (2) increase in drying rate due to the elimination of the gas-film resistance of the foreign vapor; (3) volumetric heat capacity of various vapors is usually greater than that of air; and (4) space velocity for fluidization is lower than with air, which reduces the volumetric vapor flow and consequently the size of the dust collector, air moving equipment, and other parts.

Drying of polystyrene beads is a typical example for industrial use of these dryers because of the close range of bead particle size. Also, the size of the beads permits high fluidizing velocities and therefore economic dryer sizes.

In recent years, indirect-heated fluidized beds have made inroads in almost all industries. Some of their advantages over the direct-heated FBD are: (1) the indirect heat transfer rate significantly reduces gas flow requirements; (2) there is tremendous leverage gained by the multiple of the heat transfer coefficient, LMTD, and heat transfer surface density permits very high heat inputs into low-temperature, heat-sensitive applications; (3) when a plug-flow, rectangular indirect fluid bed or low bed height is used, the solids flow counter to the thermal fluid, behaving like a countercurrent heat exchanger with all its attendant benefits; and (4) since the heat source is decoupled from the fluidizing gas source, vessel diameters and pollution-control equipment are much smaller.

Indirect fluid beds have already proved efficient in drying very heat-sensitive polymers with large constant-rate drying periods, as in drying PVC, polyethylene, acrylonitrile-butadiene-styrene (ABS) copolymers, and polycarbonates (PC).

41.3.4.5 Vibrated Fluidized Beds

A vibrated fluidized bed (VFB) is basically a long rectangular trough vibrated at a frequency of 5 to 25 Hz with a half amplitude of a few millimeters (2 to 5 mm). This kind of dryer can be used for drying

wet, sticky, and granular media and has been used successfully for drying polymers. It is often used as a second-stage dryer after a flash or spray dryer in many polymer-drying applications.

Benefits achieved from such dryers are: (1) uniform residence time distribution regardless of particle size; (2) ability to handle polydisperse solids; (3) ability to operate at low aeration rates and hence lower pressure drops; (4) gentle handling of product; (5) higher heat transfer and drying rates, and others. Since the equipment is mounted on resonance springs, the power consumption for vibration is minimal for well-designed VFB [6].

The vibration vector is typically applied at a small angle to the vertical to permit conveying of the solids in the long direction at the desired rate. This permits control of residence times and also better control of the drying or heating rates as the material progresses downstream.

41.3.4.6 Contact Fluid-Bed Dryers

Contact FB units are characterized by the residence time distribution of the individual particles inside the unit. A broad residence time distribution is obtained in a back-mixed FB in which the length/width ratio of the bed is relatively small. The narrow residence time distribution is obtained in a plug-flow FB in which the length/width ratio of FB is very large. This corresponds to a long, narrow FB. Alternatively, this can be obtained by compartmentalizing FB and is the usual practice followed in the industry.

Compared with the plug-flow FB, a back-mixed FB has a significant advantage inasmuch as the back-mixed FB can accept a feed material that is not readily fluidizable. This is possible owing to the vigorous mixing inside FB and that the material inside the bed acts as a large reservoir in which incoming feed material will be dispersed and the surface moisture will be flashed off, making the product fluidizable. This characteristic makes the back-mixed FB concept well suited as the predrying stage in many polymer-drying systems.

The plug-flow FB drying concept is particularly suitable for drying bound moisture from heat-sensitive materials since the residence time is controlled within narrow limits. In the typical polymer application, this means that the bound moisture can be removed from the polymer product at the lowest possible product temperature.

These two concepts, along with heating of the bed indirectly by immersed heat exchange surfaces, are jointly utilized in contact FBD. In FB applications for polymers with indirect heating, the temperature of the heating panels is typically limited by the softening point of the polymer.

41.3.4.7 Paddle Dryers

The paddle type of dryer, marketed by Nara Machinery Company of Japan, is an indirect dryer for granular or powdery material that dries such materials by bringing them into contact with revolving, cuneiform hollow heaters (paddles) without using gas as a heating medium. The paddles revolve at a low speed (10 to 40 rpm) inside the grooved trough fitted with a jacket (Figure 41.1).

The heating medium passes inside the hollow paddle so that the entire surface of the paddles and shafts acts as the heat transfer surface. The cuneiform blade enhances agitation of the material and at the same time prevents the powder from adhering to the heat transfer surface. For greatest heating efficiency the dryer is tilted slightly in the direction of product flow and is designed so that the material contacts all heated surfaces, both front and back. The wet product is fed continuously at the top of the dryer at one end. As the powder is agitated slowly by the heated rotating paddles, the moisture generated is conveyed out by a flow of hot air or other gas.

The main features of the paddle dryer are: (1) it is compact; (2) has high heat transfer coefficient and good thermal efficiency; (3) the paddles have an interplay for self-cleaning; (4) it is easy to control; and (5) a small amount of gas is required that minimizes dusting and other problems.

Paddle dryers have been successfully used in drying such polymers as VC resin, nylon pellets, and polypropylene (PP), as well as polyethylene. Operated in a closed-cycle mode they can recover organics from such solvent-laden products as polyethylene or PP and can reduce the air volume requirement to only 5 to 10% of that used in direct dryers.

Energy requirements for such dryers are also lower. It is seen that 1300 to 1500 Btu is required to dry 1 lb of moisture with the paddle dryer compared with 3000 Btu/lb for a suspended air unit. Because of the smaller air volume needed, the sizes of downstream condensers and refrigeration system units are reduced.

41.3.4.8 Plate Dryer

The plate dryer (PD) is an indirect dryer in which heat transfer is accomplished by conduction between the heated plate surface and the product. It comes under three major variations, e.g., atmospheric, gas tight, and vacuum.

In these dryers, the product to be dried is metered and continuously fed onto the top plate. A vertical rotating shaft provided with radial arms and self-aligning plows conveys the product in a spiral pattern

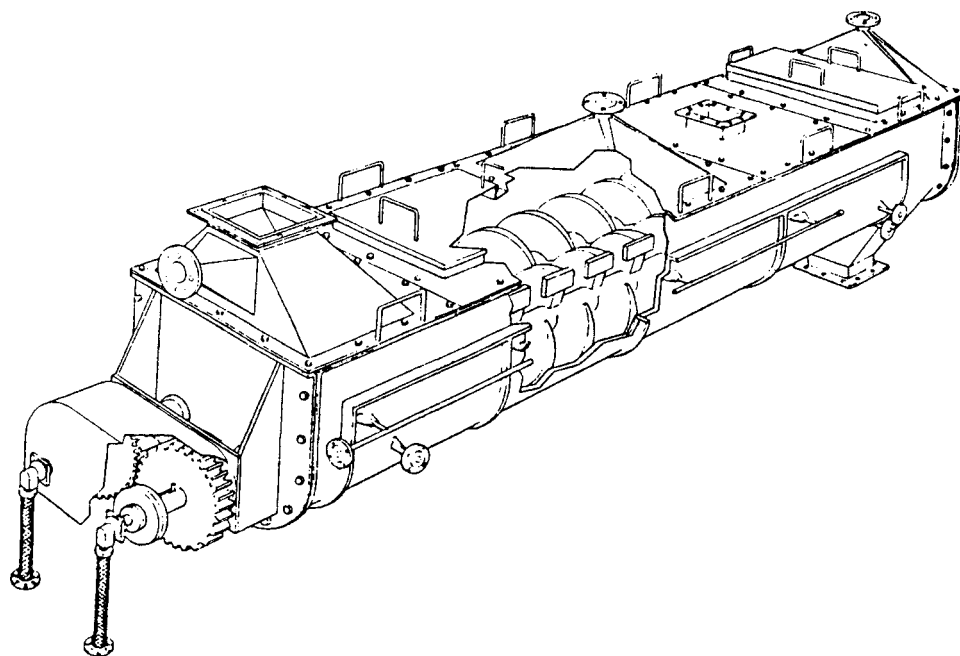


FIGURE 41.1 Paddle dryer.

across stationary plates. The plates are heated by a liquid medium or steam. Small plates with internal rims and large plates with external rims are arranged in an alternating sequence (Figure 41.2).

This arrangement makes the product drop from the outside edge of the small plate down to the large plate, on which it is conveyed to the inside edge, and then drops to the following smaller plate, where it is conveyed again toward the external edge. This design of the conveying system ensures plug flow of the product throughout the entire dryer. Each plate or

group of plates may be heated or cooled individually, thus offering precise control of the product temperature and the possibility of adjusting a temperature profile during the drying process. Thermal degradation of sensitive materials can thus be avoided, and cooling subsequent to drying can be achieved.

In the plate dryer, the product layer is kept shallow (approximately 10 mm). The entire plate surface is utilized for heat transfer. The product surface exposed to the surrounding atmosphere is even larger than the actual “wetted” heat exchange surface. The design of the product-conveying system ensures product turnover numbers in the range of 200 to 1500. A thin product layer on a large heat exchange surface coupled with high product turnover improves both heat and mass transfer rates. From vacuum plate dryers, the evaporated volatiles are removed by evacuation. Solvents can be recovered economically by simple condensation [8].

Plate dryers are typically fabricated in a modular design; this yields a wide range of dryer sizes with a heat exchange surface between 3.8 and 175 m².

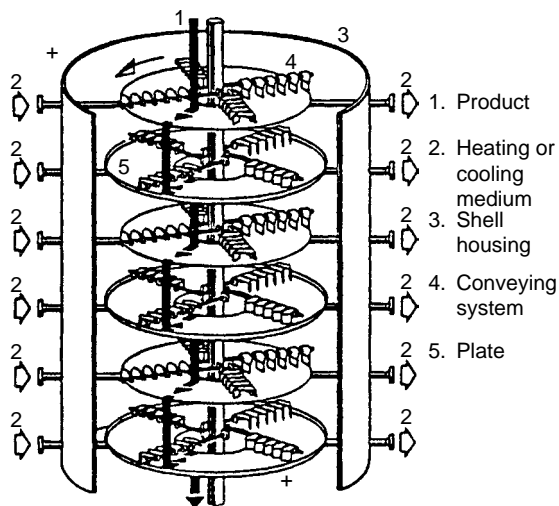


FIGURE 41.2 Plate dryer.

41.3.4.9 DRT Spiral Dryers

DRT is a recent innovation among the nonadiabatic contact dryers. It utilizes heat from a jacketed wall and transmits it to a thin, fast-moving product film rising in a spiral path along the inner wall surface (Figure 41.3). A very small quantity of the conveying medium is required to move the vapor from the dryer

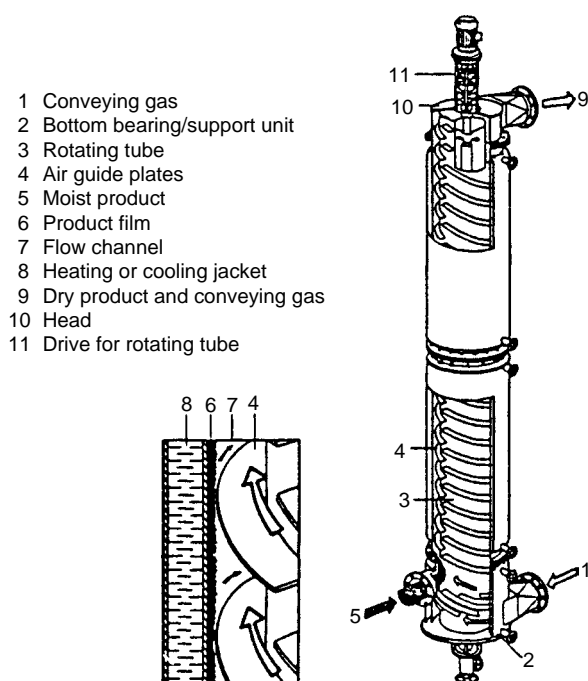


FIGURE 41.3 DRT spiral dryer.

since the heat transfer rate is very high and the cross-sectional gas flow area is small.

The jacketed outer cylinder rests on a base, which also embodies the product inlet. Product film moves spirally up the inner wall, and dry powder is discharged at the top.

A steam-heated concentric displacement body is placed within this cylinder, which rotates slowly by an external-gear motor. Many segmented air guides are provided on the outer surface of the cylinder and are arranged at a suitable angle. The distance between these plates and the inner wall must be greater than the product film thickness.

Conveying gas is blown tangentially by means of a blower into the tube base entering opposite a wet-feed metering screw. As a result, the gas disperses the wet feed by intense mixing. Product film is created by the inertial force, which threads its way upward in a spiral path until it reaches the exhaust port.

During this flight through the dryer, the conveying gas and the product are heated by the jacketed tube wall and the inner displacement body surface. Therefore, both gas and product temperature increase in the cylinder with a concomitant increase in the gas absolute humidity. This ensures that product moisture is lowered up to the discharge point, despite increasing moisture partial pressure. This increasing driving force guarantees lower final moisture contents when compared with a convectional flash dryer.

DRT is suitable for water-wet and solvent-wet chemicals, polymers, flour, and other products that are in a powdery form. For removing low-boiling solvents from polymeric products, it acts best as a predryer to a fluid-bed postdryer [9].

Among the advantages claimed are: (1) increased thermal efficiency; (2) lower power consumption; (3) compactness; (4) gentle drying; (5) reduced product holdup; (6) quick turnaround; (7) capability of using low-pressure waste steam for drying; (8) low product moisture content; (9) simple operation; and (10) minimum dust explosion potential. The unit also features low gas flow rates, short residence times (3 to 10 s), and high throughput (up to 20 t/h).

41.3.4.10 Miscellaneous Dryers

A number of proprietary dryers suitable for various polymer-drying operations are available in the market. Among them are the Solidaire, Continuator, Torusdisc, and Thermascrew (Figure 41.4) [10].

Solidaire is a continuous dryer consisting of a mechanical agitator rotating with a cylindrical housing, usually jacketed for indirect heating. The agitator is equipped with a large number of narrow, flat, adjustable-pitch paddles that sweep close to the inner surface of the housing. Residence time can be varied from seconds to 10 min by changing either the pitch of the paddles or the speed of the rotor. High paddle speed breaks up agglomerates and continually exposes new surface to the heat. It has been successfully used for drying ABS, PC, polyvinyl alcohol, polyolefins, and other polymers.

The Continuator is used primarily for removing tightly entrapped volatiles and for process applications requiring a long residence time. The mild agitation employed in this device provides gentle product mixing that minimizes “short-circuiting” while reducing particle breakup. This type of dryer can process polyethylene, PP, PVC, and other polymers.

The Torusdisc is another proprietary design particularly useful in processes that require high-capacity heating or cooling. Its chief advantage is its versatility. A single unit can be varied over a wide range of heat transfer coefficients, residence times, and temperature profiles. It consists of a stationary horizontal vessel with a tubular rotor on which are mounted the doubled-walled disks. These hollow disks provide approximately 85% of the total heating surface. It has been used commercially for drying ABS, PCs, polyolefins, and other polymers.

Thermascrew is a hollow screw, jacketed trough dryer that provides three to four times more heat transfer surface than simple jacketed screw conveyors and six times more than water-cooled drums. In either

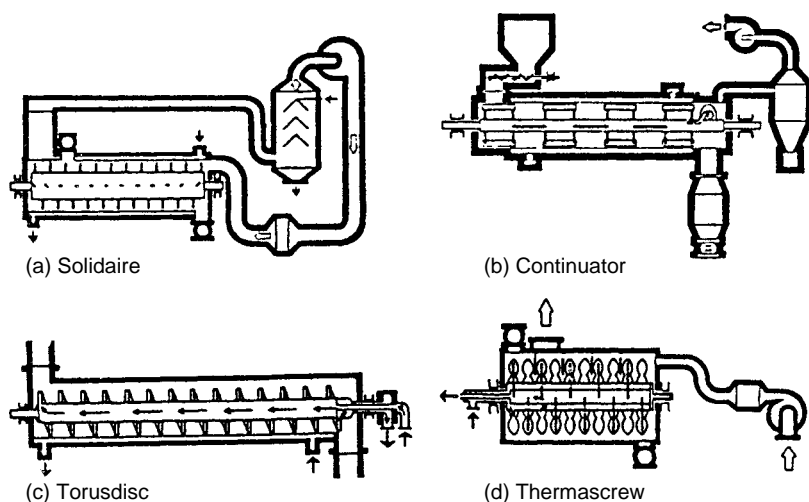


FIGURE 41.4 Some recent proprietary polymer dryers. (From Bepex Corporation, *CEP*, 79(4):5 (1983). With permission.)

continuous or batch operation, it provides efficient and uniform heating, cooling, evaporating, or other processing. It can operate in either a pressure or a vacuum environment. Polyester and polyolefins are among the materials dried with good thermal efficiency in such devices.

Among recent developments in dryers is the Yamato band FBD [11]. This is a modified FBD having all the components of a standard FBD with an additional carriage means with multiple blades mounted thereon and projecting there from for effective fluidization and transportation of materials (Figure 41.5).

The carriage includes a crank mechanism for effecting a circular or linear movement of the blades. It is driven in such a manner that the blades scratch

and fluidize the material being treated in cooperation with a heated gas. The fluidized bed is thus carried or conveyed toward the outlet port. The blades on the carriage extend in close proximity to the surface of the gas distributor plate. The blades may be straight, curved, or T-shaped. Such dryers can be used to process a variety of difficult-to-treat materials, e.g., slurries and materials containing solidified portions, as well as those having a high degree of cohesion or adhesion and/or containing lumps.

The spouted beds (SBs) can also be used to dry polymer beads. It is an efficient solid-gas contactor. In the conventional SB there is dilute-phase pneumatic transport of particles entrained by the spouting jet in the central core region and dense-phase downward motion of the particles along the annular region

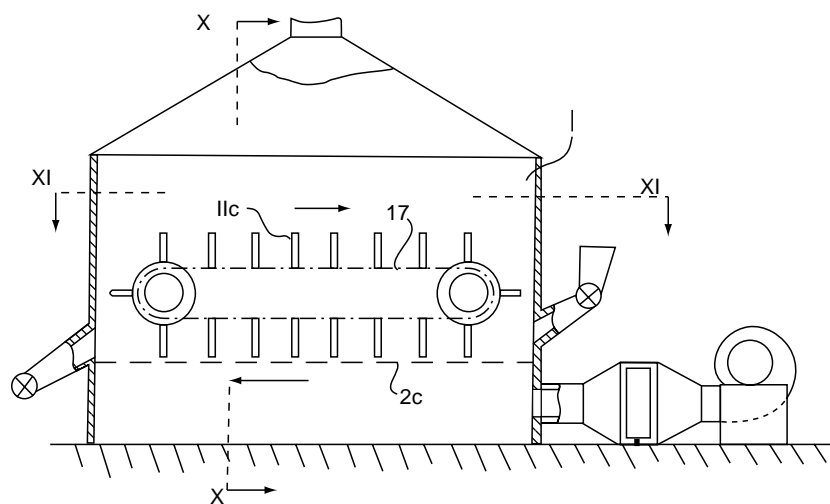


FIGURE 41.5 Yamato band fluidized bed dryer.

bounded by the cylindrical wall. Thus, the particle–gas contact is cocurrent in the core (or spout) and countercurrent in the downcomer or annulus. This characteristic recirculatory motion of the particles enables one to control the residence time of particles within wide limits by letting the particles go through a desired number of cycles prior to their withdrawal. With both batch and continuous operations possible along with the various modifications available, these beds have a strong potential as postdryers in polymer drying [12].

41.4 TYPICAL DRYING SYSTEMS FOR SELECTED POLYMERS

This section discusses briefly the drying of selected large-scale polymers. It is important to note the data in Table 41.1, which gives permissible moisture levels in various commodity resins [13].

41.4.1 DRYING OF POLYOLEFINS

41.4.1.1 Polypropylene

PP is produced by a variety of processes, most of them by a diluent phase propylene polymerization utilizing a Ziegler–Natta-activated titanium trichloride catalyst in the presence of low- to high-boiling hydrocarbons. Residual catalyst removal followed by hydrocarbon slurry centrifugation is the immediate upstream operation prior to thermal drying. Hexane is the solvent used in the major PP processes in operation today. As a result these polymers are solvent wet.

Many plants operate with two resin varieties, e.g., homopolymers and copolymers. Each requires

a different drying approach owing to different centrifuge cake-handling characteristics. Homopolymer cakes, although somewhat tacky, are much less than high-ethylene-content copolymer cakes, which tend to agglomerate, form lumps, adhere to surfaces, and so on. Considering capital cost, it is desirable to have a single dryer line for both resins. Consequently, initial dryer selection becomes a critical issue because of the feed flexibility required [14].

Both polymer centrifuge cakes are discharged hot (50 to 60°C), with diluent contents as high as 35 (wb) or 53.2% (db). Between 35 (wb) and 5% (wb), most homopolymers and copolymers exhibit constant-rate drying characteristics; i.e., all moisture evaporation is from the particle surface. Drying is rapid, and residence time is heat transfer-dependent. Since the product temperature limit for these polymers is 100 to 110°C, the solvent boiling point has a definite effect on dryer selection. Historically, again, rotary dryers are used. During the 1960s and early 1970s, a two-stage system of paddle-type dryers was used successfully. This consisted of a first stage or surface solvent dryer, with the characteristics of very high agitation, high heat transfer, and short residence time. The second stage, or bound moisture dryer, consisted of a device with low agitation, low heat transfer, and long resistance. Each dryer is provided with a recycle purge gas system to aid in controlling dew points and increase dryer efficiency. The gas flow is minimized; the amount used is that required to give a partial pressure necessary to achieve required product moistures.

With the emergence of very high capacity PP polymer lines and high-boiling point solvents, various types of dryers and drying systems have evolved. One reason for the advent of the new technologies in PP

TABLE 41.1
Percentage by Weight of Permissible Moisture (db) in Some Selected Polymer Resins

Material	Permissible Moisture		Drying Temperature (°C)
	Injection (%)	Extrusion (%)	
ABS resin	0.10–0.20	0.03–0.05	77–88
Acrylic	0.02–0.10	0.02–0.04	71–82
Cellulosics	Max. 40	Max. 30	66–88
Ethyl cellulose	0.10	0.04	77–88
Nylon	0.04–0.08	0.02–0.06	71
Polycarbonate	Max. 0.02	0.02	121
Polyethylene	—	—	—
Low density	0.05–0.10	0.03–0.05	71–79
High density	0.05–0.10	0.03–0.05	71–104
Polypropylene	0.05	0.03–0.10	71–93
Polystyrene	0.10	0.04	71–82
Vinyl	0.08	0.08	60–88

drying is the economic recovery of the flammable hydrocarbon solvents. Another reason is that PP has to be dried to a very low volatiles level and the final drying requires the drying gas to have an extremely low dew point. Usually nitrogen gas is used as a drying gas in a closed-cycle drying system. Figure 41.6 shows this by low and high gas dew point product moisture points indicating relative drying times. Reduction in recycle gas dew point is required to remove evaporated solvent first and, especially in mass transfer limited drying, reduce solvent partial pressure to increase the overall drying rate.

Accordingly, the low dew point case normally uses -20°C hexane dew point recycle nitrogen gas yielding the lowest residence time but at a higher energy expense than the higher dew point case. Since this is an expensive part of the flow sheet, refrigeration costs become a factor and recycled gas should be minimized. It is in this region that residence times vary from 30 min to over 1 h, depending on recycle gas dew point and polymer-drying characteristics; consequently, a controlled residence time dryer is desired. It is also desirable that this postdryer has an independent recycling loop to minimize energy consumption and maximize process control.

Based on these fundamentals, a two-stage flash or fluid-bed drying system has been developed for PP drying. The flash dryer disperses the feed cake in a venturi throat, with hot recycled gas breaking the cake and drying it to about 5% (wb) level. Final drying is carried out in the fluid bed in a nitrogen atmosphere. The solvent is recovered by a scrubber–condenser system. The PP, after being dried in the fluid bed, contains an extremely low level of solvent (e.g., hexane or heptane), typically 500 ppm.

A very recent development in terms of heat economy and corrosion control is the use of a spiral DRT dryer (Drallrohr Trocking) as a predryer in place of the flash predryer [9] in flash–plug-flow FBD systems. The gas/solid ratio is approximately 0.2 in these types of dryers, compared with 1.0 in the flash dryers.

Another important advantage of DRT dryer as a predryer in PP drying is its suitability in a corrosive environment. A persistent problem often seen in PP and high-density polyethylene (HDPE) manufacturing plants is the deterioration of the equipment due to free chlorides. The chlorides result from the deactivation of the activated catalysts with alcohol. Stress corrosion cracking is the most common corrosion phenomenon that results from the catalyst's chloride

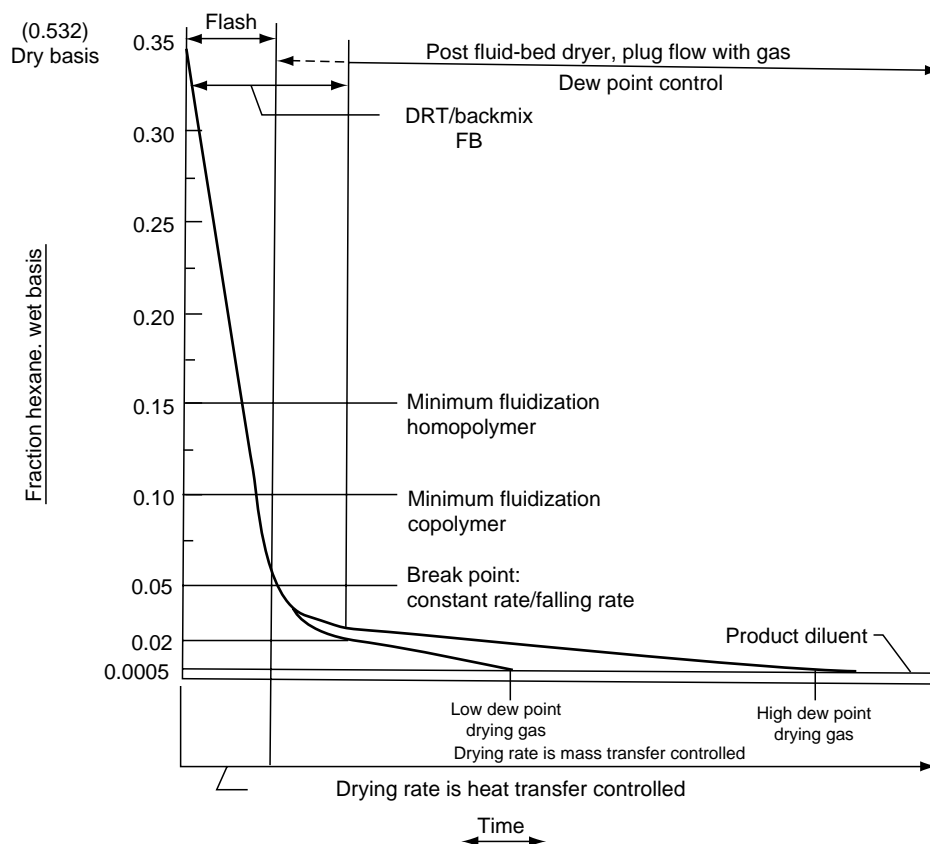


FIGURE 41.6 Typical polymer-drying curve of polypropylene.

41.4.1.2 High-Density Polyethylene

41.4.2 DRYING OF POLYVINYL CHLORIDE

41.4.2.1 Emulsion Polyvinyl Chloride

An improvement over the conventional open-cycle adiabatic spray dyers for E-PVC is the recycle exhaust spray dryer. In this type, up to 50% of the exhaust stream is recycled to preheat the supply air makeup from the atmosphere.

An improvement with respect to thermal efficiency is the two-stage dryer. This involves operating a spray dryer with a fluid-bed afterdryer. By adopting a two-stage layout with a fluid bed, powder is taken out of the spray dryer at a lower outlet temperature with higher moisture content. The cooler but higher moisture content powder is transferred to the fluidized bed, where the drying is completed to the desired

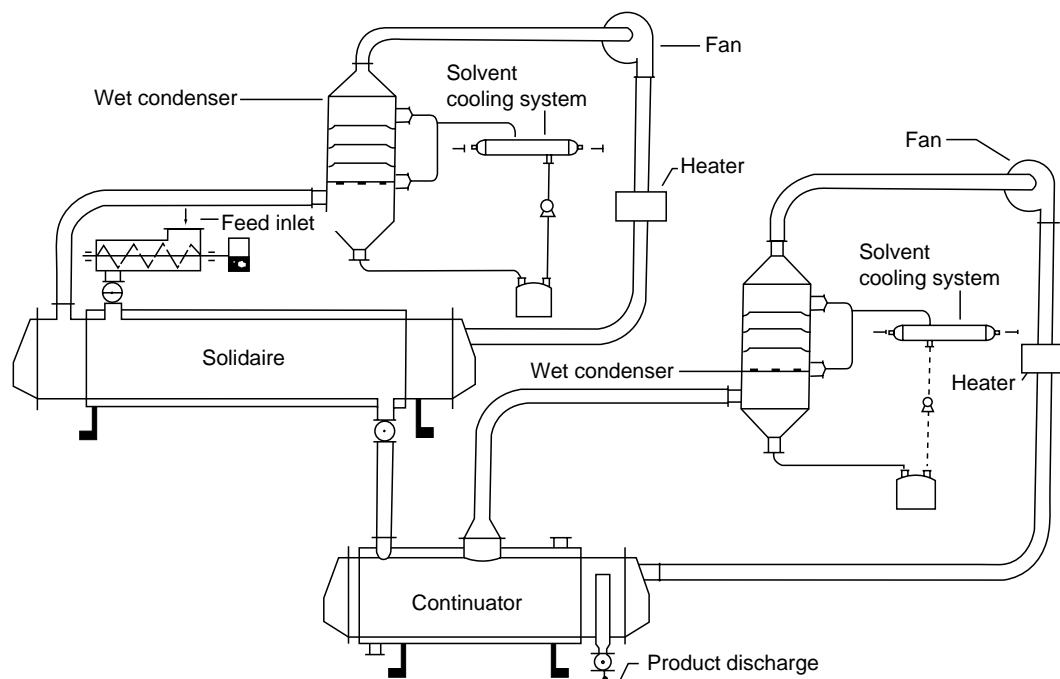


FIGURE 41.7 Drying system for polypropylene and polyethylene.

A recent improvement over the above-mentioned two-stage drying system for drying E-PVC is a spray dryer with an integrated fluid bed. The basic concept in this type of dryer is to avoid contact of the wet powder with any metal surface in the primary drying stage by transferring wet powder directly into a fluidized powder layer (second drying stage). To achieve this requirement, the fluid bed is integrated at the base of the spray-drying chamber.

41.4.2.2 Suspension Polyvinyl Chloride

Recently, a two-stage flash fluid-bed system has appeared in the market that is preferable to a rotary drying system (Figure 41.8). Most of the surface water is removed in the flash dryer stage within



seconds; some surface and all of the bound moisture are removed in the fluid-bed stage by holding the product at suitable drying temperatures for about 30 min.

Usually, wet cake with 22 to 25% water (wb) is fed to the dryer by a screw conveyor and enters by a special mill that deagglomerates the feed material, disperses it into the drying airstream, and accelerates it to duct velocity. The mill should handle the feed gently, as PVC is sensitive to high shear.

The flash dryer stage discharges the product to the fluid bed between 2 and 8% water, with the intermediate moisture chosen according to the basis of optimization used. Flash dryer air temperature may be typically 180°C at the inlet and 60°C at the outlet, depending on moisture content and the drying characteristics of the particular resin. As already mentioned, S-PVC is sensitive to shear; for this reason, dry duct velocities are kept low (around 15 m/s) and care is exercised in handling the dried product.

It is possible to arrive at the required product final moisture content by flash drying alone but, because of the residence time available in the flash dryer, the high temperatures required give an unsatisfactory product. Moreover, there is a very wide range of S-PVC homopolymers, varying in molecular weight, particle size, and other properties, and all have different dewatering and drying characteristics.

The benefits achieved in a two-stage system are its ability to handle upsets in inlet moisture in the flash dryer, a lower energy cost, and a relatively simple scale-up.

Modest improvements with respect to the most economical drying of S-PVC are a continuous, single-stage, contact fluidized bed dryer, as shown in Figure 41.9. In this type of dryer, the concepts of back-mixed fluidization and plug-flow fluidization are advantageously combined in a single unit. A broad residence time distribution is obtained in a back-mixed fluid bed in which the bed itself has a relatively small length/width ratio. In performance it can be compared with an agitated tank provided with overflow, inasmuch as the vigorous mixing inside the fluid bed will result in a uniform temperature and constant average moisture content of the particles throughout the entire bed. The product discharged from this back-mixed fluid bed has the same temperature and moisture content as the bulk material inside the fluid bed. Further, because of the excellent heat and mass transfer between the fluidized particles and the drying air, equilibrium is reached between the exhaust air and the product inside the bed. This type of fluid-bed drying concept is found to be very suitable for drying surface moisture when residence time has no impact on the drying performance.

After the mixed-bed section, a plug-flow section is provided in which the final drying of PVC takes place. This section is fairly small compared with the back-mixed section and is usually obtained by dividing the fluid bed into compartments. This concept is particularly advantageous for drying bound moisture from heat-sensitive materials since the residence time is controlled within the narrow limits and a distinct

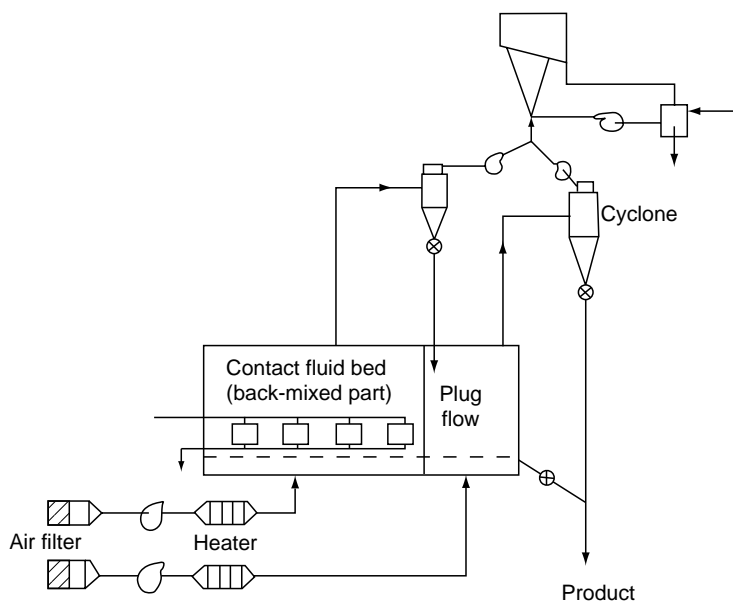


FIGURE 41.9 Contact fluidizer for suspension-grade polyvinyl chloride.

moisture profile can be obtained along the length of the unit because of a very low degree of back-mixing.

In this type of drying system for S-PVC, wet PVC cake is usually transported from the decanter centrifuge by a screw feeder to the product distributor of the back-mixed fluid-bed section. It then flows through an overflow weir into the plug-flow section where the final drying takes place. Finally, the product is discharged through the discharge weir arrangement.

The back-mixed section of the unit is provided with heating panels; no heating panels are provided in the plug-flow section, partly because the cost cannot be justified and partly because of the tendency for electrostatic deposits on the heating panel encountered with PVC at low moisture content to decrease the heat transfer coefficient.

The contact fluidized bed provided with heating panels appears to have proven to be superior to the flash fluid-bed drying system from the point of view of heat economy and overall savings. The contact fluidizer does have a few limitations. First, it is mandatory that the polymer material be readily fluidizable at a moisture level well above the moisture level in the back-mixed section to avoid defluidization of the bed during upset conditions. Second, the centrifuge cake should not be too sticky and have too much tendency to form agglomerates of the individual polymer particles. In such a case, a flash dryer is better suited as the predrying stage as better disintegration takes place in the venturi section of a flash dryer than in a back-mixed fluid bed.

Although a fluid bed as a second-stage dryer gives accurate product temperature control while providing adequate residence time, depending on the predryer load, evaporative load in this stage may be small. This results in a low airflow requirement and makes fluidization more difficult. In such cases, a vibrating fluid-bed design is a better alternative. Here, PVC is conveyed by vibration, permitting varying gas speeds without affecting the conveying rate or residence time. Also, with the low airflow rates of the vibrating fluid bed, the fines pickup problem (normally associated with high gas flow rates) is minimized and, as the vibration is at a low frequency, the overall effect of the gas and vibration is to transport the product gently, minimizing damage. The vibrating FBD must be among the most important but underutilized dryer of all granular products.

During fluidization of PVC, electrostatic charges arise of such magnitude that they affect the hydrodynamics of the system. This is disadvantageous for transfer processes in the bed, e.g., for heat transfer between the heating surface and the bed. This is a difficult problem in a fluidized bed because of intensive movement of particles and frequent interparticle

and particle-wall contact. Although charge generation cannot be prevented, one can limit its magnitude (and try to increase its dissipation) by changing process conditions. One method is the addition of a small portion of fines to the bulk; this results in the splitting of agglomerates and disappearance of the particulate layer at the walls. As a result, the bed regains its original parameters, which assure intensive running of processes in the bed.

41.4.2.3 Vinyl Chloride-Vinyl Acetate Copolymer

There is a wide difference in the difficulty of drying vinyl chloride-vinyl acetate (VC-VA) copolymers according to the degree of VA content in polymer and extent of polymerization. If the heat-resisting property of polymer is too low to use the hot air at a high temperature (even if the hydroextracting degree in the former stage is generally good, e.g., 13 to 17% wb), then it is difficult to remove VA monomer. As a result, the necessary retention time becomes longer compared with that of PVC-homo.

The equipment recommended for this application is single-stage batch fluidized bed dryer (B-FBD) or a flash B-FBD system. For proper selection it is necessary to make a detailed study on the basis of specified conditions.

An important factor that should be taken into account while drying PVC is the corrosion of the equipment due to the monomer chloride. Monomer chloride, which is always present in the wet cake, induces pitting corrosion and stress corrosion cracking in parts where the powdery materials are processed. Those parts, therefore, are made of AISI-316L and are partially coated with an acid-resistant coating. It is indispensable to make periodic inspection of the corrosion condition and to make timely replacement of the necessary spare parts. Preventive maintenance is imperative to successful operation.

Another important consideration in drying PVC is the emission of VC. U.S. EPA emission limitations of <5 ppm on VC must be strictly maintained. This criterion on VC sometimes dictates the selection of the drying equipment for PVC. In other countries the discharge limit on VC emission may be less stringent.

41.4.3 DRYING OF ACRYLONITRILE-BUTADIENE-STYRENE

In general, emulsion processes are used to make ABS of higher impact strength and bulk or suspension processes are preferred for materials with less impact strength. This three-monomer system can be tailored to end-product needs by varying the ratios in which they are combined. Acrylonitrile contributes heat

stability and chemical aging resistance; butadiene imparts low-temperature property retention, toughness, and impact strength; and styrene adds luster (gloss) rigidity, and processing ease.

The drying characteristics of ABS polymers change with changes in composition. Generally, a centrifuge cake containing 50% moisture (wb) must be dried to a final product containing less than 0.1% moisture. The critical moisture composition is around 5%. The allowable product temperature is approximately 100°C. ABS plastics are mildly hygroscopic; if dried, ABS is left in storage for some time and it must be dried again to reduce the moisture to a level ($<0.1\%$) adequate for most applications. On the basis of these physical properties, single-stage, cocurrent, and direct heat transfer rotary dryers and flash dryers are commonly used. Rotary dryers have the advantage of a longer residence time, making them suitable for drying ABS polymers with a larger particle size. The flash drying system is suitable only for small particle sizes but is more economical with regard to thermal efficiency.

In case ABS forms lumps in the course of the coagulation and/or dehydration process, it is necessary to add another process to crush the lumps, i.e., to install an FD with a cage mill or use a ring dryer in the first stage of the dryer. Since drying in the falling rate has the main objective of removing the monomers, it is necessary for the material to have a long retention time. To satisfy such a requirement, a batch FBD is widely adopted.

Drying of ABS has been commercially successful in a two-stage drying system with the combination of direct and indirect heat transfer. Since ABS requires both surface and bound moisture removal, a two-stage drying system is recommended.

The two-stage flash FBD system is advantageous in terms of thermal efficiency and product quality. The first-stage flash dryer does most of the evaporation. FBD, characterized by longer residence times, is used in the second stage. In the second stage, FBD can be replaced by a direct or indirect rotary dryer. If a fluid bed is used as the second stage, it is advantageous to use the plug-flow model since in such a bed residence time can be controlled within narrow limits.

Among the developments for drying ABS are the indirect-heated closed-loop, inert gas-heated, or liquid-heated dryers. These dryers minimize the emission of styrene monomer and oxidation of the polymer is prevented by the inert purge gas. The overall efficiency is also high. A particular type of this class of dryers is the indirect-heated FBD depicted in Figure 41.10. This type of dryer uses a rectangular bed to optimize the solids flow and heat transfer fluid LMTD effect. Also, the plenum-side inlet gas is at a low temperature, precluding any mechanical constraints. In this process, an external direct-fired heater operating at low excess combustion air heats a heat transfer fluid (e.g., molten salt, thermal fluids, steam, and others) to a temperature above that of the bed, but below the ABS degradation temperature. Since the heat source is decoupled from the fluidizing gas

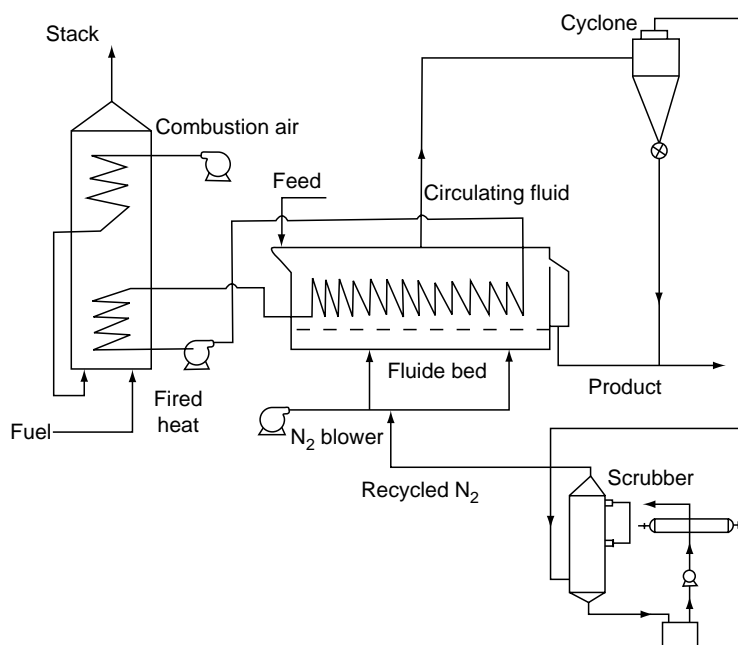


FIGURE 41.10 Contact closed-cycle fluidized bed dryer for acrylonitrile-butadiene-styrene.

source, large vessel diameters are not needed. Further, the smaller amount of fluidizing gas requires much smaller pollution control equipment.

When it is possible to obtain such wet raw material that is properly coagulated and dehydrated but with no formation of lumps and yet has a low level of moisture, the single C-FBD as shown in [Figure 41.9](#) has been widely used in recent years.

ABS group resins are highly inflammable and self-combustible and liable to cause dust explosion. It is absolutely necessary to be very alert not only in setting and controlling the hot air temperature but also in eliminating any possible kindling causes, e.g., introduction of metallic foreign substances in the raw material and overcharged static electricity. Careful maintenance is further required. Periodical cleaning to remove the resin adhering to the equipment is essential for safety.

ABS, while drying, emits styrene, a highly toxic substance. Very recently the U.S. National Institute for Occupational Safety and Health (NIOSH) has set a limit for workplace exposure of styrene. NIOSH suggests that workers should not be exposed to >50 ppm of styrene over a time-weighted average of 19 h/day, 40 h/week. Further, a ceiling concentration of 100 ppm during any 15-min sampling period is enforced in the United States.

Owing to this recent regulation, there are indeed very few optional routes left for drying ABS other than indirect-heated drying with an inert closed-loop gas system.

41.4.4 DRYING OF SYNTHETIC FIBERS

Polymers that demand special precautions during drying are common in the synthetic fiber industry. Of these, nylon and polyester chips are the two most common examples. These resins are hygroscopic and have to be dried before a spinning or molding process. Generally, these polymers are introduced to the dryer in the form of 3- to 4-mm cubic pellets.

41.4.4.1 Nylon

Nylon is the generic term for any long-chain, synthetic, polymeric amide in which recurring amide groups are integral to the main polymer chain [3]. There is a wide choice of starting materials from which polyamides can be synthesized. The two primary mechanisms for polymer manufacture are condensation of a diamine and a dibasic acid or their equivalents or polymerization of monomeric substances. Nylons are identified by a simple numerical system. The words *polyamide* and *nylon* are followed by one or more numbers. One number indicates that

the product was prepared from a single monomeric substance and also indicates the number of carbon atoms in the linear chain of the recurring polymer unit. For example, nylon-6 is manufactured by the polymerization of caprolactam and nylon-11, from 11-aminoundecanoic acid. When two numbers are used, they are separated by a comma and refer to the reactants used in the polymer's manufacture. The first number refers to the number of carbon atoms in the diacid. Thus, nylon-6,6 is prepared from the reaction of hexamethylenediamine and adipic acid. The difference in numbers of carbon atoms between the amide groups results in a significant difference in mechanical and physical properties. Although the theoretical number of nylon types is very large, a few are commercially available. Of these, nylon-6 and nylon-6,6 comprise about 75 to 80% of the nylon fiber and nylon-molding compound market.

Nylon chips are normally dried from 4 to 10% inlet moisture (wb) to <0.1% outlet moisture. If they are allowed to absorb moisture, they must be dried prior to processing. Some nylon may hold as much as 2% moisture under normal storage conditions but must still be processed satisfactorily with less than 0.1% moisture remaining in the material for reuse. Because of the low temperature limits (70 to 80°C) allowable for drying nylon, very low dew points and longer times are required to achieve even this much dryness. The common dryer for nylon is the batch vacuum tumble dryer. The drying temperature is kept controlled within 70 to 80°C, and drying time ranges from 10 to 24 h. If vacuum drying is not possible, use of recirculating dryers at 80°C and dehumidified air is the next best solution. During hot, humid weather, attention must be paid to guarantee that the recirculating air is indeed dry or moisture will be added to nylon rather than removed. Prolonged exposure to this drying condition can result in discoloration and possible property deterioration.

Nylon has a poor polymerization effect, and the chips have a high moisture content at the beginning with a propensity for holding rather low levels of moisture very tenaciously. As a result, a long time is required for drying. For these reasons, it is advantageous to use FBD and/or PDD for this process. In fact these dryers can perform drying down to 0.002% moisture content in 4 to 6 h.

Another characteristic of the nylon is that, if it is at low moisture content, it is subjected to oxidative deterioration and discoloration at high temperatures. Because of this problem, it is usual to dry it with air when the moisture content is high and then to dry in an inert atmosphere.

41.4.4.2 Polyester

A polyester fiber is any long-chain synthetic polymer composed of at least 85 wt% of an ester of a dihydric alcohol (HOROH) and terephthalic acid (TA) (*p*-HOOC₆H₄COOH). The most widely used polyester fiber is made from linear polyethylene terephthalate (PET).

PET is a linear homopolymer, i.e., a condensation polymer of TA or its dimethyl ester, dimethyl terephthalate (DMT), and ethylene glycol. The polymer is melted and extruded or spun through a spinneret, forming filaments that are solidified by cooling in a current of air. The spun fiber is drawn by heating and stretching the filaments to several times their original length to form a somewhat oriented crystalline structure with desired physical properties.

During early stages of processing of PET, drying was carried out in batch vacuum tumblers. The processing time was 10 to 12 h. As the demand for larger capacity gradually increased, the multistage, batch-type fluidized bed drying system replaced the older vacuum tumbler dryers.

A characteristic of a PET chip is that, if the raw material is heated at 90 to 100°C, its composition is rearranged from a vitreous to a crystalline form. The chips stick to each other owing to surface melting when they are heated at a high temperature. In order to avoid this problem, the drying system is divided into two stages. In the first stage crystallization and preheating are accomplished; in the second stage drying is completed. In the first stage, the heating is gradual. Agitation is required to prevent sintering or sticking of the product at this stage. Usually, a fluidized bed or agitated vessel is used for this purpose.

After surface crystallization is performed, the chips do not show adhesiveness before the temperature rises to the melting point. Advantage is taken of this property of the chips, which are then discharged into a continuously moving bed dryer. Usually nitrogen, with a dew point temperature of -40°C, or dehumidified air is passed countercurrent to the product flow. In continuous operation, a 2-h gain in residence time could be achieved.

In recent years, with the diversification of the applications of PET, there is a demand to miniaturize the equipment and to save energy. This has motivated various special dryer designs exclusively for PET. One is PDD. A combination of B-FBD and PDD has a chip retention time close to that of an ideal piston flow, thus enabling considerable savings in the energy cost for drying.

41.4.5 MISCELLANEOUS

Polystyrene (PS) and acrylonitrile-styrene (AS) are two other polymers produced in bulk quantities. Previously, these polymers were dried with FD. Later, C-FBD replaced all previous FD dryers because of their energy savings advantage. In recent years, paddle dryers have made rapid gains. The heat-resisting power of these materials is comparatively low. Melted material will adhere to the walls of the equipment if the processing temperature is not properly regulated.

PC is another commodity resin that demands careful drying. When the polymer was first commercialized, it was common to use FD plus B-FBD with the steam-stripping process. In recent years this has been gradually switched to PD with the idea of energy saving and of the direct process of chloride solvent without steam stripping. The Solidaire dryer is another possible choice. Since PC has comparatively high heat resistance, the drying process is not difficult.

Polypropylene oxide (PPO) is a recently developed resin with an application that is rapidly expanding. It requires a comparatively long drying time since it contains superfine particles and has high affinity for water. Of various kinds of polymers, this is the one that requires the most difficult processing techniques. The paddle dryer is found to process this material economically.

41.5 DRYING OF POLYMER RESINS

In order to avoid surface defects in molded parts and sheets made from resins, it is usually necessary to dry the pellets before processing. Residual moisture above some critical level can cause a finished product with unsatisfactory surface finish and properties. Drying is required to reduce the moisture content of the pellet below some critical value. The degree of dryness depends on the specific nature of each converting operation; some require more critical moisture control than others. For example, PET and nylons are very hygroscopic but for different reasons. PET in normal storage conditions contains about 0.15% moisture (db). It must be dried to a level of 0.005% (db) or better for processing. Although PET is not difficult to dry because of the high temperature that can be used, it can have absolutely no exposure to atmosphere between drying and processing operations. On the other hand, some nylons may hold 2% moisture under normal storage conditions but can be processed satisfactorily with 0.1 to 0.15% moisture in the material. Because of the low temperature limits (70 to 80°C) allowable when drying nylon, very low dew points and longer drying times are required to achieve even this much dryness.

41.5.1 GENERAL OBSERVATIONS

Depending on the degree of affinity for moisture, plastic resins can be divided into two classes: (1) hygroscopic and (2) nonhygroscopic. Moisture adsorption and/or absorption capability depends on the type of resins as well as the ambient temperature in which it is placed. In some instances, exposure of only few minutes can be detrimental. If the material is exposed to a certain temperature and relative humidity for a period of time, it will reach the equilibrium point, referred to as the *equilibrium moisture content* (EMC). Prior to drying it is important to know the permeability (product of the diffusion constant of water vapor–polymer system and the solubility coefficient) of polymer to water vapor since this dictates the condition for relative humidity for the safe storage of the polymer [16].

41.5.1.1 Nonhygroscopic Resins

Polyethylene, polystyrene, and PP fall under the classification of nonhydroscopic resins. These types of polymer resins collect moisture on the surface of the pellet only. The moisture can originate from several potential sources. Such moisture in some cases can be removed very easily by moderate preheating immediately prior to feeding the material into the mold. In some cases it is sufficient to provide vents at the transition from the hopper to the mold cavity. In some situations the moisture can be removed by passing warm air over the material. The equipment utilized to heat air and dry resins is usually very simple, e.g., an inlet air filter, a blower, and a controlled electric heater, as shown in Figure 41.11.

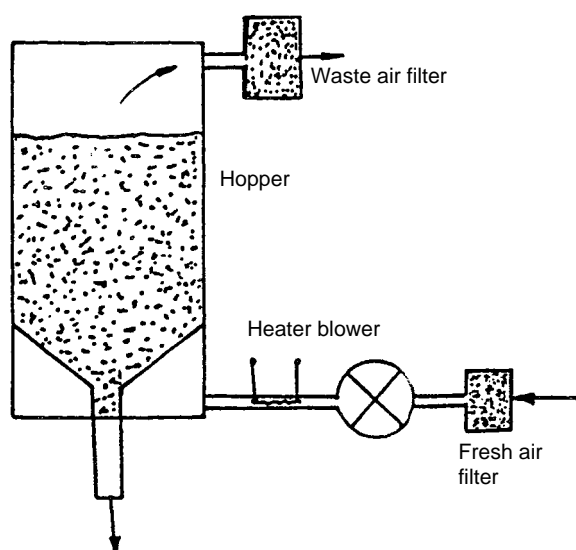


FIGURE 41.11 Resin dryer (open cycle).

41.5.1.2 Hygroscopic Resins

PET, nylon, ABS, and PC come under the classification of hygroscopic resins. These types of polymer resin collect moisture inside the pellet itself. Removal of this moisture requires dry air as well as heat. These resins therefore demand proper design and careful machine selection for each application. Desiccant dryers are the dominant technology for these resins.

41.5.2 DRYING METHODS

41.5.2.1 Drying with Heat as Transfer Medium

In these types of dryer, air is used exclusively as a heat transfer medium. A distinction is made between

1. Dryers with fresh air only (open system)
2. Air circulation dryers with partial supply of fresh air (semiopen system)
3. Desiccant dryers

Figure 41.11 shows the simplest type of dryer with fresh air operation. Heated air flows through the bed of granules, normally from bottom to top, uniformly heat the bed of granules and at the same time carry off the moisture. The air temperature at the inlet is kept approximately 20°C above the desired temperature of the granule bed. Its advantages are: (1) these units are inexpensive; (2) easy to handle and clean; (3) readily attachable to molding machines; and (4) have a high degree of efficiency (30 to 80%). Among its disadvantages are: (1) drying depends on dew point temperature (i.e., weather and climate); (2) it has only a moderate efficiency for hygroscopic resins (20 to 30%); (3) there is possible contamination of environment and granules (pollution); and (4) the exhaust air is at a high temperature (40 to 60°C). Usually these types of hoppers or dryers are suitable for nonhygroscopic plastics, e.g., polyolefins and polystyrene.

In dryers with partial recirculation (Figure 41.12), all the exhaust is not vented into the atmosphere since it still contains energy. Instead, 70 to 90% of the exhaust air is recirculated. The fresh air makeup usually ranges from 10 to 30% of the total flow, which increases the drying capacity of the recirculated air. These dryers are more energy efficient than the open type but have the same relative advantages and disadvantages and are particularly suitable for nonhygroscopic and mildly hygroscopic resins, e.g., ABS, PC, PMMA, PPO, and SAN.

In desiccant dryers, the heat transfer medium (generally air) is given an additional treatment to lower the dew point in a desiccant chamber, where the moisture is removed. This dried air passes through a heating chamber and the fixed bed of granules from

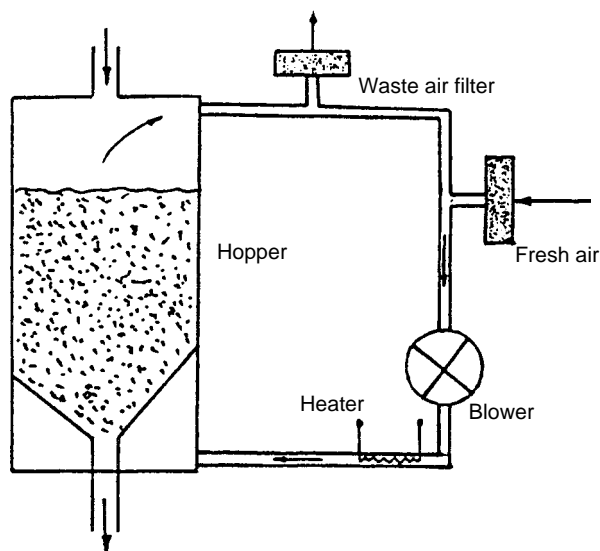


FIGURE 41.12 Resin dryer (semiopen cycle).

bottom to top. Two most commonly used desiccants are silica gel and molecular sieve. New polymeric desiccants have recently been developed. When the inlet concentration of moisture in the airstream is high, silica gel removes more moisture by weight. For lower inlet moisture conditions, the molecular sieve works best. The incoming airstream to the desiccant bed in a plastic dryer is warm, generally above 40°C. This makes use of a molecular sieve necessary to remove moisture. Another advantage of the molecular sieve is that it produces 1000 kcal/kg of moisture absorbed. As a result, a bed with a molecular

sieve is not only capable of achieving lower moisture dew points, but also requires less energy input (as heat) to achieve drying rates.

Figure 41.13 shows a desiccant bed system in the semiopen design. In this unit the smaller stream, heated to approximately 200°C, passes through a desiccant chamber where moisture is removed from the adsorbent and is then vented. After this regeneration of the adsorbent, the chambers are rotated. Commercial units are also available in which the circulating air is not exchanged. This design features redrying in one chamber at a time by preheating fresh air. Since in such units the entire desiccant battery is removable, the adsorbent is redried outside the granule-drying circuit. This ensures almost constant drying capacity. By cooling the returning airflow with an additional cooler, it is possible to lower the dew point far below the ambient temperature.

Generally, hygroscopic resins, e.g., nylon-6, nylon-6,6, PET, PBT, and ABS, are dried in desiccant dryers. The dew point of the drying medium has a significant impact on drying hygroscopic resins. For example, PET absolutely requires dew points in the -40 to -50°C range to be adequately dried. For other hygroscopic resins, dew points in the range of -15 to -25°C are adequate.

41.5.2.2 Drying without a Heat Transfer Medium

An alternative to drying polymer resins is the use of vented barrels for drying without a transfer medium. This technology for drying resins is gaining ground. In one of the proprietary designs, an annular

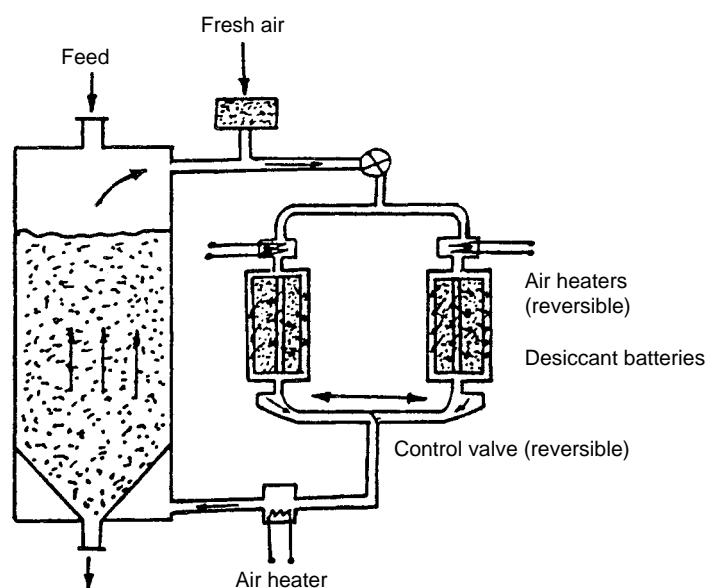


FIGURE 41.13 Resin dryer with desiccant batteries.

chamber formed by a tubing made of mesh in the center of the dryer and a perforated external shell or barrel surrounding the flowing materials are used. The material enters the feed port on the top of the dryer and then flows by gravity. The perforated shell is covered with bands to heat the material, and the air is drawn through it. This assembly is further enclosed in an external protective jacket. As the material flows through the annular chamber, air entering between the protective jacket and the inner perforated barrel is heated and drawn up the mesh tubing or chimney in the center of the dryer. Air travel is controlled by a compressed-air venturi. As heated air passes through the plastic granules, it drives off surface moisture and preheats the material before it enters the feed throat of the screw. Any internal moisture remaining in the hot pellets is flashed off almost instantaneously by shearing action. Water vapor flashed off in the barrel is drawn out by the mesh tubing chimney, providing an unobstructed escape path for the moisture, which is exhausted into the atmosphere.

Advantages of venting are: (1) little risk of contamination; (2) operation independent of moisture content; (3) reliability; (4) consistency of quality; and (5) removal of residual monomers under favorable conditions.

41.6 DRYING OF SELECTED POLYMERS

From the discussion above, it is obvious that there is an application for several dryer types for drying of polymers and resins. For instance, suspension PVC is usually dried either in two-stage systems involving a flash predryer followed by a fluid-bed second-stage dryer (with or without tubes), or by a single drying stage such as a rotary drum dryer or, more commonly, an FBD with internal heat transfer surfaces (e.g., tubes, coils, or plates). Emulsion PVCs, on the other hand, are mainly processed in spray dryers. PP is dried in similar systems to those used for suspension PVC, and the various forms of polystyrene are processed in either flash or FBDs. Polyacrylonitrile, which is frequently produced as a filter cake, is dried either on a band dryer after being preformed into a suitable shape, or dried in a single- or two-stage flash dryer. Some forms of polyethylene require drying and here again systems are either flash dryers or FBDs with in-bed heat transfer tubes. For polyamides, column dryers are mainly used under a nitrogen blanket in order to avoid oxidation. Some applications employ low-temperature fluid beds to dry the granules.

Polyester granules (e.g., PET) are used for the production of bottle polymer, film (either video or wrapping), and fiber or filament. These types of polyester require quite a different system as the product first

undergoes a crystallization stage before reducing the moisture to very low levels, below 50 ppm. Due to the very special requirements for this type of polymer, special processing systems have been developed. The following presents an application for the economic drying of polyester chips to very low moisture for the production of microfilaments. The same drying systems can be used for any of the other polyester products, as well as for the drying of PBT and some PC granules.

In recent years, the trend in the production of polyester yarn is to produce ultrafine microfilament at 5 to 7 μm diameter that requires drying to 20 ppm. Traditional filament yarn and staple fiber having a diameter of 18 to 22 μm typically require 50-ppm final moisture.

The first continuous PET drying system, originally developed by Rosin Engineering (London), was a combination of horizontal paddle crystallizer with a vertical column dryer. This system was used extensively for the production of all types of fibers, e.g., industrial yarn, bottle polymer, and film. Although quite versatile in that it can be used with different types of granules having completely different sizes, it has the disadvantage that there is a slight formation of dust due to the mechanical action of the paddles, and also that space has to be left at one end of the crystallizer for the withdrawal of the rotor shaft. However, at the same time, fluidized bed units for solid-phase polymerization (SPP) of PET and polyamide were being developed. It was observed that there were several advantages in using a fluidized bed for the initial heating and precrystallizing phase as compared with the rotary paddle type of other existing systems. Rosin manufactured its first combined fluidized bed crystallizer and column dryer for PET drying in 1970 (Figure 41.14).

The system consists of a fluidized bed heater/precrystallizer and a column dryer for PET. The fluid-bed section has five main functions:

1. Evaporation of surface and some internal moisture from PET
2. Transformation of PET from the amorphous to the crystalline condition
3. Heating of the chips to the temperature required for drying in the column
4. Provide sufficient turbulence to avoid sintering or chips sticking together
5. Removal of dust from the incoming granules

PET chips are fed into the fluid bed by a variable-speed vibro feeder and a fixed-speed rotary valve that acts as a gas seal. They meet an oncoming stream of heated gas (e.g., nitrogen or air) and become partially suspended in the flow. As more chips are fed in, the

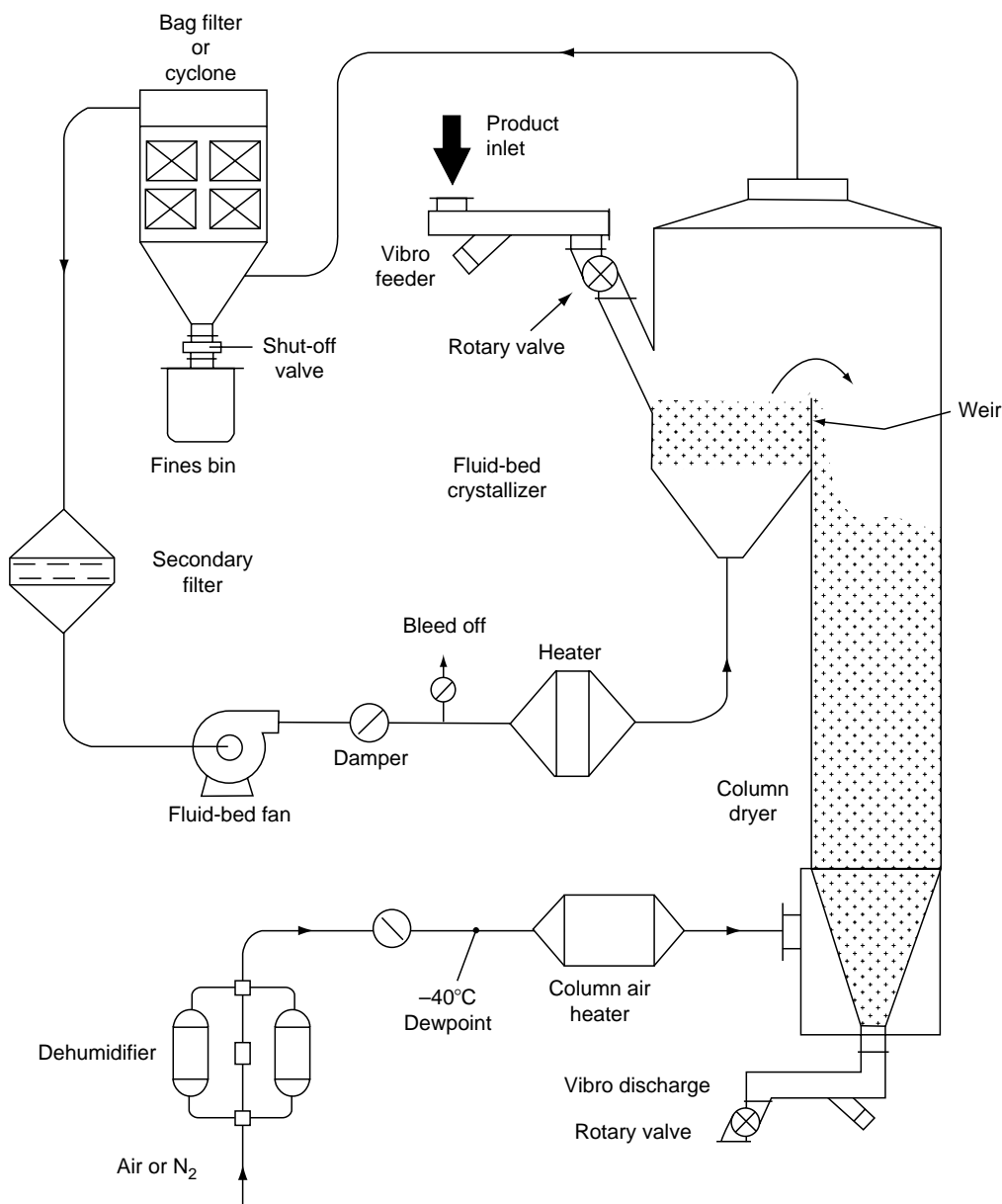


FIGURE 41.14 Continuous fluid-bed crystallizer/column dryer for polyester. (Courtesy of Rosin Engineering/Rosin Americas Ltd.)

fluid bed becomes established as a deep agitated mass of material exhibiting many properties of a fluid.

When a wet chip (typically 0.5% moisture) falls into the fluid bed, the surface moisture is rapidly evaporated. As the chip is then further heated, crystallization occurs. This amorphous-to-crystalline transformation of PET is an exothermic reaction and the heat given off is quite sufficient to raise the surface temperature of PET to above the softening point. If the chips are not moving as they do in the fully developed fluidized state, this will produce large solid lumps of agglomerated chips.

It is important to prevent agglomeration of chips so that the subsequent drying stage may proceed uniformly. Agglomerates may not dry completely, which gives rise to subsequent processing problems, particularly with microfilament production. In addition, agglomerates can lead to material flow problems if left unchecked and can shut down entire operations. Plug flow in the fluid bed is achieved through a system of internal baffles (which can be adjusted if necessary to alter the residence time) so that each chip is given a very similar thermal treatment and therefore achieves uniform crystallization.

When the chips reach the last section of the fluid bed in the crystallized and heated condition, they pass over a weir and descend into the column dryer. This unit was originally developed by Rosin in conjunction with ICI over a period of 2 y to arrive at a design that gives true piston or plug flow.

A constant, gentle flow of heated, dehumidified air or nitrogen is provided upward through the column. The dew point of this gas is carefully controlled by either a molecular sieve absorption system when using compressed air, or a combination of two-stage refrigeration/drying system for low-pressure gas supplied by a fan. The compressed air systems are more cost effective for plants with capacities up to 1000 kg/h. Above this rate, the ambient air system is usually preferred. When nitrogen is involved, it becomes more cost effective to use a closed-loop, low-pressure dehumidifying system for all but the smallest of plant capacities.

The production of chips with a final moisture content of 50 ppm typically requires a residence time of 2 h using heated gas with a dew point of -40°C . Whereas this moisture level is satisfactory for normal yarns and staple fibers, microfilament quality requires a moisture level of 20 ppm, which is achieved by corresponding adjustments to the retention time and gas dew point.

In [Figure 41.14](#), the gas supply to the column dryer is heated after dehumidification using a similar heat exchanger as in the fluid bed. This relatively small amount of gas mixes with the gas above the fluid bed and the same quantity is then vented off to atmosphere (air) or to a return line (nitrogen) after the supply fan, such that the whole system is kept in pressure balance. This small air loss is in fact the only energy loss in the system. Since the gas for the fluid bed is recycled, the heat input closely matches that required to heat the chips, which in any case are heated for the subsequent extrusion process. In many instances, the column dryer is positioned directly above the extruder.

41.7 CONCLUSION

It is clear from the discussion in this chapter that numerous dryer types can be used for drying of polymers. Rotating double-cone vacuum dryers, e.g., can be used up to 300°C and 0.1 torr absolute pressure for processing PET, PBT, and liquid-crystal polymers. The combination of high temperature and low pressure assists semicrystalline and noncrystalline polymers to crystallize and align to increase the strength of the polymer. Polyester, nylon, fluoroplastics, and polyurethane can also be dried in rotating-cone

vacuum dryers. Such dryers can handle flakes, chips, pellets, and crystals. The slow tumbling action does not cause changes in the shapes of particles. Units are commercially available in volume from 0.2 to 350 ft^3 . For solvent-wet materials, closed-system operation including solvent recovery is possible. Higher maintenance costs are the limitations of such dryers.

Among the new types of dryers suited for polymers, one may cite the centrifugal pellet dryer marketed by Gala Industries (Eagle Rock, VA), which can be used to dry polyethylene, PP, polyester, rubber, and so on in three distinct phases: (1) predewatering; (2) impact dewatering; and (3) air drying. Up to 95% of the water is removed by impact and gravity through vertical perforated plates in the first stage. The predewatered pellets are fed into a turbinelike rotor encased in stationary cylindrical screens. As the pellets move spirally (in the second stage) from the bottom to the top of the rotor, the water content is reduced to a value between 0.5 and 1%. Finally, air drying reduces the moisture to below 0.05% in the upper part of the rotor as air is forced through the moving pellets.

Numerous papers have appeared in recent years on drying of polymers and resins. Shah and Aroara [17] have reviewed the state of the art of drying suspension-PVC. They compare, in depth, continuous FBDs, rotary dryers, and cyclone dryers. The impact on energy consumption, maintenance costs, and product quality is assessed and compared. They show that FBD with immersed heat exchanger has some limitations when several grades of polymers are to be dried with frequent grade changes.

Readers interested in mathematical modeling of polymer drying may refer to Vergnaud [18].

Following is an example of how selection of the dryer is affected by quality of the dried product that may be used as raw material to produce different consumer products. Shah and Arora [19] have surveyed the various possible dryers used for crystallization/drying of polyester chips from initial moisture content of about 0.3 to 0.5% (wb) to under 50 ppm. Aside from low average moisture content it is also necessary to ensure uniform distribution of moisture, especially for certain products, e.g., production of thin films. The uniformity constraint is less severe if the chips are to be used to make PET bottles. [Figure 41.15](#) shows schematics of the crystallization/drying steps involved. Generally, it is a two-step process. The material is heat-sensitive. The initial crystallization/drying is faster than the drying step at low moisture levels. A two-stage dryer is indicated and is commonly used. It is possible to use different dryer types for each stage as shown in [Figure 41.16](#). A single dryer type (e.g., column or packed bed dryer with

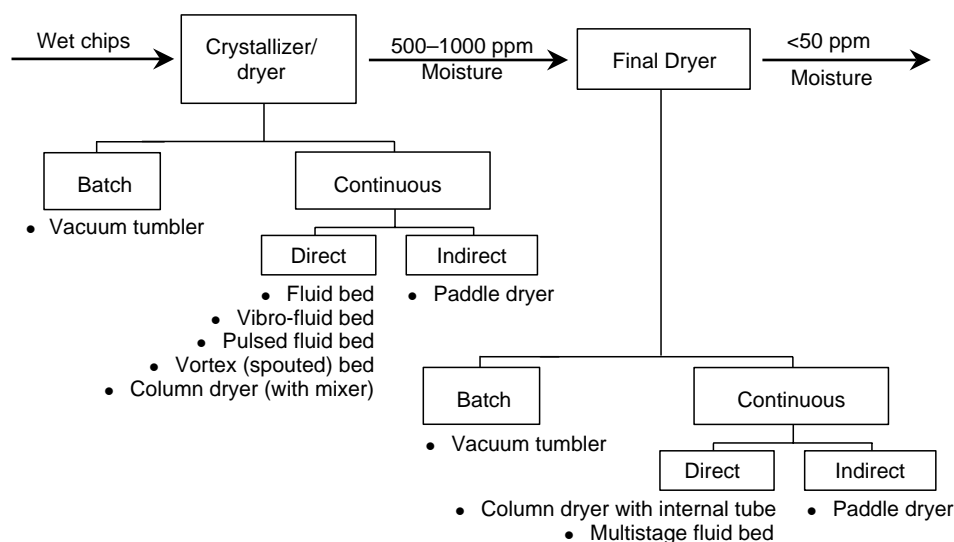


FIGURE 41.15 Schematic diagram of crystallization/drying steps in the production of polyester chips.

the chips moving downward slowly under gravity) is cheaper and hence recommended for the lower quality grade but a more expensive fluid bed followed by another fluid bed or column dryer may be needed for the higher quality grade. Note that numerous alternatives are possible in each case. It is also important to operate the dryers at the correct conditions of gas flow rate, temperature, and humidity. Dehumidified air is needed to achieve low final moisture contents in accordance with the equilibrium moisture isotherms of the product.

Another example of dryer selection is related to the choice of a suitable atomizer for a spray dryer. A spray dryer is indicated when a pumpable slurry, solution, or suspension is to be reduced to a free-

flowing powder. With proper choice of atomizer, spray chamber design, gas temperature, and flow rate it is possible to “engineer” powders of desired particle size and size distribution. Table 41.2 shows how the choice of the atomizer affects chamber design, size, as well as energy consumption of atomization and particle size distribution. The newly developed two-fluid sonic nozzles appear to be especially attractive choices when nearly monodisperse powders need to be produced from relatively moderate viscosity feeds (e.g., under 250 cp) at capacities up to 80 t/h by using multiple nozzles. More examples may be found in Kudra and Mujumdar [21].

New dryers are being developed continuously as a result of industrial demands. Over 250 U.S. patents

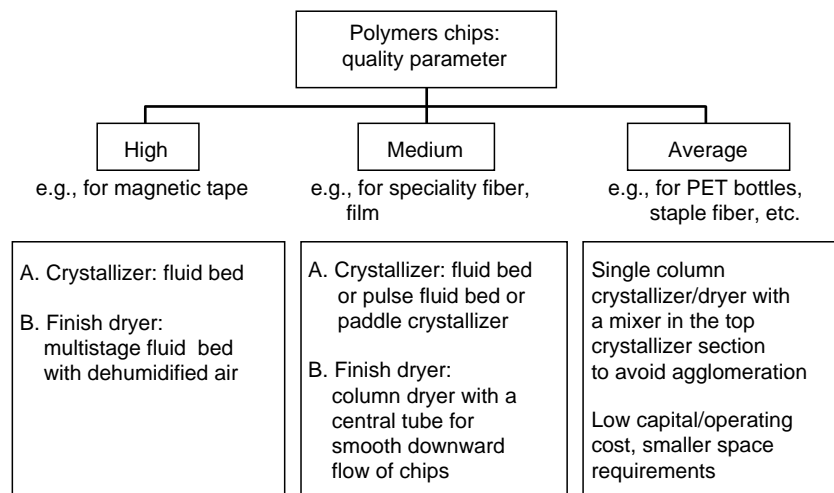


FIGURE 41.16 Possible dryer types for drying of polyester chips.

TABLE 41.2

Spray Drying of Emulsion-PVC. Effect of Selection of Atomizer on Spray Dryer Performance: A Comparison between Different Atomizers

Parameter	Rotary Disk	Two-Fluid (Sonic)	Two-Fluid (Standard)
Dryer geometry	Conical/cylindrical $H/D \approx 1.2\text{--}1.5$	Tall-form cylindrical $H/D \approx 4$	Tall-form Cylindrical $H/D \approx 5$
Evaporation capacity (water) (kg/h)	1600	1600	1600
Chamber ($D \times H$) (m)	6.5×8	3.5×15	3×18
Number of nozzles	1,175-mm disk 15,000 rpm	16 nozzles 4 bar pressure	18 nozzles 4 bar pressure
Power for atomizer (W/kg slurry)	25	20	80
Capital cost	High	Medium	Medium
Operating cost	Medium	Low	High

are granted each year related to dryers (equipment) and drying (process); in the European Community about 80 patents are issued annually on dryers. Kudra and Mujumdar [21] have discussed a wide assortment of novel drying technologies, which are beyond the scope of this chapter. Suffice it to note that many of the new technologies (e.g., superheated steam, pulse combustion—new gas-particle contactors as dryers) will eventually replace conventional dryers in the next decade or two. Among the more popular new dryers for polymers is the pulsed bed dryer. This dryer uses pulsating motion imparted to the bed of particles by periodically relocating the fluidized region of the vessel. This type of dryer has been claimed to have a higher efficiency and lower air consumption for fluidization and for drying. It is discussed in detail by Kudra and Mujumdar [21]. New technologies are inherently more risky and more difficult to scale-up. Hence, there is natural reluctance to their adoption. Readers are encouraged to review the new developments in order to be sure that their selection is the most appropriate one for the application at hand.

It is well known that most polymers leaving the polymerization reactor contain various but small amounts of unreacted monomer, solvents, water, and/or various reaction by-products. The presence of these volatiles in the polymer is undesirable. Their concentrations may range from several parts per million to several tens of percentage. Their separation from bulk polymer is necessary to improve polymer properties, to recover monomer and solvents, to meet health and environmental regulations, to eliminate odors, and/or to increase the extent of polymerization. This process of devolatilization is usually performed above the glass transition temperature of the polymer. The reader is referred to Albalak [20] for

detailed discussion of the theory of devolatilization and various devolatilizing equipments.

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42 Drying of Enzymes

Ana M.R. Pilosof and Virginia E. Sánchez

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42.1 INTRODUCTION

Enzymes are protein catalysts of high molecular weight, which are produced not only by plants and animals but also mainly by microorganisms as a result of fermentation processes. Enzymes fall into two categories: (1) bulk industrial enzymes, which mainly include proteases for detergents, amylases for textile desizing and starch hydrolysis, pectinases for fruit-juice clarification, and proteases for the leather industry (Table 42.1); and (2) analytical enzymes.

In general, enzymes have a disadvantage in that they are deactivated due to heat-induced structural changes or, in the case of proteolytic enzymes, due to decomposition by themselves. It is therefore desirable to distribute and use enzyme preparations in the form of solids, such as powders and granules, instead of liquids. Although drying itself is a valuable tool in the improvement of the enzyme storage stability, the process step itself often causes a substantial loss of activity and the final product is still susceptible to inactivation.

Such solid enzyme preparations are conventionally produced by means of freeze drying or spray drying. As freeze-drying process is unsuitable for large-scale productions, spray drying is used as the most fitted process for the industrial mass production of solid enzyme preparations. Especially in the case of solid enzyme preparations to be used in detergents, spray-drying process is most frequently used.

Analytical enzymes are invariably dried in small quantities by freeze drying or by spray drying using low temperatures. The spray drying of pancreatin, for example, must not have inlet drying temperatures above 95°C [1].

The design of a proper drying process should guarantee a high level of active enzyme. Generally, enzyme activity after drying is a function of the composition of the initial liquid to be dehydrated, the process parameters, and the physicochemical characteristics of the enzyme [2], so that drying of each enzyme product should be considered on an individual basis.

Relevant properties of dried enzymes are listed in Table 42.2. By modifying the spray-drying process, it is possible to alter and control the properties that are mentioned earlier for spray-dried enzymes.

A major concern in spray drying of enzymes is the retention of their activities, whereas this complication is not seen in the case of purely chemical systems. Therefore, the enzyme activity retention must be close to 100% in the spray-drying operation; and moreover, the shelf life of the dried enzyme products must be excellent, i.e., enzyme activity must be retained for long-time storage.

The formation of dust during handling of enzyme preparations in finely divided solid forms is also a problem. The dust of enzyme preparations incorporated in detergents and washing compositions could be dangerous to the health of the workers in detergent factories and for the end users.

TABLE 42.1
Major Industrial Applications of Industrial Enzymes

Application	Enzyme	Source
Detergents	Protease	<i>Bacillus</i>
	Amylase	<i>Bacillus</i>
	Lipase	<i>Humicola</i> , <i>Pseudomonas</i>
Starch industry	Cellulase	<i>Bacillus</i> , <i>Humicola</i>
	Amylase	<i>Bacillus</i>
	Glucoamylase	<i>Aspergillus</i>
	Glucose isomerase	<i>Bacillus</i> , <i>Streptomyces</i>
Dairy	Protease	<i>Rhizomucor</i>
	Lipase	<i>Aspergillus</i>
	Lactase	<i>Klyveromyces</i> , <i>Aspergillus</i>
	Sulfhydryloxidase	<i>Aspergillus</i>
Wine and Juice	Pectinase	<i>Calf Stomach</i>
	Cellulase	<i>Aspergillus</i>
	Cellobiase	<i>Aspergillus</i> , <i>Trichoderma</i>
	Glucose oxidase	<i>Aspergillus</i>
Distilling industry	Polyphenol oxidase	<i>Aspergillus</i>
	Amylase	<i>Trametes</i>
	Glucoamylase	<i>Aspergillus</i>
Brewery	β -glucanase	<i>Aspergillus</i>
	Acetoacetate decarboxylase	<i>Aspergillus</i> , <i>Bacillus</i>
	Amylase	<i>Bacillus</i>
Bakery	Protease	<i>Aspergillus</i> , <i>Bacillus</i>
	Glucose oxidase	<i>Aspergillus</i>
	Amylase	<i>Aspergillus</i>
	Cellulase	<i>Bacillus</i>
Textiles	Catalase	<i>Trichoderma</i> , <i>Humicola</i>
	Phytase	<i>Aspergillus</i>
	Cellulase	<i>Aspergillus</i>
	Plant cell wall-degrading enzyme	<i>Trichoderma</i> , <i>Humicola</i> , <i>Aspergillus</i>
Animal feed	Xylanase	<i>Aspergillus</i>
Pulp and paper	Protease	<i>Trichoderma</i> , <i>Bacillus</i>
Leather	Tannase	<i>Aspergillus</i>
Tea		

Source: From Oxenboll, K., *Aspergillus* enzymes and industrial uses, in *The Genus Aspergillus*, Powell, K. ed., Plenum Press, New York, 1994, 147–153.

There are relatively few studies on the drying of enzymes in the chemical and biological literature. This could partly be due to the industrial nature of the subject with the concomitant proprietary knowledge and confidentiality agreements about products and specific process parameters. However, a large number of patents exist in this field. Selected patents where drying plays the dominant role in the manufacture of enzyme products are listed in [Table 42.3](#).

TABLE 42.2
Quality Parameters for Dried Enzyme Products

Retention of activity
Dust properties
Solubility and dispersibility
Enzyme stability <i>per se</i>
Enzyme stability in detergent
Flow properties
Enzymes protein purity
Mean particle size and particle size distribution
Homogeneity
Bulk density
Color
Odor

42.2 SPRAY DRYING

Spray drying is a convective drying technique that uses hot air to transfer the heat and remove the evaporated water. It is a short-time process in the range of few seconds; and if processing conditions are optimized and stabilizers are added, it is suitable even for heat-sensitive enzymes. The process may be summarized in three phases: (1) spray formation, (2) drying, and (3) air–powder separation.

A wide range of feed and drying properties are successfully handled to produce the powder with desired qualities. The wide application of spray drying has been due to the flexibility of the system and the development of different drying-chamber designs that combine with rotary or nozzle atomizers to handle particular products in continuous operations.

If the desired powder specification cannot be achieved in the single-stage process, multistage operations are necessary. By combining atomization, fluidization, and agglomeration technologies in advanced spray-dryer designs, it is possible to meet the quality specifications of the end product, enzyme, within a safe, hygienic, and environmentally friendly process.

When the spray-drying operation is cocurrent, i.e., hot air introduced into the dryer close to the atomizing device, there is less danger of overheating as the evaporation rates are high (34 to 160 kg/h/m² of particle area) [3]. Thus, cocurrent drying chambers are preferred to minimize heat deactivation of enzymes during the process.

In the case of enzymes that are particularly sensitive to oxidation, it is preferable to use inert gas, such as nitrogen, during spraying and drying [4].

42.2.1 DRYING-CHAMBER LAYOUTS

Several drying-chamber designs may be used to attain the desired powder specifications. Standard cocurrent

TABLE 42.3
Applications of Enzyme Drying

Issue	Patent No.	Patentee	Date
Process for freeze-drying enzymes	US 4 180 917	Rohm and Haas Co., Philadelphia, PA	January 1, 1980
Spray drying with additives	US 4 233 405	Rohm and Haas Co., Philadelphia, PA	November 11, 1980
Enzyme spraying onto a heated fluidized bed of inert particles	US 4 617 272	Economics Laboratory Inc., St. Paul, MN	October 14, 1986
Disaccharide-stabilized enzyme preparation	EP 0 501 375 A1	KAO Corporation, Tokyo, Japan	July 2, 1992
Spray auxiliary composed of hydrophobic silicas	US 5 318 903	BASF, Aktiengesellschaft, Germany	June 7, 1994
Process for storage of materials	US 5 928 469	Inhale Therapeutic Systems, San Carlos, CA	July 27, 1999
Enzyme preparation for leather	RU 2 127 311	Sergeevna et al.	March 10, 1999
Salt-stabilized enzyme preparations	US 5 972 669	Gist-brocades, B.V., The Netherlands	October 26, 1999
Microgranular enzyme composition	US 6 120 811	Genencor International Inc., Palo Alto, CA	September 19, 2000

conical chambers (Figure 42.1a) with either rotary atomizer or nozzles enable both fine- and coarse-particled powders to be produced. This layout is used for thermostable enzyme products where fairly high outlet air temperatures may be used. For a single-stage spray drying, enzyme activity losses may be

significant. Table 42.4 shows the effect of some spray-drying parameters on the loss of activity of several enzymes. Enzyme activity decreases with increasing outlet air temperatures. Low outlet temperatures were also found to be important for the preservation of tyrosinase, glucose oxidase, β -galactosidase, alkaline phosphatase, and lactate dehydrogenase (LDH) [5–7].

Enzyme activity also decreases with increasing inlet temperatures. Significant enzyme inactivation occurs when moisture decreases below 15 to 20%.

The loss of enzymatic activity can be reduced and even almost completely avoided when a controlled combination of spray drying and fluidized-bed drying is used.

Spray driers with integrated fluid bed represent the latest in the spray-dryer design whereby the completion of drying is accomplished in a fluid bed located in the lower cone of the chamber. The operation of the fluid bed enables lower outlet temperatures to accomplish completion of drying, leading to lower powder temperatures and higher dryer thermal efficiencies. The integration of fluid beds into drying chambers allows to produce, under low product temperature conditions, nondusty, free-flowing, coarse powders of individual particles or agglomerates.

Chamber designs are shown in Figure 42.1b and Figure 42.1c. Figure 42.1b shows a modification of the standard layout where the fluid bed located in the base of the drying chamber is of an annular design enabling the exhaust air to be ducted out through the center of the chamber base.

The chamber in Figure 42.3c has cyclone-like dimensions, i.e., a short cylindrical side in relation to an extended cone section, and has a mixed flow concept with air entering and leaving the top of the chamber. This chamber is ideal for heat-sensitive, sticky products and has enabled many products to

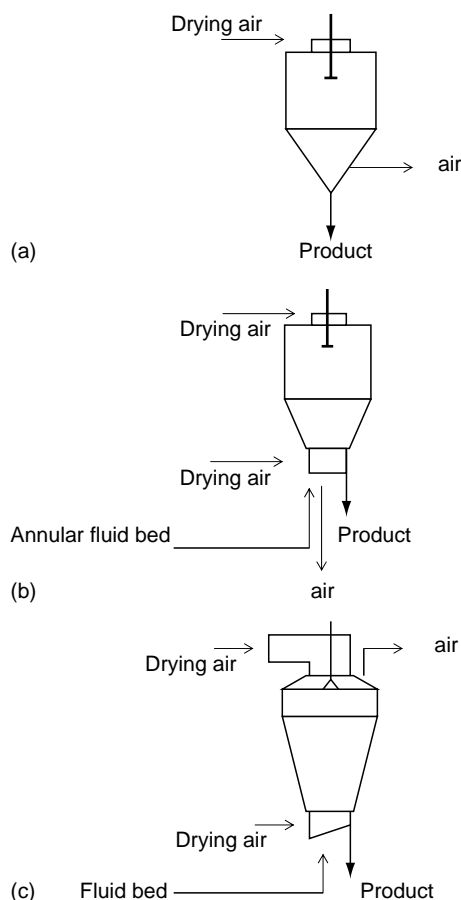


FIGURE 42.1 Drying-chamber designs.

TABLE 42.4
Effect of Some Spray-Drying Parameters on the Loss of Enzyme Activity

Enzyme	Spray Dryer	Temperature (°C)		Moisture Spray-Dried Product (%)	Loss of Activity (%)	Reference
		Inlet	Outlet			
Bacillus alkaline protease (Alcalase, Novozymes)	Conventional spray-drying tower with spinning disk as atomizer	131.0	73.0	10.2	2.7	British patent 1 360 969
		131.0	70.0	7.6	13.6	
		145.0	40.0	26.8	1.5	
		146.0	57.0	15.3	7.8	
Fungal α-amylase prepared by reverse osmosis	Conventional spray-drying tower with spinning disk as atomizer	150 ± 5	63.0	5.2	19.0	British patent 1 360 969
		150 ± 5	66.0	8.3	10.0	
		150 ± 5	78.0	4.4	26.0	
		150 ± 5	80.0	2.1	38.0	
Fungal amyloglucosidase	Conventional spray-drying tower with spinning disk as atomizer	150 ± 5	90.0	14.0	34.0	British patent 1 360 969
		150 ± 5	80.0	17.0	20.0	
		150 ± 5	70.0	20.0	10.0	
<i>Aspergillus oryzae</i> protease	Small-scale spray drier	160.0	75.0	14.1	72.7	US patent 4 233 405
		120–130	70.0	9.4–10.1	28–49	
Neutral fungal protease prepared by ultrafiltration	Small-scale spray drier	154.5	79.5	6.5	21.0	US patent 4 233 405
<i>Bacillus subtilis</i> neutral bacterial protease prepared by ultrafiltration	Small-scale spray drier	154.5	76.6	5.07	22.3	US patent 4 233 405

be spray dried successfully for the first time. Although it has been developed for food and dairy products, it is also used for dyestuffs, agrochemicals, polymers, and detergents [8].

This system features a rotary atomizer or nozzle located in an air-dispersed roof. Primary drying takes place in the conventional manner, but the partially dried product, still having significant moisture content, passes directly into an established fluidized layer.

As it has been found that generally the loss in enzymatic activity increases when the water content in the spray-dried enzyme concentrate is lowered (Table 42.4), it is desirable that the product leaving the spray dryer has a moisture content of not less than 8 to 10%, preferably about 20%. This results in extremely low outlet temperatures from the drying system combined with controlled second-stage drying that takes place within the static fluid bed at a lower temperature (40 to 50°C) to achieve the desired moisture content. The product from the dryer can be postcooled or postdried in a vibrating fluid bed.

A novel feature of the design is the removal of the exhaust air from the roof of the drying chamber [8]. Fine materials elutriated from the fluidized bed are carried within the airflow and act to powder the surface of the drying chamber thereby limiting deposit formation. A proportion of the elutriated fines comes into contact with the cloud of atomized droplets resulting in agglomeration prior to their entry into the fluidized layer. The fine materials separated from the exhaust air are recycled to the chamber for further agglomeration. The process produces a dustless free-flowing powder with a mean particle size ranging from approximately 100 to 700 μm depending upon the product characteristics.

42.2.2 SOLUTE-INDUCED PROTECTION IN SPRAY DRYING

Several methods have been applied in the field of stabilizing enzymes against losses during drying and subsequent storage and handling. The bulk of these applications are concerned with the addition of carbohydrates and, more specifically, sugars, polyols, and salt components to the enzyme concentrate [9].

Also known is the inclusion of components into the formulation with the aim to produce a glassy product at storage temperature, thus improving enzyme stability. It will generally be preferred to employ a carrier substance, which must be hydrophilic (either water-soluble or water-swelling), that must exist in a glassy amorphous state with a glass-transition temperature (T_g) above 20°C (a desirable range is therefore between 60 and 200°C) and should be sufficiently chemically inert [10]. However, significant enzyme

inactivation occurs also in formulations kept well below their glass-transition temperatures [9].

During the final stages of air drying, the major stress that must be overcome is the removal of the enzyme's hydration shell, which, for at least some labile enzymes, can result in irreversible inactivation upon rehydration [11]. The mechanism of this level of protection is different from that occurring in solution [12]. It has been suggested that sugars can continue to protect the dried protein by hydrogen bonding to the protein at some critical point during dehydration [13,14], thus serving as water substitutes when the hydration shell of the enzyme is removed.

Moreover, most researchers agree that protection by compounds such as lactose and trehalose depends on the formation of an amorphous phase with the protein [15]. The proteins are mechanically immobilized in the glassy, solid matrix during dehydration. The restriction of translational and relaxation processes is thought to prevent protein unfolding, and spatial separation between the protein molecules (i.e., dilution of protein molecules within the glassy matrix) is proposed to prevent aggregation [12].

Taken together, these studies support the conclusion that the importance of amorphous behavior of the protein and the additive is that it allows for effective hydrogen bonding between the additive and the protein. A glassy additive that does not have the interaction will not protect the protein against dehydration damages [12].

Sugars are an important group of glass-forming substances, which are also good stabilizers during drying. Among them maltose- and trehalose-type disaccharides are preferred. Examples of the maltose-type disaccharides include maltose, cellobiose, gentiobiose, melibiose, and lactose, and examples of the trehalose-type disaccharides include trehalose, isotrehalose, sucrose, isosucrose, etc. The amount of the disaccharide to be included in a solid enzyme preparation may vary depending on the type of the enzyme used, but generally it is more preferably from 10 to 100% by weight of the enzyme [16].

An addition of 0.5% lactose in a *Bacillus* alkaline protease spray-drying process is enough to improve the recovery of active enzyme after drying and thermal treatment at 90°C (Table 42.5). In the same way, the exhaust air temperature could be increased by 10°C reaching severe drying conditions without significant loss of protease activity (Table 42.6). The solid enzyme preparation obtained in this process was also excellent in resistance to mechanical pressure [16].

In addition to plain carbohydrates, other polyhydroxy compounds can be used, e.g., carbohydrate derivatives and chemically modified carbohydrates (i.e., carbohydrates that have undergone chemical reaction

TABLE 42.5
Effect of Lactose Addition on Production
of Stabilized Enzyme Powders

Lactose Addition (%)	Residual Activity (%) after Drying	Residual Activity (%) after Heating 1 h at 90°C
0.0	83	84
0.5	90	95
1.0	96	96
2.5	96	96
5.0	95	97

Patent reference: EP 0 501 375 A1.

Enzyme: Bacillus alkaline protease K 16 5% + calcium chloride 0.2% + sodium sulfate 2.5%.

Drying technology: Atomizer-type spray dryer T^{in} : 150°C; T^{out} : 60°C.

to alter substituents on the carbon backbone of the molecule but without alteration of the backbone).

Proteins are also suitable. Thus albumin can be used and also hydrolysis products of gelatin, like Byco A (Croda Colloids Ltd.) that allowed an excellent enzyme storage stability after hard spray-drying conditions of LDH type XI (ex rabbit muscle) (Table 42.7).

Sugar copolymers may be employed as glass-forming substances. Ficoll (Pharmacia[®]) includes copolymers with molecular weights between 5000 and 1,000,000, containing sucrose residues linked through ether bridges to bifunctional groups. Such groups may be alkylene of two or more carbon atoms but not normally more than ten carbon atoms. The bifunctional groups serve to connect sugar residues together. These polymers may, for example, be made by reaction of the sugar with a halohydrin or a bis-

TABLE 42.6
Effect of Lactose Addition on Production of
Stabilized Enzyme Powders by Spray Drying with an
Exhaust Air Temperature Variation from 60 to 70°C

Lactose Addition (%)	Residual Activity (%) after Drying		
	60°C	65°C	70°C
0	83	79	70
0.5	96	99	96
1.0	94	95	95

Patent reference: EP 0 501 375 A1.

Enzyme: Bacillus alkaline protease K 16 5% + calcium chloride 0.2% + sodium sulfate 2.5%.

Drying technology: Atomizer-type Spray Dryer T^{in} : 150°C.

epoxy compound [10]. Ficoll 400 DL (Pharmacia) is a water-soluble copolymer of sucrose and epichlorohydrin that has a T_g of 97°C. A mixture of Ficoll 400 DL 4% and LDH type XI (ex rabbit muscle) shows that enzyme activity was effectively preserved through the spray-drying procedure and subsequent long-term storage. Good stabilization was also reached for alcohol oxidase: after 30 d, the spray-dried material retained 90% of its activity while the freeze-dried material lost all activity in 20 d (Table 42.7).

Another approach is the addition to the formulation of one or several components that are able to bind moisture. This will reduce the water activity of the final preparation or temporarily prevent the interaction of water penetrating from the surroundings with the enzyme itself.

The use of organic and inorganic salts as a processing aid (e.g., to improve flowing behavior of the product) or as a bulking/standardizing agent is well known. Solid enzyme formulations with improved drying yield and storage are achieved by preparing a solution comprising an enzyme and a water-soluble inorganic salt. The presence of salt prior to drying, while the enzyme is still in solution, results not only in a higher yield during drying but also in an improved storage stability as well as processing stability of the obtained dry enzyme preparations. Addition of magnesium sulfate to different industrial enzymes allowed a recovery of more than 100% active enzyme after drying and after long-term storage [17].

Preferably, an inorganic salt of a divalent cation, like zinc or magnesium sulfate, can be added to the enzyme solution. Also, a combination of salts as well as a combination of enzymes can be used. The addition of divalent cations is preferred because they provide the best storage and processing stability. Sulfate is preferred as anion because it provides the best drying yield (Table 42.8).

Drying of a solution containing the enzyme and the salt will result in a solid composition that is homogeneous with respect to the distribution of the enzyme and the salt. The stabilizing effect of the salt increases with increasing dosage of the salt to the enzyme solution, until at a certain point further increases in salt dosage no longer produce further improvement of the enzyme stability. For this reason, between 5 and 90% of salt is added to the enzyme solution based on the weight of the enzyme in solution.

42.3 POWDERED DETERGENT ENZYMES

Enzymes today are key strategic ingredients for washing and cleaning formulations. Enzymes not only remove stains but also improve textile fiber properties.

TABLE 42.7
Solute-Induced Protection of Enzymes during Spray Drying and Storage

Enzyme (w/w %)	Brand Name	Additive (w/w %)	Drying Technology	Temperature (°C) Inlet–Outlet	Residual Activity after Drying (%)	Storage		Residual Activity after Storage (%)
						Time Temperature (Weeks)	(°C)	
1. Fungal phytase (11)	Gist-brocades	MgSO ₄ ·7 H ₂ O (24)	Buchi lab-scale spray drier	130–85	—	8	30	
2. Bacillus alkaline protease (12)	Genencor International Inc.	MgSO ₄ ·0 H ₂ O (8.5)	Niro STREA-1 lab-scale fluid-bed coater	80–50	119	6	35	121
3. Trichoderma β -glucanase + endoxylanase (25)	Gist-brocades	MgSO ₄ ·7 H ₂ O (24)	Glatt WSG-60 fluid-bed coater	80–50	105 + 120	12	30	111 + 120
4. Lactate dehydrogenase type XI (0.005)	Sigma Chemical Co.	Ficoll 400 DL ^a (4)	Drytec pilot-scale spray drier	210–70	82	20	10–35	100
5. Lactate dehydrogenase type XI (0.005)	Sigma Chemical Co.	Ficoll 400 DL ^a (10)	Lab-Plant SD-04 spray drier	170–75	91	9	10–35	103
6. Lactate dehydrogenase type XI (0.005)	Sigma Chemical Co.	Byco A ^b (4)	Drytec pilot-scale spray drier	210–70	88	15	25	113
7. Alcohol oxidase (0.00005)	Provesta Enzymes	Ficoll 400 DL ^a (4)	Drytec pilot-scale spray drier	150–70	52	4	35	90

^aCopolymer of sucrose and epichlorohydrin (Pharmacia Reg. Trade Mark).

^bCold water-soluble protein obtained from gelatin by enzymatic hydrolysis (Croda Colloids Ltd.).

Patent references 1–3: US 5 972 669.
 4–7: US 5 928 469.

TABLE 42.8
Effect of Salts Addition on Spray Drying and Storage
Losses of Fungal Phytase

Salt Type	Spray Drying Losses (%)	Storage Losses (%) after 8 Weeks at 35°C
None	6	52
Magnesium sulfate	7	15
Magnesium chloride	26	43
Magnesium nitrate	32	27
Zinc sulfate	5	9
Zinc chloride	48	5
Calcium chloride	40	18
Calcium nitrate	44	13
Sodium sulfate	11	51
Potassium sulfate	17	36
Ammonium sulfate	6	46

Patent reference: US 5 972 669.

Enzyme: Fungal phytase concentrate (Gist-brocades) 17 w/w % + 800 mM salt.

Drying technology: Lab-scale Buchi 190 mini spray dryer T^{in} : 140°C; T^{out} : 80°C.

Prior to the introduction of compacts, the use of enzymes in detergents was limited primarily to one class of enzymes—the proteases. Proteases catalyze the hydrolysis of protein-based soils like blood and grass. Most powder and liquid laundry detergents on the market today, both low density and compacts, employ a protease [18]. Recently, protein engineering has been used to construct detergent proteases with improved stability and performance characteristics.

In addition to proteases, a limited number of brands employ amylases. Detergent amylases catalyze hydrolysis of the α -1,4-glycosidic linkages in starch. As such they show benefits on a number of common food soils like gravies, sauces, pastas, and baby foods.

Advances in genetic and protein engineering have led to new classes of enzymes with novel benefits for use in compact products. In 1988, lipase appeared in one of the first compact powders to hit the Japanese market. Since then lipase has found broad application in the global detergent market. Second-generation lipases with improved cleaning efficiency were developed in the compact detergent market [19].

Products containing up to four different enzymes—protease, amylase, lipase, and cellulase—are now on the market. The patent literature suggests that even more novel detergent enzymes are on the way.

As detergent enzymes have now become commodities and are rather low-priced products, full attention

is paid to the production process that, despite all the technical demands, has to be very economical.

The very first enzyme products that were introduced on the market in the 1960s were powders, and using today's terminology they contained 100% dust. Today the powders have been replaced by various types of granulates. However, certain kinds of agglomerates, e.g., fluidized-bed agglomerates, are still available. These agglomerates have an acceptable particle size distribution but the physical strength is normally poor.

To protect both detergent plant operators and end users, the enzyme particles have to be coated in such a way that no active enzyme dust is present or released during handling. Besides avoidance of dust, the coating has to stabilize the particles against abrasive forces and protect the granule against chemical agents such as water, hydroperoxide, and peracetic acid (the bleach that migrates in detergent formulations). On the other hand, the coating should not be sticky because this would result in oversized particles during the coating process and would deteriorate the free-flowing properties of the particles in the dosing systems of detergent manufacturers [20]. Finally, the coating has to dissolve readily in the washing liquor, even at low temperatures.

Since the introduction of enzymes into the detergent and other industrial segments, many developments have been made regarding the granulation and coating of enzymes to reduce enzyme dust. However, in today's state of ever-increasing environmental concern and heightened awareness of industrial hygiene, there remains a continuing need for low-dust enzyme granules. The following are the most important processes to granulate enzymes [20].

42.3.1 GRANULATION BY FAST MIXER SYSTEMS

With fast-rotating mixing systems of “ploughshare” type mixers or Schigi type with horizontal or vertical shafts, equipped with blenders, the high turbulence in the rotating mixture of ingredients determines the particle size.

Usually, a premix of dry powders is loaded into the equipment and the liquid enzyme concentrate is injected and mixed. After a certain time, depending on the recipe and moisture content, granules are formed that are discharged and dried in a fluid dryer. After sieving the granules, over- and undersized materials are milled and recycled to the premix.

42.3.2 GRANULATION BY THE PRILLING PROCESS

The basic principle of this process (Figure 42.2) is that the total mixture of ingredients is transferred into a

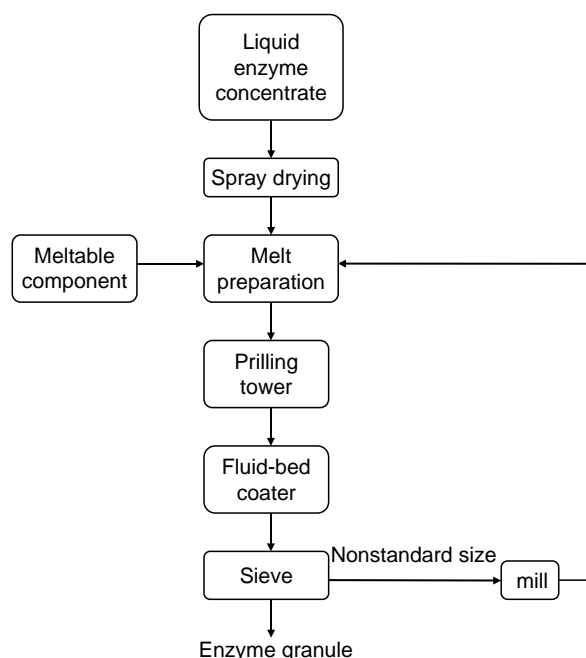


FIGURE 42.2 Prilling process.

molten mass of low viscosity in which insoluble ingredients have to be homogeneously distributed. The melt is pumped through insulated tubes to the top of a tower where it is sprayed by nozzles or a rotating disk.

The size of the droplets that fall down the tower is determined by the diameter of the nozzle, the rotation frequency of the disk, the surface tension, and the viscosity of the melt. The heat capacity and the melting heat of the droplets dictate the distance needed to solidify the droplets to nearly ideal balls and, therefore, also the height of the prill tower. The final steps are cooling in continuous fluid beds and sieving of the solidified prills. Over- and undersized materials are separated and recycled.

With this process, it is of some disadvantage that all the ingredients, including the enzyme, have to be anhydrous. Therefore, the enzyme has to be brought into a dry state, which is often costly because of energy demand, enzymes losses, and inactivation because of the (high) drying temperature. The preparation of the melt at elevated temperatures also inactivates a certain amount of the enzyme, which additionally increases the cost of the process.

As prilling agents, meltable ingredients, such as polyethylene glycols, are used as binders. Salts are used to make the particles brittle and tough, and they combine the advantage of being cheap.

42.3.3 GRANULATION BY EXTRUSION PROCESS

The extrusion process is very well established in the plastic and food industry. This technology for the

manufacturing of the enzymes-containing granulate combines the advantage of a homogeneous particle size distribution with low-cost ingredients.

After mixing the dry ingredients, the dry premixture and the liquid enzyme concentrate are fed batchwise in a mixing system to obtain a moist doughlike mixture. This mass is fed into a twin-screw screen-type extruder and presses through thin perforated metal sheets with holes that are the diameter of the desired particle size. The extruded noodles fall by batchwise feeding into a spheronizer. This machine, equipped with a fast-rotating disk, breaks the noodles down to cylindrical particles, which are then transported on the disk to the walls of the apparatus.

After drying the particles in a fluid-bed dryer and removal of over- and undersized materials by sieving, the beads are coated in a fluid-bed coater with one or two coating layers to obtain white or colored beads and to have a tough protecting layer that completely avoids the development of enzymatically active dust.

42.3.4 GRANULATION BY SPRAY COATING OF CORE PARTICLES

Granulation by spray coating generally refers to producing a particle having an average size between 20 and 400 μm by fluidizing a core material in a heated airstream to pass through an area of atomized liquid (Figure 42.3). The atomized liquid droplets, which contain dissolved or suspended solids, form a film on the surface of the core material. The coated core material is then transferred from the spray zone into a drying zone. The solvent in the liquid—generally water—is dried, leaving the dissolved or suspended solids as a film on the core material. This process is continued until the desired amount of film is formed.

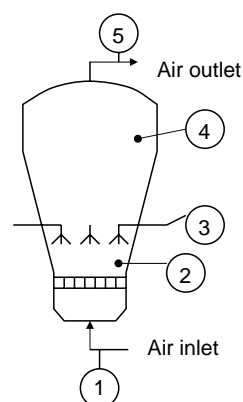


FIGURE 42.3 Principle of the core-coating fluid-bed apparatus: (1) air inlet; (2) product bowl; (3) spray nozzles; (4) expansion chamber; (5) air outlet.

TABLE 42.9
Typical Carriers and Binders Used in Granulation of Enzymes by Spray Coating

Binders	Carriers Used as Core Materials
Alginate	Sodium chloride
Carrageenan	Sodium carbonate
Cellulose fibers	Urea
Gelatins	Calcium alginate
Xanthan gum	Saccharose beads
Locust bean gum	Polyvinyl alcohol beads
Gellan gum	Starch
Soluble or hydrolyzed starch	Soy flour, guts
Polyethylene glycols	Corn flour
Ethoxylated fatty alcohols	Cellulose-type materials
Polyvinyl alcohols	
Polyvinyl pyrrolidones	
Ethoxylated phenols	

Finally, the coated particles, which have an onion-like structure, may be screened to obtain the desired range of particles.

Methods for making low-dust granules include:

1. Loading a suitable carrier as a core material into a fluid-bed granulator
2. Blending an aqueous enzyme source and one or more suitable binders
3. Spraying the blend of enzyme and binder of step 2 on the carrier
4. Spraying the product of step 3 with a water-soluble, food-grade polymer at a rate to form a coating and to reach the desired particle size

Typical carriers used as core materials and binders are listed in Table 42.9. As used herein, "binder" indicates one or more materials, which act either alone or in combination with sugars (such as sorbitol) to bind the enzymes to the carrier material.

Two objectives are attained during the spraying process: (1) the enzyme is attached to the carrier and (2) the particle is built up to a granular form (within the desired size range). A suitable food-grade polymer is then sprayed onto the granulated particles to envelope the enzyme and to hold the agglomerate or granule together.

This process is economically attractive because the moisture sprayed onto the carrier is "flashed off" as the liquid is sprayed on the carrier, and thus a large amount of aqueous enzyme can be loaded on the carrier.

Atomizing spraying can be done countercurrently (down spray) to the fluidizing air or cocurrently (up spray) with the fluidizing air. Down spray usually

results in more agglomeration and is useful when fine powders are coated to increase the particle size, resulting in lower dust granules. Cocurrent spraying (up spray) results in less agglomeration and is used when the core particle size already approximates the final product size [20].

Many variables affect the efficiency of the coating process. The three most important parameters for manufacturing the microgranules are bed temperature, fluidization air rate, and spray rate. Careful adjustment of all the engineering parameters is required to set the optimal conditions for a dust-free granulate of desired specifications.

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43 Drying of Coal

Jerzy Pikoń and Arun S. Mujumdar

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43.1 INTRODUCTION

Coal drying is of much theoretical and economic importance. Coal is a valuable fuel and raw material for many chemical synthesis processes that are becoming more and more important considering the increasing price of crude oil. Drying of coal is carried out to increase its calorific value and facilitate its transport. Wet coal is difficult to load or unload from railway cars owing to freezing, which is a problem in colder climates. The presence of moisture causes a reduction in friability of coal, makes it difficult to control blending operations, worsens the

quality of grinding (if coal is ground), and impedes separation and classification as well as the pneumatic transport of pulverized coal. Friable coal suitable for combustion in modern steam boilers is obtained only when the moist coal is dried. Coal must also be dried for the following processes: (a) briquetting; (b) coking; (c) gasification; (d) low-temperature carbonization; (e) liquid fuel synthesis, and others. The final moisture content requirement of coal is different depending on the process in which it is used. The following is a summary of approximate ranges of moisture content of coal required for various processes.

Hard coal:

- Coking processes (based on the ramming method), 8 to 12%
- Coking processes (based on the charring method), <8%
- Briquetting processes, <4%
- Low-temperature carbonization process, ~0%
- Hydrogenation process, ~0%
- Coal combustion process in the pulverized fuel-fired furnace, <2%
- Brown coal: Briquetting process, 8 to 18%
- Gasification process, 5 to 15%
- Low-temperature carbonization process, <15%
- Hydrogenation process, ~0%
- Coal combustion process in the pulverized fuel-fired furnace, 12 to 15%

Coal drying and preheating are of particular importance in coke production because at a relatively small investment cost (for the installation of dryers) it is possible to increase the production capacity of the coke ovens by about 30 to 50% in preheating and about 10 to 15% in drying. The preheating of coal in dryers makes it possible to utilize in the mixture a greater proportion of gas coals, which give coke characterized by better mechanical strength, very low grainability, and homogeneous graining. The moisture content is very important in determining the usability of coal for further technological processing. In coal, the moisture may be present in the form of surface or hygroscopic moisture. Surface moisture is not dependent on the coal type because it depends on the classification and washing processes performed in the coal mine and on the soaking process during transport and storage. The evaporation of surface moisture takes place in the first drying period at a constant drying rate.

Hygroscopic moisture depends on the rank of coal; it decreases with the age of coal. The evapor-

ation of hygroscopic moisture takes place in the second drying period at a falling drying rate.

43.2 TYPICAL DESIGNS OF DRYERS USED FOR COAL DRYING

Both direct and indirect dryers are in use for coal drying. Combustion gases or steam may be used as the heating medium. Steam-heated dryers (e.g., drum, tray, and tube dryers) are used mainly for drying of brown coals in the coal briquetting process. Frequently, coal drying is carried out in convection dryers, e.g.,

- Rotary dryers
- Pneumatic dryers
- Fluid-bed dryers with spouted bed
- Vibratory fluid-bed dryers
- Shaft dryers
- Mill-type dryers

43.2.1 ROTARY DRYERS

In many industrial plants, rotary dryers are used for the drying of coal and coal muds. Generally these dryers operate in the cocurrent mode to avoid the possibility of ignition. The drying medium is hot air or combustion gases derived from natural gas or coal combustion. A typical rotary dryer is shown schematically in Figure 43.1. The main component of this dryer is the steel shell (3) lined with a refractory lining and set up on rollers (7) by means of bandages (hoops, 10) located on the shell. The shell is rotated typically using a toothed gear (5). The shell is set up obliquely with a slope of 2 to 5° to the horizontal. Inside the shell, there are lifters fastened to the inner surface of the drum. During operation, these lifters lift the coal granules and shower them gradually inside the shell in the stream of the flowing heating gas. Various lifters are shown in Figure 43.2.

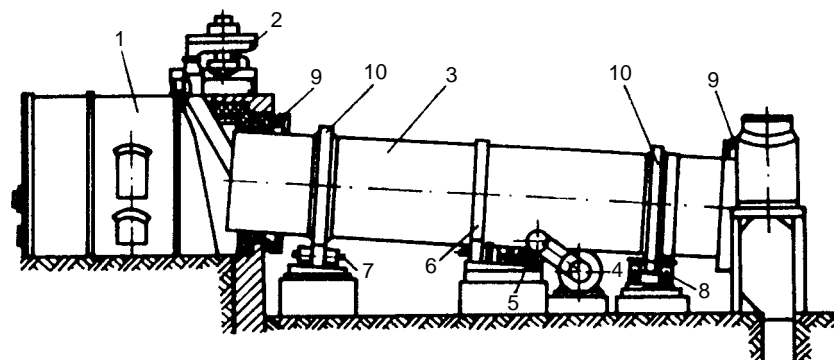


FIGURE 43.1 Schematic of a rotary dryer.

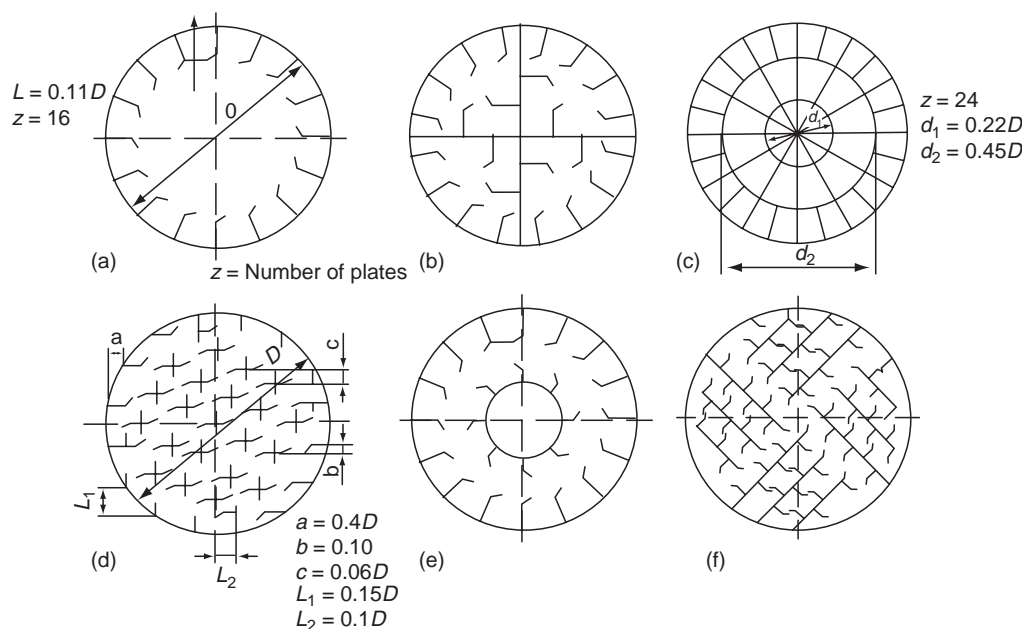


FIGURE 43.2 Scheme of the lifting-mixing devices for rotary dryers.

For drying of big lumps of coal that stick to the shell wall, type *a* devices may be used. For lumpy coals of low friability, type *b* devices may be used. For finer coal granules of high friability, the lifting-mixing devices are made in the form of separate cells (type *f*). For superfine coals (dust), a system of closed cells (type *c*) is used. The ends of the dryer shell are sealed labyrinthinely to the inlet and outlet heads. The combustion gases leaving the dryer carry coal dust, which may be up to 15% of the coal fed to the dryer. Hence the dryer exhaust is directed to cyclones or electrostatic precipitators to separate the entrained dust. The final loss of coal does not exceed 0.5 to 1.5%, depending on the type of equipment used for cleaning the combustion gases. Typical rotary drum dryers for coal drying are in the following sizes:

- (a) Shell diameter, 1.5 to 3.5 m
- (b) Shell length, 12 to 30 m or more

The drive motor power required depends on the dryer diameter, its length and rotating speed, and holdup weight, and is as follows:

<i>D</i> (m)	<i>L</i> (m)	Motor Power (kW)
1.5	8	8
1.8	12	14
2.2	12–14	17–20
2.8	12–14	28–35
3.2	18	75
3.5	27	120

Typical operating parameters for rotary dryers used for coal drying are given in Table 43.1.

The heat consumption in such dryers amounts to about 3700 kJ/kg H₂O. The drying time is about 15 to 40 min at a holdup fraction of 0.15 to 0.25. The gas velocity at the dryer outlet should not be more than 2 to 3 m/s for the drying of fine coals derived from washing and not more than 0.5 to 1 m/s for the postflotation concentrates to avoid excessive carryover of coal. The dimensions of the rotary dryer are determined based on the drying rate per unit volume, which is given in Table 43.1.

Generally, one cannot know the suitable drying rate per unit volume because it varies with the dryer diameter, the flow rate, the temperature of the drying medium, and the rotation speed of the shell. One has to calculate dryer volume based on the heat transfer coefficient.

Knowing the drying rate per unit volume (amount—kilograms—of moisture evaporated in time unit from 1 m³ of shell volume), one can calculate the volume of dryer shell *V_b* from the formula

$$V_b = \frac{W}{N_v} \quad (43.1)$$

From the volumetric flow rate of the drying medium *V_g*, assumed fractional holdup *β* (0.05 to 0.3), and the velocity of drying medium in the drum *u*, one can calculate the diameter of the dryer as

$$D = \frac{1.13}{\sqrt{1-\beta}} \sqrt{\frac{\dot{V}_g}{u}} \quad (43.2)$$

TABLE 43.1
Operating Parameters for Rotary Dryers Used for Coal Drying

Coal Type	Coal Moisture (%)		Heating Medium Temperature (°C)		System of Mixing Devices	Drying Rate per Unit Volume, N_v (kg/m ³ h)
	Before Dryer	After Dryer	Before Dryer	After Dryer		
Hard coal	9	0.6	900	60	Figure 43.2, type a	35–40
Fine coal mixed with postflotation concentrate	17	5	740	110	Figure 43.2, types a, b	93
Postflotation concentrate	22	5	770	105	Figure 43.2, types a, b	116
High-ash mud, grain size 0–2 mm	30	1	750	120	Figure 43.2, types b, d, e, f	120

The length of the dryer is calculated from the formula

$$L = \frac{V_b}{0.785D^2} \quad (43.3)$$

The length–diameter ratio L/D is usually in the range of 5 to 10. The speed of rotation of the shell n depends on the length L , diameter D , angle of inclination of shell α , and the drying time τ . It is determined by the empirical formula

$$n = \frac{k_1 k_2 L}{\tau D t g \alpha} \quad (43.4)$$

where k_1 = coefficient characterizing the motion of material in the drum. For cocurrent dryers used for coal drying, $k_1 = 0.2$ to 0.7 ; k_2 = coefficient taking into account the type of distributing or mixing device. For lifting devices, $k_2 = 0.5$; for the cell and sectional devices, $k_2 = 1.0$

In practice, the peripheral speed of rotation is usually 15 to 25 m/min. The dwelling time of coal in the dryer τ is very important in the drying of coking coals. Excessive drying time can worsen the agglomerating property of the coal. The drying time of coal can be determined from the formula

$$\tau = \frac{2\zeta_u \beta}{N_v} \frac{X_1 - X_2}{2 - (X_1 - X_2)} \quad (43.5)$$

43.2.2 ROTARY-TUBE DRYER

The rotary-tube dryers are used widely for coal drying in brown coal briquetting plants. They are also used for drying of hard coals. These dryers are indirect dryers heated by saturated steam at pressures of 0.15 to 0.55 MPa. This dryer (Figure 43.3) consists of a sloping drum (1), in which the perforated walls

have seamless tubes fastened to them (2) at diameters of about 102 to 108 mm. The dryer shell rolls on special rings (6) and rollers (7) and is driven by power transmitted (8) by means of a gear wheel (9). The heating steam enters the drum through the pin (3). The moist coal is fed inside the heating tubes. As a consequence of the slope and rotation, the coal is displaced gradually toward the heating tubes from the inlet to the outlet. In the heating tubes are installed screw guides, which control the displacement of coal. The variable-pitch guides installed in tubes prevent the displacement of coal in tubes very fast. The dried coal is collected in the lower part of the chamber (5). The steam condensate is drained through a pin placed in the lower end of the drum. In this type of dryer, the convective heat transfer coefficient from steam to coal is about 25 W/m²·K. Some technical data for the drum-tube dryers are as follows:

- Diameter $D = 2500, 2800, 3130, 3350, 3750$, and 4000 mm
- Drum length $L = 7$ to 8 m
- Angle of inclination of drum $\alpha \cong 8^\circ$
- Speed of rotation $n = 5$ to 9 rpm
- Drying rate per unit exposed surface $N_F = 5.4$ to 8 kg/m²·h
- Temperature of vapor, 90°C
- Coal temperature at outlet, 80°C
- Heat consumption, 2950 to 3100 kJ/kg H₂O
- Dust content in vapor in drying of brown coal, ~ 25 g/m³

In brown coal briquetting plants, these dryers are heated by exhaust steam from the briquetting machines. The steam carries away oil droplets, which are carbonized, forming deposits on the tube walls. To remove impurities from the dryer tubes, these tubes may be washed by circulation of trichloroethylene at

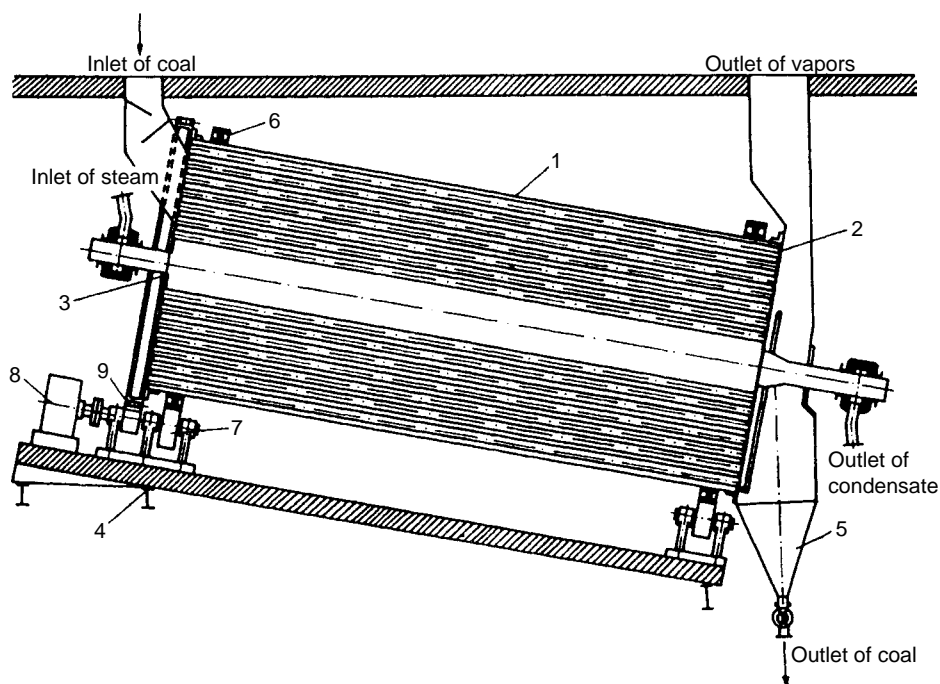


FIGURE 43.3 Drum-tube dryer.

70 to 80°C for about 3 h. The spent trichloroethylene is distilled for reuse.

43.2.3 CHAMBER DRYER EQUIPPED WITH STIRRERS

Chamber dryers equipped with stirrers are commonly used for drying of flotation concentrates (Figure 43.4). The dryer operates in a cocurrent mode. Hot combustion gases supplied to the dryer by a duct (5) meet at the outlet the cold and moist coal supplied by another duct (7). The dryer consists of a chamber (1) fitted with baffles (9) and two shafts (2) rotating in opposite directions on which paddles are mounted

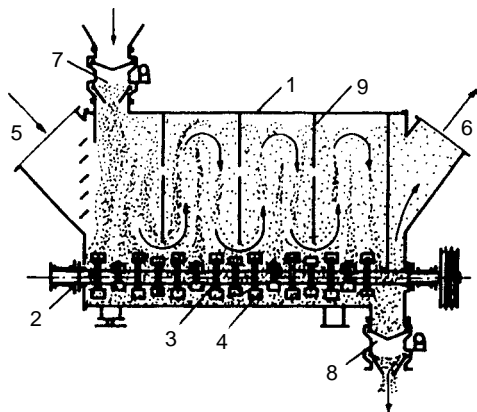


FIGURE 43.4 Chamber dryer equipped with stirrers.

(3,4). The moist coal supplied by a duct (7) is lifted and thrown by stirrers (3,4) from the inlet to the outlet (8). The stirrer paddles (3,4) are also useful for breaking sintered coal. Some technical data are as follows:

- Combustion gas temperature at inlet, 700°C
- Combustion gas temperature at outlet, 110°C
- Chamber width, 2000 to 2400 mm
- Chamber length, 2000 to 3400 mm
- Drying rate per unit volume, $N_v = 600$ to $750 \text{ kg H}_2\text{O/m}^3\cdot\text{h}$
- Coal moisture content at inlet, 18 to 22%
- Coal moisture content at outlet, 8.5%
- Heat consumption, $\sim 3150 \text{ kJ/kg H}_2\text{O}$
- Speed of rotation of shaft, 7 to 10 rpm

For pasty feeds, which tend to cake during drying, chains are fastened to the paddles to break the cakes formed during drying.

43.2.4 PNEUMATIC DRYERS

Pneumatic dryers are widely used for drying coal and flotation concentrates. The basic element of the dryer is a vertical tube of diameter 650 to 1100 mm and length 14 to 35 m through which the hot drying medium (e.g., combustion gases or air) flows from bottom to the top. The coal being dried is lifted by the stream of drying gases and transported from

bottom to top. During pneumatic transport, the coal grains are heated and dried. Thus, the drying gas velocity depends on the grain size of coal being dried; in practice this amounts to 10 to 40 m/s. In a commercial dryer, the disintegrator connected to the vertical drying tube at the bottom is installed, which serves to crush and dry moist lump coal from the feeding chute. Drying hot gas enters the disintegrator and meets the moist coal. The moist coal is highly disturbed and mixed with hot gas, and drying is strongly accelerated. Generally, about 50% of the water to be dried is removed from the coal in the disintegrator. The most intensive drying occurs in the first 2 to 3 m of the dryer. The drying process is conditioned by the heating medium velocity in relation to the grain size of coal. This velocity increases with grain size, as shown in Figure 43.5. This is very important because coarser grains reside a longer time in the heating medium. The large surface of contact of the grains with the heating medium allows rapid drying, which for finely ground coals is of the order of a few seconds. The short drying time relates only to the first drying period, i.e., to the evaporation of surface

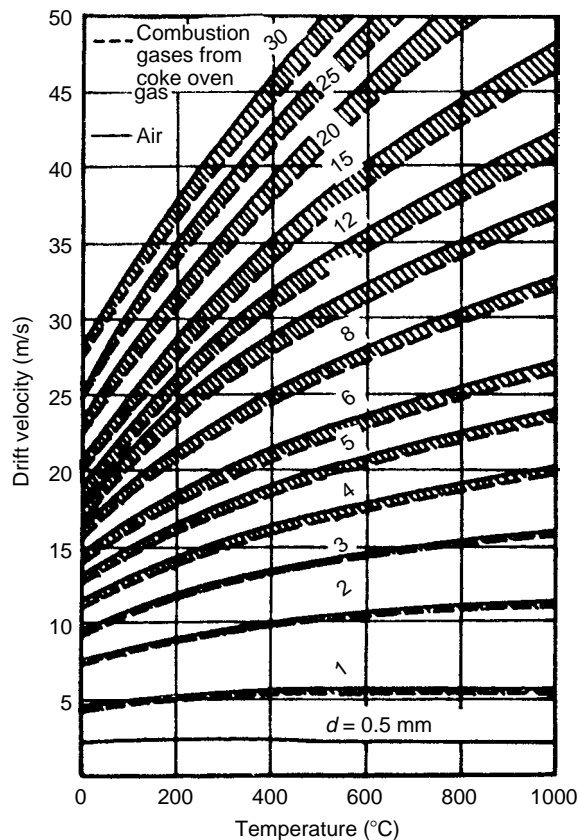


FIGURE 43.5 Drift velocity of coal grains according to Rammler and Augustin.

moisture. The heating medium parameters have a decisive effect on this period. The pressure drop characteristics of a pneumatic dryer are illustrated in Figure 43.6, given by Barth. In the vertical tube, the air flows upward. Into the airstream is supplied a grained material shaped as globules. At low air velocities the globules fall downward; at a high velocity they are entrained by the airstream. The pressure drop is increased rapidly at the moment when the globules are entrained because the material begins to collect in the tube. Beyond the critical zone characterized by a rapid drop of pressure is the operation zone of pneumatic dryers. To select the optimum flow velocity, one must avoid operation at the clogging zone. The clogging zone is idiosyncratic to each dryer. Here, a slight increase in solid loading or slight fall in pressure drop causes the clogging of the pipe. The clogging zone separates the critical transition zone from the zone of pneumatic transport. Although Figure 43.6 does not refer to coal transport, behavior is similar when transporting ground coals. In pneumatic transport, heat and mass exchange take place mostly by convection. The heat exchange between the drying medium and the coal particles suspended in this medium can be defined by the Frössling formula

$$Nu = 2 + 0.55Pr^{1/3}Re^{1/2} \quad (43.6)$$

In the first 2 to 3 m of drying tube, heat transfer to the coal particles takes place promptly, so the heat transfer coefficient in this zone is defined as

$$Nu_{\max} = 0.95 \times 10^{-4} Re^{2.15}, \quad 400 < Re < 1300$$

$$Nu_{\max} = 0.76 Re^{0.65} \quad 30 < Re < 400$$

Figure 43.7 shows the relationship between the Nusselt number and the Reynolds number for heat transfer between particles transported by the air at 0 to 100°C.

The heat supplied to the coal particle suspended in the gas stream is

$$\dot{Q} = h_c \pi d_p^2 \Theta_m \quad (43.7)$$

The mean temperature difference Θ_m is calculated as the logarithmic mean

$$\Theta_m = \frac{\Theta_1 - \Theta_2}{\ln \Theta_1 / \Theta_2}$$

where

$$\Theta_1 = T_1 - T_{m1} \quad \text{and} \quad \Theta_2 = T_2 - T_{m2}$$

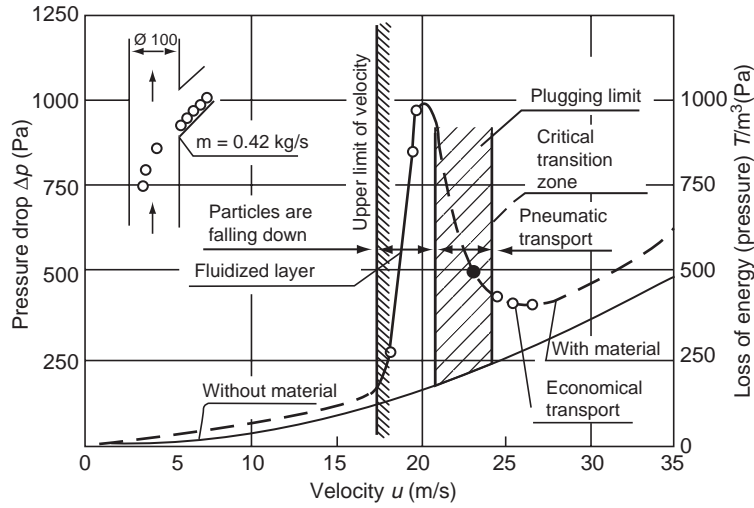


FIGURE 43.6 Pressure drop of drying medium in empty tube as well as during transport of material (type of material: spheres, 7.5 mm diameter).

The quantity of heat necessary to evaporate the surface moisture is

$$Q = \frac{\pi d_p^3}{6} \zeta_s (X_1 - X_2) \Delta H \quad (43.8)$$

After substitution of this value in Equation 43.9, the drying time is given by

$$\tau = \Delta H \frac{d_p^2 \zeta_s (X_1 - X_2)}{12 \Theta_m \lambda} \quad (43.10)$$

The time of the drying process is estimated from the formula

$$\tau = \frac{Q}{\dot{Q}} = \frac{d_p \zeta_s}{6 h_c} \frac{X_1 - X_2}{\Theta_m} \Delta H \quad (43.9)$$

At low Reynolds numbers ($Re \cong 0$) the Frössling formula gives

$$Nu = \frac{h_c d_p}{\lambda} \cong 2$$

In practice, the moist coal particle diameter is larger than the dried, and the particle diameter continues to change with the moisture decrease, so the correction for the drying time based on the coal particle diameter must be taken into account. The length of the drying duct, i.e., the effective length of the drying tube, is then

$$L' = u \tau \quad (43.11)$$

The velocity u is the mean velocity with which the coal grain flows in time τ through the lifting tube.

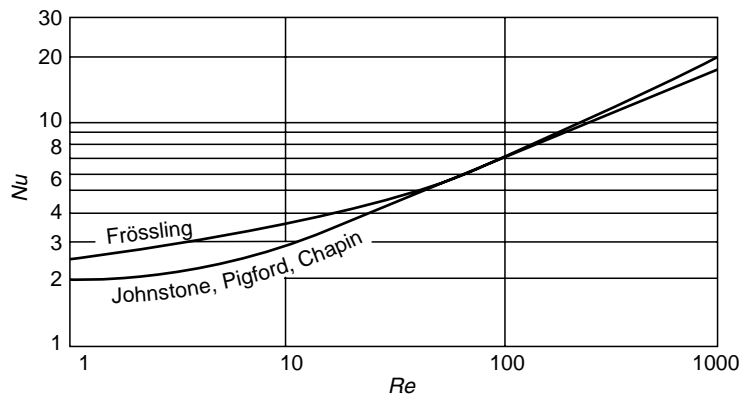


FIGURE 43.7 Heat reception by globules in air at temperatures of 0 to 100°C.

$$u = u_g - u_0 \quad (43.12)$$

The velocity of the drying medium is taken as

$$u_g = (1.1 - 1.25)u_0 \quad (43.13)$$

The sedimentation velocity can be defined on the basis of the Archimedes number.

$$Ar = \frac{g d_p^3 \rho_s - \rho_g}{\mu_g^2 \rho_g} \quad (43.14)$$

Knowing the Archimedes number, one can calculate the Reynolds number; for $Ar < 84,000$,

$$Re = \left(\frac{Ar}{13.9} \right)^{1/1.4} \quad (43.15)$$

For $Ar > 84,000$

$$Re = 1.71 \sqrt{Ar} \quad (43.16)$$

From the Reynolds number we estimate the theoretical sedimentation velocity of the coal particle as

$$u_0 = \frac{Re v_g}{d_p} \quad (43.17)$$

The tube length calculated based on formula [11] is too small since with coarse grains (0.2 to 1.0 mm) this does not take into account the starting length over which the grains reach their terminal velocity. For the correction of this starting (transient) length we suggest the formula

$$L'' = k_L u_g d_p \quad (43.18)$$

where $k_L = 10^3$ s/m. The total length of the lifting tube will be

$$L_T = L' + L'' \quad (43.19)$$

The quantity of drying medium required depends on the heat needed for the drying according to the heat balance equation

$$\begin{aligned} \dot{G} C_p (T_1 - T_2) = \dot{L}_0 (C_s + C w_1) (T_w - T_{m1}) \\ + \dot{L}_0 (w_1 - w_2) \Delta H + L_0 (C_s \\ + C w_2) (T_{m2} - T_w) \end{aligned} \quad (43.20)$$

where \dot{L}_0 = dry solid flow rate (kg/s), C_s = specific heat of solid (kcal/kg·K), W = dry basis moisture content, T_w = wet bulb temperature (°C), and T_m = solid temperature (°C).

In industrial practice, the length of the lifting tube is determined most frequently based on the drying rate per unit volume, which is as follows:

$$N_v = 400 - 600 \text{ kg H}_2\text{O/m}^3 \cdot \text{h} \quad \text{for ground coal}$$

$$N_v = 700$$

$$- 900 \text{ kg H}_2\text{O/m}^3 \cdot \text{h} \quad \text{for postflotation concentrate}$$

N_v varies generally with hot gas temperature, solid–gas ratio, and gas velocity. Therefore, suitable drying tube volume must be calculated on the basis of the equation of heat transfer and heat balance. N_v gives the approximate value of the drying tube volume. The drying rate per unit volume is defined by

$$N_v = \frac{\dot{W}}{V_p} \quad (43.21)$$

The diameter of the lifting tube is calculated from the formula

$$D = \sqrt{\frac{\dot{V}_g}{0.785 u_g}} \quad (43.22)$$

The volumetric gas rate of the drying medium can be calculated based on the heat balance Equation 43.20:

$$\dot{V}_g = \frac{\dot{G}}{\zeta_g} \quad (43.23)$$

From Equation 43.21, the volume of the lifting tube V_p can be calculated. Knowing the lifting tube volume V_p and its diameter, we can calculate the length of the lifting tube as

$$L_e = \frac{4 V_p}{\pi D^2} \quad (43.24)$$

In general, the lengths of lifting tubes do not exceed 35 m. The coal feed to the lifting tube is situated 4 to 6 m above the gas feed. The dryer capacity depends on the diameter of lifting tube.

Diameter (mm)	700	830	900	1100
Capacity (Mg/h)	20–35	30–55	40–65	50–75

The heating medium temperature at the dryer inlet is 550 to 700°C and 70 to 170°C at the dryer outlet. The initial moisture of coal is usually 10 to 15%; the final moisture is about 4 to 6%. A typical pneumatic dryer setup for coal drying is shown in Figure 43.8. Wet coal from the bin (11) is proportioned by feeder

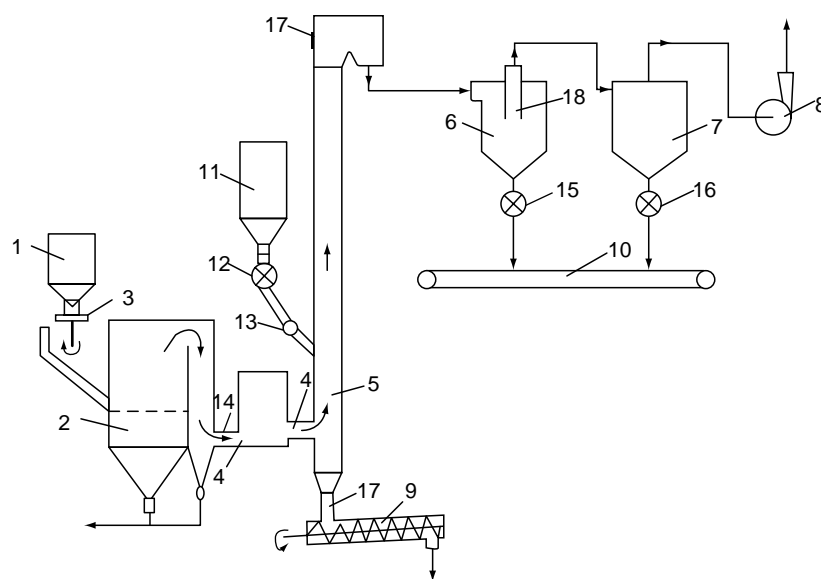


FIGURE 43.8 Schematic of pneumatic dryer for coal drying.

(12) to the lifting tube (5), where it is carried away by the combustion gas stream derived from fuel combustion in the furnace (2). Drying occurs in the tube, after which the coal is separated from combustion gases in the cyclone (6) and then in a bank of cyclones (7). Negative pressure in the system is generated by a fan (8). The dried and separated coal is directed to the conveyor (10) through the shutter closures (15, 16), which are the lower seals of the cyclones (6 and 7). The heavier coal particles, which fall down in the lifting tube (5), are directed to the dried coal conveyor (10) through the screw (9). The system is equipped with explosion flaps (17).

The Parry-type dryer for coal drying is shown in Figure 43.9. Here, wet coal is proportioned into the drying chamber (2) by means of a feeder. Around the drying chamber are installed cyclones to separate the coal from combustion gases. The combustion gases necessary for drying are generated in the chamber (1) by combustion of a solid or liquid fuel. The air for combustion is supplied by the fan (6). Another fan (5) allows recycling of a part of the combustion gases to the drying process. The capacity for this type of dryer generally exceeds 50 Mg/h at initial and final coal moisture levels of 35 and 4%, respectively.

The dimensions of a typical dryer of this type are as follows:

- Diameter = 2.7 m
- Height = 6 m
- Diameter of furnace chamber = 4.2 m
- Height of furnace chamber = 10 m
- Thermal power of chamber = 13.4 MW

The combustion process is controlled automatically; the temperature of the combustion gases outflowing to the environment is the signal for control.

43.2.5 FLUID-BED DRYERS

Fluid-bed dryers are widely applicable for coal owing to the high intensity of this drying process. Intense drying is achieved as a result of good mixing, use of a high-temperature heating (and fluidizing) medium,

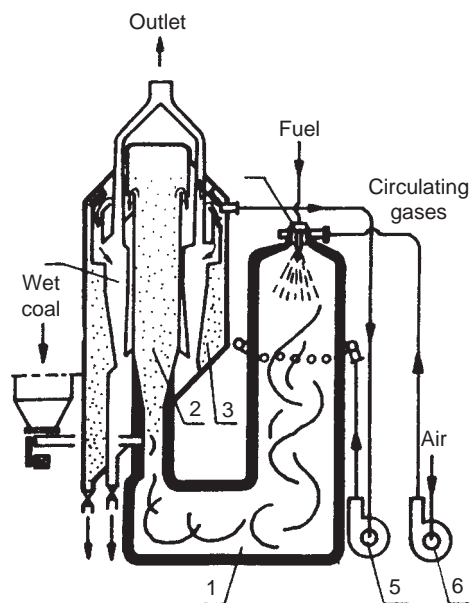


FIGURE 43.9 Schematic of Parry-type dryer for coal drying.

and ease of control. A fluidized state is achieved when gas with a proper velocity passes through the coal layer. The gas flow velocity at which the packed bed is converted into a fluidized bed is known as the minimum fluidization velocity. One feature of the fluidized bed is its high porosity, defined by

$$\psi = \frac{V'_p}{V_z} = 1 - \frac{V_m}{V_z} \quad (43.25)$$

where V'_p = volume of solid-free space, V_z = total volume of bed, and V_m = volume of solid material in the bed ($= L/\zeta_s$).

A packed bed of coal has a porosity in the range of 0.4 to 0.5. The porosity of a fluidized bed can vary over a wide range, depending on the gas flow velocity ($\psi = 0.4$ to 1). The minimum velocity of fluidization for the coal and other friable materials for which the porosity $\psi_0 = 0.4$ can be calculated from the formula

$$Re_{cr} = \frac{Ar}{1400 + 5.22\sqrt{Ar}} \quad (43.26)$$

For $\psi_0 = 0.48$

$$Re_{cr} = \frac{Ar}{710 + 4\sqrt{Ar}} \quad (43.27)$$

From the critical Reynolds number Re_{cr} , one can calculate minimum (critical) velocity of fluidization as

$$u_{cr} = \frac{Re_{cr}u_g}{d_p} \quad (43.28)$$

Equation 43.26 and Equation 43.27 are shown graphically in Figure 43.10. For beds of coal, the pressure drop can be estimated from the formula

$$\Delta p = H(1 - \psi)(\zeta_s - \zeta_g)g \quad (43.29)$$

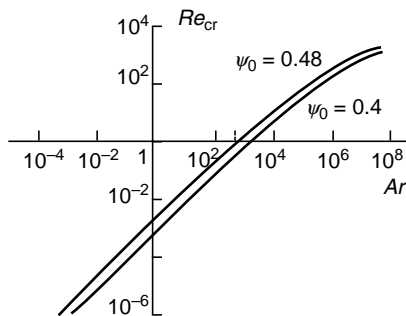


FIGURE 43.10 Relationship between critical Reynolds number and Archimedes number.

The porosity of the fluidized bed can be estimated using

$$\psi = \left(\frac{18Re + 0.36Re^2}{Ar} \right)^{0.21} \quad (43.30)$$

The height of the fluidized bed is readily given by

$$H = H_0 \frac{1 - \psi_0}{1 - \psi} \quad (43.31)$$

For coal dryers, a space of about 3.5 to 4.5 times the fluid-bed height should be allowed as the disengagement height h_1 :

$$h_1 = 3.5H - 4.5H \quad (43.32)$$

The convective heat transfer coefficient between gas and solid in the fluidized bed can be estimated from the correlation

$$Nu = 10^{-2} \frac{Re}{\psi} Pr^{1/3} \quad (43.33)$$

Drying kinetics in the fluidized bed requires in addition the fulfillment of the relationship

$$H = \frac{\dot{G}\tau}{\zeta_s(1 - \psi)S} \quad (43.34)$$

The mean residence time τ of coal in the fluid-bed dryer is determined empirically and τ can be calculated by

$$\tau = \frac{3600H_0A\zeta_s}{F}$$

where F = coal feed rate (kg/h), A = fluid-bed area (m^2), and H_0 = static coal bed height (m).

In coal drying, the temperature of drying gases at the dryer inlet is 300 to 700°C and the temperature of the fluidized bed is 70 to 80°C.

Figure 43.11 shows the relationship between gas velocity and particle diameter for various porosities of bed for coal of density $\zeta = 1200 \text{ kg/m}^3$ and for gas temperature $T = 100^\circ\text{C}$. Between curves 1 and 5 there is a zone of dryers operating as a loose bed. Industrial fluid-bed dryers operate in the range of $\psi = 0.55$ to 0.75. Zone 2 includes coal particles from 30 to 40 μm up to 6 to 7 mm. For fountain-type dryers, this zone can widen (3) up to 25 to 30 mm. The aerofountain-type dryers (4) operate at bed porosity $\psi = 0.75$ to 0.9. In dryers of this type, one can dry coal-containing particles from several micrometers to tens of millimeters. Owing to their simple design and high capacity, fluid-bed

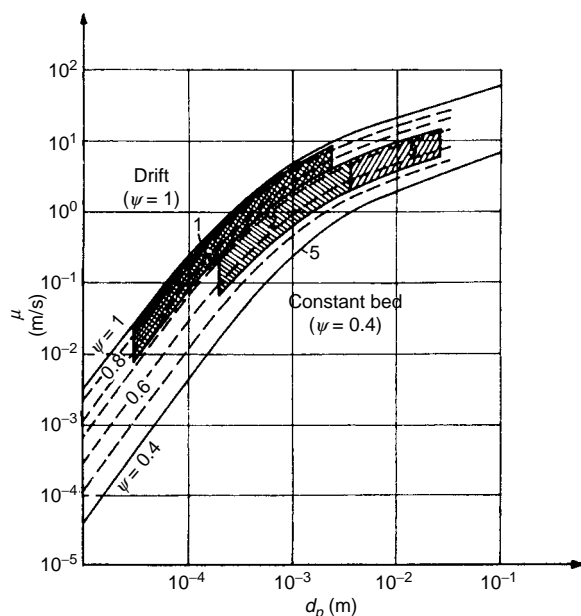


FIGURE 43.11 Relationship between gas velocity and particle diameter.

dryers are widely applicable in the industry for coal drying. One type of fluid-bed dryer commonly used for coal drying is the Fluo-Solids dryer marketed by Dorr-Oliver (United States). This dryer operates with a small coal bed (300 to 400 mm) and has a very high drying rate per unit exposed surface of 2000 kg H₂O/m²·h. Figure 43.12 schematically shows the Fluo-Solids dryer with the roller chamber designed by Dorr-Oliver for coal drying. Moist coal from the bin (1) is proportioned by feeder (2) to the drying chamber (3). Combustion gases from the combustion chamber (4) are mixed with air to obtain the desired temperature (300 to 700°C). The heat consumption is typically in the range of 3100 to 4000 kJ/kg H₂O. The hot gases pass

through a screen and the coal layer, causing fluidization of coal. The exhaust gases leaving the dryer are cleaned in a cyclone (5) and cloth filter (6). Some technological data for typical coal fluid-bed dryers for coal operating in the United States are given in Table 43.2.

43.2.6 FLUID-BED DRYER WITH FOUNTAIN BED (SPOUTED BED)

In recent years, fluid-bed dryers with a fountain (spouted) bed have become popular for drying of coal. The fountain bed is achieved in a conical-cylindrical apparatus (Figure 43.13). The hot gas stream supplied to the dryer by tube (3) carries away the coal grains supplied by screw (2) and moves them upward. The lifted coal grains fall aside and move downward along the annulus.

Figure 43.14 shows a schematic of a spouted bed dryer for coal as designed by Fiodorov and Michailov. Wet coal is supplied to the conical drying chamber. The gas velocity decreases gradually in the conical chamber owing to which coarser coal grains fall downward. The finest grains are entrained by the gas stream and then separated in the cyclone (3). The intermediate fractions are circulated in the drying chamber (1), after which they pass through a vertical slit in the chamber wall for discharge. Control of the slit opening allows control over the dryer holdup. The coarse coal particles fall into the bin (5) from which they are recycled to the dryer by a bucket elevator (6). For spouted bed dryers, it is very important to determine the gas velocity at which the spouting process begins and ends. The critical velocity for spouting depends on the bed height and the physical properties of the coal and gas, as well as on the geometric shape and dimensions of the dryer. The incipient spouting velocity can be estimated from the correlation

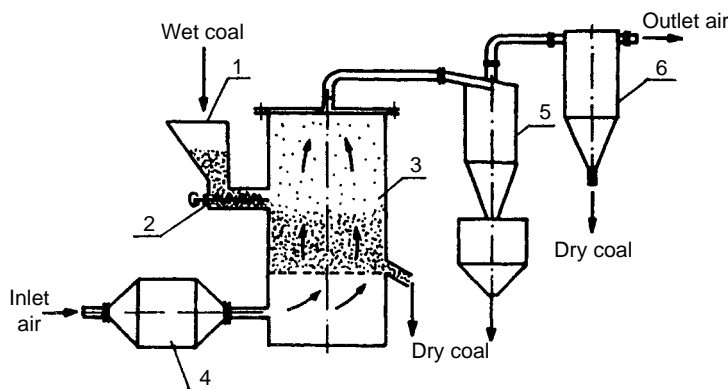


FIGURE 43.12 Schematic of Fluo-Solids-type fluidized bed dryer.

TABLE 43.2

Operating Parameters for Some Typical Fluid-Bed Dryers for Coal Used in Various Parts of the United States

Parameters	Indiana	West Virginia	Kentucky	Utah
Diameter, m	2.1	4.2	4.2	4.2
Screen area, m ²	3.46	13.9	13.9	13.9
Capacity, Mg/h	100	700	230	800
Drying rate per unit exposed surface, kg/m ² · h	2900	2500	1800	1800
Coal grain sizes, mm	0–6	0–10	0–15	0–38
Moisture at inlet, %	18–22	14.5	—	—
Moisture at outlet, %	2	4.8	—	—
Gas temperature at inlet, °C	650	410	—	—
Gas temperature at outlet, °C	80	70	—	—

$$Re_{cr} = 0.364 Re_f \left(\frac{D}{d_0} - 1 \right)^{0.82} \left(\tan \frac{\phi}{2} \right)^{0.1} \quad (43.35)$$

where Re_{cr} = critical Reynolds number at which spouting starts, D = diameter of upper bed section, d_0 = diameter of lower bed section, and ϕ = apex angle of cone; generally, $\phi = 16$ to 70° .

Then

$$u_{cr} = \frac{Re_{cr} u_g}{d_p} \quad (43.36)$$

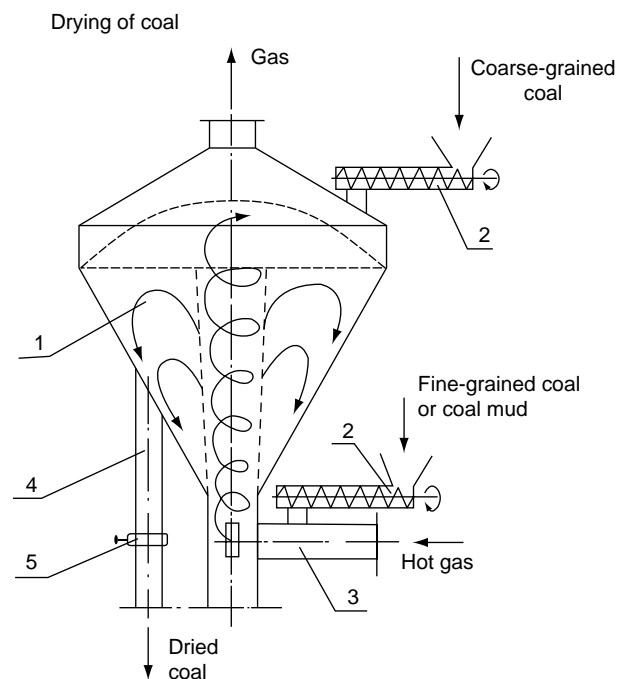


FIGURE 43.13 Schematic of dryer with spouted bed.

In polydisperse systems, the maximum particle diameter is taken as the particle diameter in Equation 43.36. For optimum geometry one may use

$$\phi = 30 - 50^\circ \quad \text{and} \quad \frac{D}{d_0} = 2 - 4$$

For fine grains ($Ar < 10^4$), the bed becomes unstable and may pulsate with a frequency at a low amplitude. It is found that the dryer height H and separator diameter D_s also affect the minimum spouting velocity. These effects are included in the correlation

$$Re_{cr} = 0.176 Re_f \left(\frac{D}{d_0} - 1 \right)^{-0.087} \left(\frac{H}{d_0} \right)^{0.6} \left(\frac{D_s}{d_0} \right)^{0.94} \left(\tan \frac{\phi}{2} \right)^{-0.323} \quad (43.37)$$

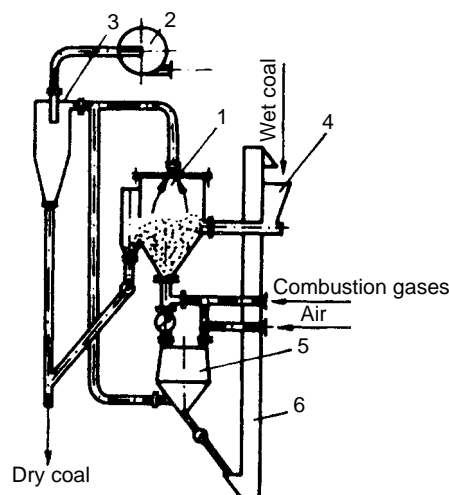


FIGURE 43.14 Schematic of fluidized system with spouted bed.

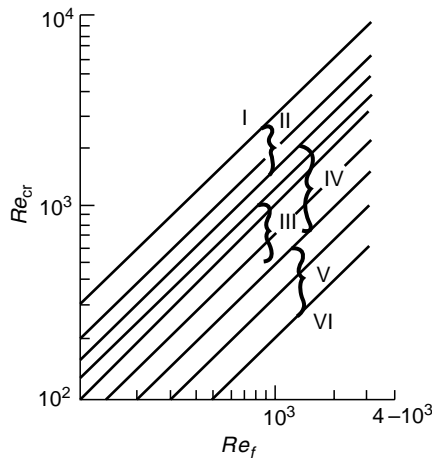


FIGURE 43.15 Relationship between critical Reynolds number at incipient spouting and Reynolds number calculated for sedimentation velocity Re_f .

Figure 43.15 presents the relationship between Re_{cr} and the Reynolds number for the settling velocity of grain Re_f as well as various operating zones of spouted beds. To estimate the pressure drop across a spouted bed, one can use the approximate formula

$$\Delta p = (0.64 - 0.75)g\zeta_u H_0 \quad (43.38)$$

It is very difficult to calculate the values of convective heat transfer coefficients for a spouted bed. Uemaki and Kugo give the following correlation for convective heat transfer from the gas to the solid particles.

$$Nu = 5 \times 10^{-4} Re_{min}^{1.46} \left(\frac{u}{u_{min}} \right)^{1.30} \quad (43.39)$$

where the Reynolds number is defined by formula

$$Re_{min} = \frac{u_{min} d_p}{u_g}$$

where u = flow velocity and u_{min} = minimum velocity for spouting.

From Equation 43.39, it appears that the convective heat transfer coefficients in the spouted bed are in the range of 3.4 to 17 W/m²·K; according to other investigators these coefficients are much higher, say, up to 51 to 142 W/m²·K. The drying rates per unit volume in such dryers are high: $N_v = 110$ to 290 kg/m³·h. For coal drying, the temperature of the drying medium (e.g., combustion gases or air) is usually not more than 200°C, unlike the case of conventional fluid-bed dryers, which operate with much higher inlet temperatures.

43.2.7 VIBRATORY DRYERS

A vibratory dryer designed by Escher-Wyss is sketched in Figure 43.16. Vibratory dryers are used for drying hard and brown coals, especially if it is necessary to combine coal transport or proportioning operations with the drying operation. Drying is accomplished by hot air or combustion gases passing through the vibrating coal layer. The coal to be dried is led into the perforated trough oscillated by an electromagnetic oscillator. The trough is inclined to the horizontal at an angle of about 1 to 5°. Owing to the slope of the trough the coal is conveyed with a velocity dependent on the slope of the trough and on the vibration amplitude and frequency. For an electromagnetic vibrator at frequency $f = 50$ and 100 Hz and amplitude $x = 0.05$ to 3 mm, the conveying velocity of coal is 0.01 to 0.3 m/s. This type of unit can also be used to cool coal. In vibratory dryers, the heat requirement is much lower as a consequence of better utilization of the drying medium.

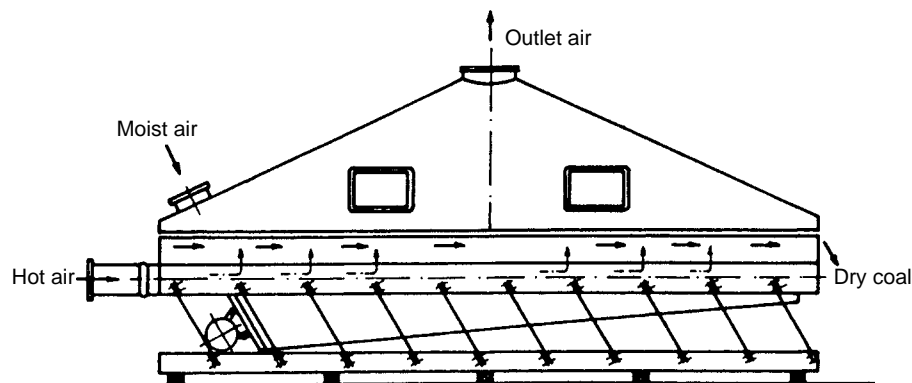


FIGURE 43.16 Schematic of Escher-Wyss-type vibratory dryer.

The capacity of vibratory dryers depends on the trough width B , coal layer height h_0 , and coal drift velocity u . It can be calculated using the formula

$$\dot{L} = Bh_0u\zeta_u \quad (43.40)$$

The coal layer in the trough is typically

$$h_0 = 20-30 \text{ mm, for ground coal}$$

$$h_0 = 40-60 \text{ mm, for coal in lump form}$$

For vibratory dryers, the length of troughs does not usually exceed 10 m. In case the drive is located in the center of the trough, the trough length can be as high as 30 m. For long troughs, difficulties are encountered with proper distribution of the heating medium. Sometimes the electromagnetic oscillator is replaced by a power drive in the form of rotating disks that push the trough that is mounted elastically. Some technoeconomic data for coal drying in vibratory dryers are given in Table 43.3.

43.2.8 MILL-TYPE DRYERS

Mill-type dryers are used for coal only when it is advisable to combine grinding and drying operations. For satisfactory grinding performance, a low moisture level must be achieved. The mill-type dryers are used generally in power plants using steam boilers fired by pulverized coal. Figure 43.17 shows a mill-type dryer, which is really a ball mill in which the coal is broken and partly ground by the balls. The dryer consists of a horizontal drum with cones fitted on both ends. From inside the drum is lined with the protective lining of basalt, cast iron, carbon, or manganese cast steel (14% Mn).

The drum is filled with balls (diameter, 30 to 80 mm). The filling ratio of the drum is typically 14 to 30%. The filling ratio of the mill has a decisive effect

TABLE 43.3
Technoeconomic Data for Vibratory Dryers for Coal

Parameter	Sizes of Grains	
	3–10 mm	10–32 mm
Capacity, Mg/h	40	80
Moisture at inlet, %	12	7
Moisture at outlet, %	1.0	1.5
Gas temperature at inlet, °C	370	275
Gas temperature at outlet, °C	60	54
Dried coal temperature, °C	66	40
Pressure drop, Pa	380	320

on the quality and capacity of coal grinding. The balls are made of manganese or carbon steel toughened to obtain high hardness and durability. The ball consumption is 80 to 100 g/Mg for soft coals, and it can reach up to 800 g/Mg for hard coals. The drums are of 2 to 3 m diameter and 3 to 5 m long. The speed of rotation of the drum is 16 to 30 rpm. For grinding, the electricity consumption varies over a wide range, from 7 kWh/Mg for short brown coals up to 30 kWh/Mg for anthracite. Coal grinding in mill-type dryers requires additional electrical energy (5 to 10 kWh/Mg), which is consumed by the fan. Coal is fed into the mill through one of the pins; the heating medium in the form of hot air or combustion gases is also supplied through the same pin. The rotary motion of the drum causes the movement of balls near the drum wall up to about three quarters of drum diameter as the falling down of balls under gravity. The falling balls break and partly grind the coal. The hot air or combustion gases supplied to the drum simultaneously heat the balls and coal as well as dry the coal.

The quantity of heating medium used in typical commercial units is 1.3 kg/kg for brown coals and 2.0 kg/kg for anthracite. The heating medium temperatures depend on the moisture of the coal. Fig-

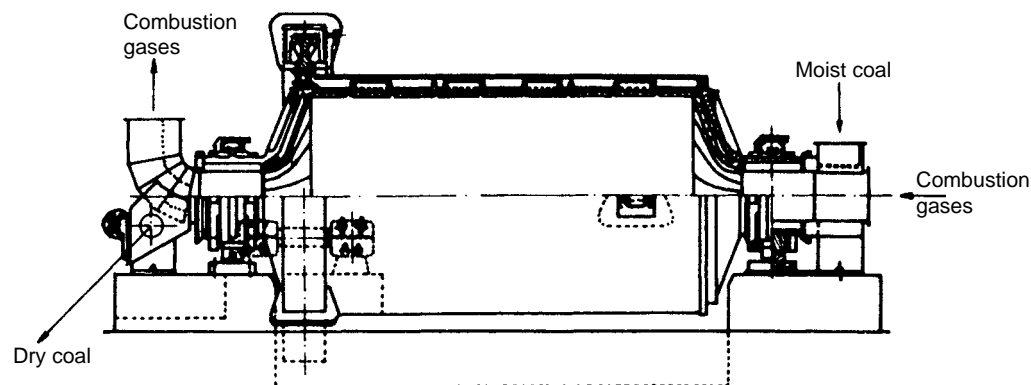


FIGURE 43.17 Schematic of mill-type dryer.

ure 43.18 shows the relationship between the combustion gas temperature, moisture content in combustion gases at the dryer inlet, and the coal moisture content. The flow velocity of the hot medium in the mill is in the range of 1 to 3 m/s.

The temperature of the coal-air mixture is 70 to 80°C for hard coal and 80 to 90°C for brown coal. The capacity of the mill-type drum dryers is 50 Mg/h or more. For drying hard and brown coal, beater mills may be used. A beater mill, including the separator, is shown in Figure 43.19. Here, drying and grinding proceed without the use of a fan. The mill consists of a separator housing and articulated rotating beaters. The beaters number 2 to 12 per row and weigh 4 to 10 kg each. The beaters are made up of manganese steel, manganese cast steel, cast iron, or carbon steel. The beater size can be different. The beaters are subjected to abrasive wear and must be replaced periodically. In grinding quick coke, the lifetime of the beaters is 200 to 400 h. The rotational speed of the beaters is 40 to 60 m/s; the number of revolutions of the shaft is 720 to 1420 rpm. Wet coal from a bin is charged by gravity into the mill through the feeder; the hot air or combustion gases are supplied by lateral stub pipes. The rotational motion of beaters causes bouncing of coal particles up to a certain height of the shaft (up to about 5 m), which causes intensive drying of coal as well as grinding of coal grains owing to the direct impact. Coarser grains are recycled for grinding; finer grains are entrained, forming a dust-air mixture supplied directly to the boiler furnace.

The power consumption for grinding is typically 15 to 20 kWh/Mg for hard coal and 4 to 14 kWh/Mg for brown coal. The metal consumption is 60 to 160 g/Mg for hard coal, 20 to 80 g/Mg for brown coal, and 2000 g/Mg for quick coke. For coals at moisture levels of 10 to 14%, air at 300 to 400°C may be used as the heating medium. When the moisture content is above 15 to 25%, combustion gases at temperature of 700 to

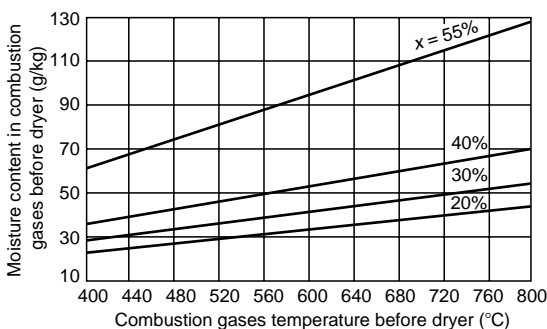


FIGURE 43.18 Combustion gas temperature; relationship between combustion gas humidity and coal moisture content.

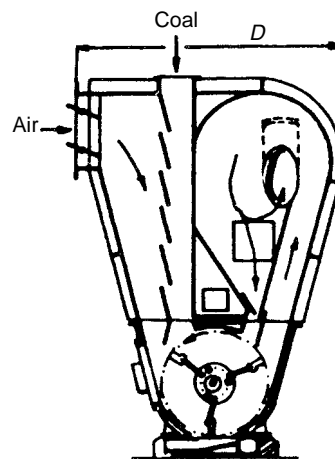


FIGURE 43.19 Mill-shaft dryer (beater mill including separators).

1000°C should be used. Coal at a high moisture content (25 to 35%) should be predried partially at the chute to the mill by means of hot combustion gases.

Figure 43.20 shows pulverized coal production with simultaneous drying of coal. Wet coal is supplied to the ball mill through the scales, feeder, and drying tube. Hot air together with the coarse particle recycling stream from the separator is directed to the mill and the combustion gases are supplied to the drying tube.

43.2.9 SHAFT DRYERS

A shaft dryer designed by Konrer-Ledant for drying of coal fed to coke ovens is shown schematically in Figure 43.21. This dryer consists of two vertical shafts of rectangular section. These rotating shafts, fitted with paddles, are intended to fluff the coal and to transport it downward. The dryer has two zones, a drying zone and a cooling zone. Combustion gases generated in the combustion chamber flow to the two vertical shafts, where they contact with the wet coal moving down the shaft. Cooling is achieved by means of ambient air supplied to the lower part of the shaft. The combustion gas is at 650°C, and after mixing with the air leaving the cooling zone its temperature drops to 250 to 300°C. At the dryer outlet the exhaust gas temperature is 60 to 70°C. In this type of dryer, the pressure drop is about 700 Pa.

The combustion gases leaving the dryer shaft are cleaned in a battery of cyclones and in a wet scrubber. The dried coal withdrawn from the lower part of the shaft by screw conveyor or a Redler-type conveyor, the trough of which is hermetically sealed,

the slit size of the stub pipes K_2 . The basic advantage of this dryer is its compact construction and simple design, as well as the possibility of full automation. It is possible to select the flow velocity to minimize the carryover of particles from the bed; in many cases this allows one to do away with any dedusting equipment. Low-temperature (150 to 200°C) waste combustion gases from steam boilers can be used as the drying medium. Low-temperature operation greatly reduces the danger of ignition. Also, the drying rate per unit volume is very high, up to about 0.2 kg/m³·s at an inlet air temperature of 200°C. For comparison, for rotary and pneumatic dryers this amounts to about 0.014 and 0.14 kg/m²·s, respectively. Rotary, pneumatic, and fluid-bed dryers are operated at much higher temperatures (400 to 800°C). The low velocity (about 1 mm/s) of movement of the coal bed considerably decreases the wear of the dryer walls from abrasion. The heat consumption is 2800 to 3500 kJ/kg H₂O. The drying time is the key design parameter for the calculation of the moving bed dryer. The drying time is equal to the residence time in the dryer. Hence the drying time has an effect on the velocity of the bed in the dryer. The drying time at constant parameters of the heating medium can be calculated from

$$\tau = \frac{1}{\mathcal{E}} [(X_1 - X_{cr}) + (X_{cr} - X^*)] \ln \frac{X_{cr} - X^*}{X_2 - X^*} \quad (43.41)$$

The drying coefficient \mathcal{E} characterizes the rate of drying under constant drying conditions and can be estimated from the correlation

$$\Pi = 0.469 Re^{0.896} K^{-0.633} \left(\frac{H}{d_e} \right)^{-1} \left(\frac{s_u}{s_g} \right)^{-1} \quad (43.42)$$

The equivalent particle diameter is calculated from the formula

$$d_e = \frac{1}{\sum \frac{x_i}{d_i}} \quad (43.43)$$

where x_i = mass fraction of diameter d_i kg/kg, u = superficial gas velocity at inlet temperature, and $\Theta = T_s - T_D$ = temperature difference between gas at inlet T_s and wet bulb temperature T_D .

The bed pressure drop Δp through the layer of coal can be calculated from the well-known Ergun's equation:

$$\Delta p/H = 150 \frac{(1 - \psi_0)^2}{\psi_0^3} \frac{\mu_g u}{d_e^2} + 1.75 \frac{(1 - \psi_0)}{\psi_0^3} \frac{s_g u^2}{d_e} \quad (43.44)$$

43.2.11 SUPERHEATED STEAM DRYING OF COAL

The advantages of using superheated steam as the drying medium are well known and noted in the chapter on this subject in this handbook. Among those of major interest in coal drying are no fire or explosion hazard, ability to dry at elevated temperatures, faster drying rates leading to small equipment size, and others (Beeby and Potter, 1984; Weiss et al., 1991; Wolf et al., 1988).

Steam-fluidized beds with steam have been in successful operation in South Africa for nearly a decade (Faber et al., 1986). Faber et al. (1986) have shown that fluid-bed steam drying of activated carbon pellets saves energy (15%) as well as capital costs (14).

Figure 43.23 shows conceptually the steam-fluidized bed drying (Dampf-Wirbelschicht-Trocknung [DWT]) described by Weiss et al. (1991). Typical operating conditions for drying of brown coal are given below.

Fluid-bed pressure drop and temperature	1–10 kPa, 110–120°C
Fluidizing steam	15–25 kPa, superheated
Heating steam	400–500 kPa absolute, saturated
Coal grain size (feed)	up to 6 mm
Coal grain size (output)	up to 4 mm
Dry coal moisture	10–20%, dry basis

Note that RBC stands for raw brown coal. Dry brown coal (DBC) is withdrawn through the inflow plate. Exhaust steam is cleared in an electrostatic precipitator and is partially recycled by a compressor as fluidizing steam. Excess steam is condensed and the latent heat is recovered at 100°C. If this heat is utilized elsewhere, the net energy consumption for coal drying is very low. The internal heat exchanger uses 4 to 5 bar saturated steam and is taken out as clean condensate. The condensation temperature must be 30 to 50°C above the bed temperature for efficient heat exchange between the bed and the exchanger tubes. The bed is operated at a slight overpressure to eliminate air in leakage.

Figure 43.24 shows a schematic of a steam-fluidized bed dryer with combined generation of power and heat. In a conventional coal-fired power plant, up to two thirds of fuel energy is lost since the latent heat of turbine exhaust steam is dissipated unused to the cooling water because of its low temperature level. In DWT process, the latent heat can be used to dry the input coal. Figure 43.24 shows a coal-fired power plant schematic with a circulating fluid-bed

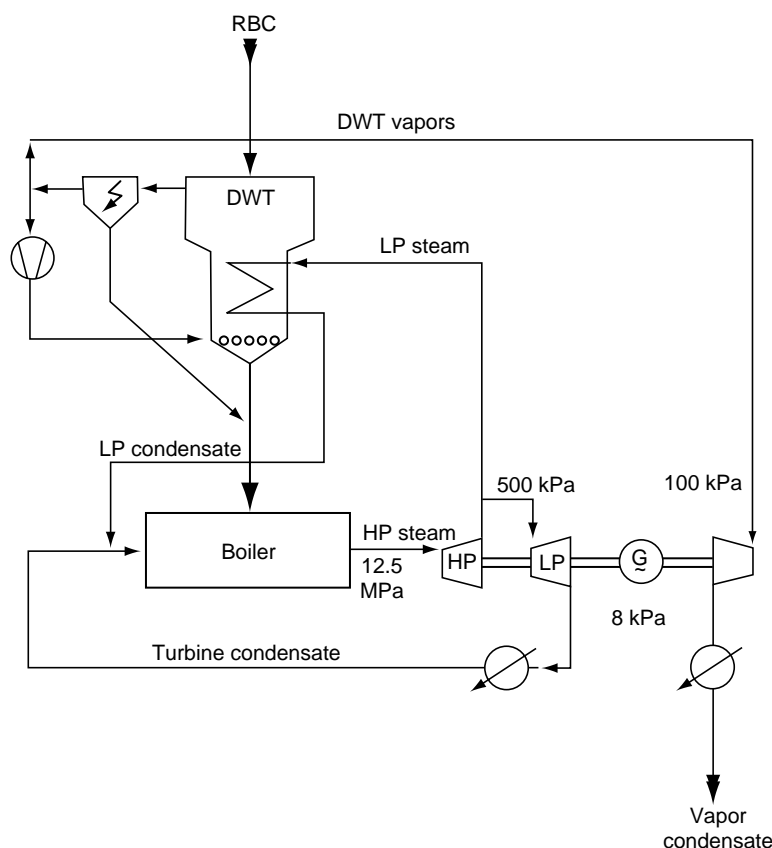


FIGURE 43.23 Steam-fluidized bed dryer (DWT, Dampf-Wirbelschicht-Trocknung) with internal heat exchangers.

boiler, turbogenerator set, and condenser for turbine exhaust steam. The top half is the DWT scheme. Use of DWT is claimed to improve the energy efficiency from 34 to 37%. If the DWT latent heat is also utilized the efficiency is expected to rise to 39%. The reader is referred to Faber et al. (1986) for details. Large-scale pilot plants have successfully demonstrated this technology. The pilot-plant capacities range from 1 to 24 t/h evaporation capacity. The latter can generate 20 t/h of dry coal (12% moisture) from 44 t/h of raw coal (60% moisture).

43.3 RECENT DEVELOPMENTS AND STATUS OF COAL TECHNOLOGIES

Coal is a raw material for many chemical syntheses as well as fuel. Coal is dried to increase its calorific value and simplify loading, unloading, transport, and to improve boiler combustion efficiency. It is also dried for processes like briquetting, coking, gasification, carbonization, and liquid fuel synthesis. Coke oven efficiency can increase 30 to 50% in preheating and 10 to 15% in drying.

Only high-moisture coals need to be dried prior to usage. The type of dryer and extent of drying

required depends on many factors. Some of the principal ones are the utilization of the mined coal, whether it needs to be ground for firing into combustion chambers and whether the application is near “mine-mouth” or far away from the mine. If the coal contains substantial amounts of water (could be 50 to 70% wet basis depending on the coal), it may be necessary to dehydrate it to reduce the transportation cost. The calorific value also increases this way while the combustion efficiency is also enhanced. Recent work at Monash University in Australia has demonstrated in laboratory and pilot tests that reduction of moisture from Australian brown coal or lignite can reduce greenhouse gas emissions from a power plant by 30%.

Direct dryers (e.g., rotary, pneumatic, fluid-bed, vibrating fluid-bed, and shaft dryers) can be used with hot air or combustion gases as drying media at 700 to 900°C before dryer and 60 to 120°C after dryer. It is important not to have high oxygen content in the drying gas to avoid explosion and fire hazard. This is a key issue in coal drying. Low-rank coals can be highly reactive and hence they are more susceptible to fire and explosion hazard due to spontaneous combustion. Hence indirect dryers have some advantages

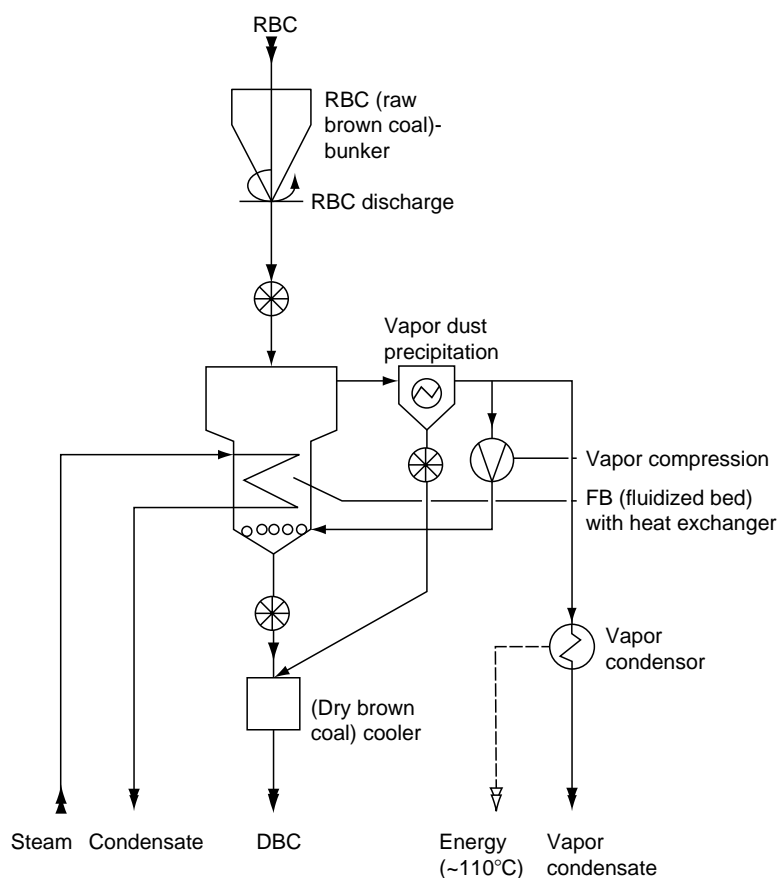


FIGURE 43.24 Steam-fluidized bed dryer with combined generation of power and heat.

in this regard. Rotary dryers with indirect heating are used for hard coals. These have higher energy efficiency, about 3100 kJ/kg water evaporated. For air fluidized bed dryers the corresponding figure is 3100 to 4000 kJ/kg water evaporated. A commercial vibratory dryer for hard and brown coals (manufactured by Escher-Wyss of Switzerland) uses a vibration frequency of 50 to 100 Hz and amplitude of 0.5 to 3 mm giving a conveying velocity in the range of 0.01 to 0.3 m/s with an angle of inclination of 5° to the horizontal. Low gas velocities are needed since vibration suspends most of the pseudo-fluidized beds. The efficiency is better than a conventional fluid-bed employing high gas velocities. Attrition is reduced and gas cleaning requirements minimized in a vibrated bed dryer (Mujumdar, 1989; Erdez and Mujumdar, 1991).

In pilot trials Potter (Potter, 1979; Potter and Keogh, 1979, 1981; Potter and Beeby, 1982, 1986; Potter et al., 1988) have shown that extremely favorable heat transfer rates as well as drying efficiencies are obtained when drying brown coal in a steam-fluidized bed with internal heat exchanger tubes im-

mersed within it. Typical processing conditions are reported as:

High tube temperature	140–170°C
Bed temperature	110–127°C
Minimum fluidizing velocity	5.7 cm/s (approx)
Steam temperature	130–155°C
Steam velocity (m/s)	0.20–0.30
Coal feed rate (kg/h) w.b.	40–70
Product (kg/h) w.b.	16–28
Drying rate (kg/h)	24–32

Using steam exhausted from one dryer stage as carrier steam for another stage, multiple-effect operation (similar to that common to evaporators) can be achieved yielding a steam economy of 1.9 for a triple-effect dryer.

Potter et al. (1988) used a continuous fluid-bed dryer for drying Victoria brown coal. The fluidized bed (FB) dryer was 0.3 × 0.3 × 3 m (high) with four bubble caps to distribute steam. The disengaging region was 2.5 m. Both horizontal and vertical tube bundles were tried.

Faber et al. (1986) have compared drying rates in air and steam-fluidized beds of pulverized coal. They confirmed existence of the inversion temperature above which steam drying is faster than air drying. Above about 180°C the steam drying rate in the constant-rate period in fluid-bed drying exceeds that in (dry) air drying. For a 2000 kg/h dryer for alumina they found the capital cost to be 20% lower for steam dryer while the total energy cost was 15% lower. No credit was given to the steam produced in the steam dryer.

Faber et al. (1986) also report on a successful industrial installation using steam dryer for activated carbon pellets (2000 kg/h, dry basis) from an initial moisture content of 50 to 2% (dry basis). The pellets are dried to 8% on dry basis before they are fed to an evacuated chamber in which the final moisture content of 2% is achieved. The steam enters the dryer at 300°C and leaves at 150°C. The steam discharged is used to preheat the feed. The authors report smooth operation of the dryer since 1985. The installed cost of the steam drying system was 40% lower than that for a conventional air dryer. The air dryer can operate at a maximum temperature of 125°C to avoid combustion in the dryer. The energy costs (1986 data) were estimated to be about \$3.60 per ton of dry product in South Africa.

A Russian book on drying and thermo-aero classification of coal presents interesting results on steam drying of 1- to 13-mm coal particles and volatiles evolution during drying. Mujumdar (1990) notes that in steam drying the drying time (actually residence time in dryer) does not affect the volatiles liberation unlike air drying. Further, he found that, under conditions of his experiment, the constant-rate drying period is 6 to 7 times longer in steam drying and the heat transfer rate is 1.7 to 2 times than that in air drying. He also reports on favorable industrial experience in steam-fluidized bed drying in a 35 × 9 × 12 m, 46 t/h of brown coal with an evaporative capacity of 25 t/h. No details are given about the steam reuse.

From energetic and safety viewpoints, superheated steam drying appears to be the most attractive technology for drying of coal. This could be done using a flash dryer, a fluid-bed dryer, a vibrating bed dryer or a rotary dryer, if traditional dryer equipment is to be used. A part of the energy can be supplied indirectly using immersed heat exchangers. As Mujumdar (1990) has noted there may be potential to use variants of multistage impinging stream dryer using superheated steam for coal drying applications, but no work appears to be done yet on this design. Furthermore, use of a two-dimensional design is expected to allow modular design and reduced scale-

up problems. If used as a fuel after briquetting, there may be opportunities to blend pulverized wet coal with wood residues and various organic sludges either before or after drying since these biomasses are also excellent candidates for superheated steam drying.

43.4 DRYING OF LOW-RANK COALS

Low-rank coals (LRCs)—e.g., brown, lignite, and subbituminous coals—represent nearly one half of the estimated coal resources in the world and are the only source of low-cost energy in many developing nations. LRCs are typically present in thicker seams with less overburden than bituminous coals, thus making them recoverable by low-cost strip mining.

From a user angle, LRCs have a lower fuel ratio (i.e., fixed carbon to volatile matter) and are typically more reactive than bituminous coals; many also have extremely low sulfur contents (a few tenths of 1%). Low mining costs, high reactivity, and extremely low sulfur content would make these coals premium fuels if not for their high moisture levels, which range from around 25% to more than 60%. Among coal importers, high moisture creates a mistaken perception of inferior quality and hence many positive features of LRCs are neglected. LRCs can be combusted either as a blending component with high-rank coal in existing boilers, or in new boilers designed for LRCs. For example, more than one third of U.S. electrical power currently comes from power stations that are fired by LRCs. Until recently the high moisture levels of LRCs have also excluded them from the rapidly growing coal-water fuel (CWF) market. Of all the coal-based alternative fuels, CWFs appear the most promising (Willson et al., 1992).

It is now accepted that no single process can be suitable as a universal drying technology for all LRCs. The needs of the end user dictate the type of process. If the end user requires dried lump coal for stoker applications, a process that uses or generates fines would not be a reasonable option. An end user with advanced combustion applications will require finely ground coal. From a producer's angle, a preferred process could make use of both technologies: one to produce sized dry coal and the other to make coal-water fuel from the fines for a different market. However, three stability issues must be solved before bulk-dried LRCs can be used: (a) moisture reabsorption; (b) dust generation; and (c) spontaneous combustion. Since the strength of LRCs is significantly reduced when their gel-like structure is destroyed by drying, the dried product breaks down rapidly, generating large amounts of dust, and becoming more liable to spontaneous combustion.

Direct drying of brown coal by recirculating boiler flue gas or its equivalent, has become standard practice in Victoria, Australia. It should be noted that significant drying occurs during normal coal grinding and other operations associated with preparation of the feed to a boiler.

Recently, a steam-fluidized bed drying process being implemented at a plant operated by the State Electricity Commission of Victoria, Australia (SECV) uses a heat exchanger supplied by an external high-pressure steam source to dry finely ground brown coal. A tube network immersed in the bed supplied heat. The high heat transfer rates of the bed, together with condensing steam, yield a compact heat exchanger. Water evaporated from the coal is used to fluidize the system. Test results show that the water content of Victoria's brown coal can be reduced from 60 to 15% and carbon dioxide (CO₂) emissions can be decreased to 17%.

In Australia, indirect drying has been used for many decades to prepare brown coal for briquetting. In this process, low-pressure steam is condensed on the outside of tubes conveying brown coal to provide process heat; the more water removed in liquid form, the higher the efficiency. After reducing the moisture to nominally 15%, dried brown coal is compacted into strong briquettes using stamp presses.

43.5 HOT OIL DRYING

Hot oil is an alternative drying medium used by a number of process developers as early as 1926. In the two-stage Carbontech process, raw coal is first dried in hot oil. Most of the oil is recovered in the second-stage flue gas stripper. Some of the oil is absorbed, which allegedly helps to stabilize the product and increase its heating value. The process costs for this system will depend on the amount of oil that can be economically recovered.

The Exxon donor solvent, direct liquefaction process also used hot oil drying. In this process, LRC is dried by a high-pressure hydrogenation reactor contact with hot recycle hydrogen donor solvent prior to entering. Unfortunately, no data were developed for solvent recovery after drying, because the dried coal and vehicle solvent were reacted immediately with hydrogen in the liquefaction reactor.

Other methods have been tested to improve the stability of the dried LRC, including spraying with residual tars or oils and briquetting or palletizing dried pulverized coal. All of these additional processing steps increase the cost of the final product and must be evaluated on a site and coal-specific basis.

43.6 HOT WATER DRYING

Hot water drying (HWD) developed at EERC, is a drying process that produces a safe, quasiliquid fuel. This technology features high temperature, high pressure nonevaporative drying that removes much of the inherent moisture and allows the production of CWFs with solid loading in the range of those of commercial bituminous CWFs.

In HWD process, ground LRC is treated at coal-specific temperatures, beginning at as low as 240°C, and the corresponding saturated steam pressure for less than 10 min. Moisture is removed from the coal by expansion and expulsion from the micropores by CO₂, which is liberated during decarboxylation. Devolatilized tars/oils, being hydrophobic, remain on the coal surface in the pressurized aqueous environment. It is hypothesized that this produces a uniform coating that seals the micropores and limits moisture reabsorption, which is a major advantage of the process. Because the coating retains most of the LRC's volatile matter, high energy recovery and excellent combustion performance can be obtained. The developers claim that alkali cations, a major source of boiler fouling, associated with the carboxyl groups, are released in the aqueous phase in their process and are removed during the final mechanical dewatering step.

The technical feasibility of HWD and low-rank coal-water fuel (LRCWF) production has been demonstrated in a 7.5-tpd pilot plant at EERC with LRCs from around the world. As a general rule, the energy densities of LRCWFs produced after HWD are around 30% for subbituminous coals, 50% for lignite, and >100% for brown coals, vs. those prepared from raw coal. Costs of dewatering will vary with coal grade and location.

Currently the successful commercial LRC drying processes appear to be those in which the dried LRC is utilized immediately and not stored. When stored, the products from most drying systems can have stability problems, which result in excessive fines and spontaneous heating.

43.7 COMBINED GRINDING AND DRYING

Often coal is ground prior to utilization. Heat produced during grinding can reduce the moisture content significantly while reducing the particle size. In some other applications, e.g., drying of sludge from a deinking plant, initial moisture content of 54% can be reduced to as low as 12% (wet basis) during grinding. This also destroys any pathogens that may be present. One of the commercial grinder dryers of interest to coal drying is the KDS Micronex grinder/dryer.

A typical KDS grinder chamber has a diameter of 1.3 m and encloses a set of 8 spinning chains and a stationary torus above it. The chains are spun horizontally at high rotational speeds; the chain tip speed can reach about 200 m/s causing high frictional heating due to aerodynamic drag. The bottom surface of the torus that is flat provides a surface for the particles to collide on and shatter. This results in grinding action accompanied by drying. The mechanism of drying is partly thermal and partly mechanical dewatering. This saves energy for drying. Air temperatures in the chamber can reach between 70 and 90°C. For a typical application for drying coal at the feed rate of about 2200 kg/h with initial moisture content of 15% on wet basis the manufacturer reports final moisture content of 4% on wet basis, and average particle diameter of 75 μm and energy consumption of 2020 kJ/kg, which is less than the latent heat of vaporization for water. Note that the Micronex unit claims to consume about 70% less energy than a traditional rotary dryer since a part of water removal takes place in liquid phase. In fact for drying of deinking sludge, manure, the manufacturer claims energy consumption of 975 to 1700 kJ/kg. The electrical power consumption for the coal dryer/grinder is rated at 140 kW. There appears to be no commercial installation for coal drying using this technology at this time.

43.8 FLEISSNER PROCESS

This is a very old process for drying low-rank coals, first developed in Austria in mid-1920s. This process consists of treating the low-rank coal in a saturated steam environment that avoids vaporization of the moisture in the coal. The coal particle “sweats” effectively and no evaporation occurs. As noted earlier decarboxylation occurs during this process causing some shrinkage of the dried product and a reduction of its equilibrium moisture content. The U.S. Bureau of Mines evaluated this process for dewatering of lignites and subbituminous coals in the 1940s and 1950s. The more recent nonevaporative drying processes for coal essentially appear to be based on the principle of the Fleissner process.

43.9 A NONTHERMAL BIOMASS DRYER

Recently GulfTex Environmental Services of Texas, United States have developed a new nonthermal dryer called Pulverizing Air Dryer (PAD). This dryer does not use heat but can reduce moisture content from 80 to as low as 10% depending on the biomass. The PAD technology uses high-velocity air streams to accelerate and then collide against the particle streams subjected

to drying. Interestingly they claim that the particles can range from large rocks to sludge material. The PAD process affects surface, loose matrix-bound as well as intracellular moisture. The drying chamber is designed as a cyclone in which the centrifugal drying chamber separates the material by density, with water coming out from the top and the dry material out from the bottom. Capacities of up to 20 t/h are possible with this novel dryer. According to the manufacturer the capital and energy costs of this PAD dryer for coal are low and can result in a saving of up to \$150 per ton of the dry product compared with a thermal dryer. Note that this figure is probably valid for a specific coal at 2004 prices. Unfortunately, no technical details are available on this apparently interesting new technology.

43.10 MECHANICAL THERMAL EXPRESSION PROCESS

A process that uses mild heating and mechanical squeezing, and appropriately termed the Mechanical Thermal Expression (MTE) process has generated renewed interest in an essentially nonthermal dehydration process for high-moisture brown coals and lignites. Developed at Monash University, this process uses mechanical expression (1 to 10 MPa) at elevated temperatures. The high pressure does not permit phase change of water and thus makes it more energy-efficient. When depressurized the wet coal also undergoes some flash evaporation and removes about 7% of the moisture. Changes in physical structure and chemical changes are hypothesized to be responsible for dewatering of the coal despite immersion in water. The precise mechanics are still unknown although there are effects of shrinkage, removal of trapped air, and changes in physical and chemical structure making the coal particles possibly more hydrophobic that make the dehydration effective. Capital costs involved in this process are still not clear but the process has definite potential. Tests on a 10 t/h pilot plant are under way in Australia at the present time (Clayton et al., 2006). Interestingly, this process has been successfully tested on laboratory scale for dewatering of biomass, e.g., bagasse, sludges, and biosolids

43.11 CONCLUSION

With rising energy costs and dwindling supplies of oil, there is little doubt that within the next decade coal technologies will play very significant role in meeting the world’s energy needs. Coal gasification and liquefaction will become increasingly important all around

the world to reduce dependence on oil and gas. Many of these processes need conditioning of the coal used that includes reduction of its moisture content to desirable levels for the processes involved. The world reserves of low sulfur but high moisture content are rather extensive and must be utilized cost-effectively by reliable energy-efficient and safe drying technologies. Different technologies will be needed depending on whether the coal is used near the mines or after shipping over great distances. Cost of transportation can be reduced by lowering the moisture content but there is danger of autoignition in storage due to most such coals being highly reactive. Thus, drying, storage, and handling of dried coal pose many technological challenges.

Over the last three decades over 300 patents were issued by the U.S. Patent Office with the word dryer or drying of coal in their titles. Of course, only a few of these patented technologies are truly viable. Many of the ideas suggested are not practical, e.g., use of acoustic radiation, ultrasonics, and super critical extraction to reduce sulfur content. Some suggest rather complex designs of rotary dryers that are unlikely to be cost-effective. For example, U.S. Patent 4,014,106 (1997) describes a rotary dryer for coal comprising of an insulated stationary cylinder within which an open-ended metallic cylindrical tube is rotated coaxially in an almost horizontal position. Wet particles are fed at the open end and hot air is circulated between the stationary cylinder and the tube. Metal balls are mixed with the wet material in the rotating tube. Perforated fins and baffles extend from the external wall of the tube to enhance conductive heat transfer to the tube and the metal balls. The dried coal is discharged at the opposite end through a screen section of the tube. This patent also suggests additives to the wet mix to coat the dried particles for reducing dusting or to prepare the mix for palletizing. There is no reported commercial application for this process. Although technically sound, it appears to be rather complex for industrial application.

On the other hand, superheated steam seems to provide all the required advantages but few vendors have developed these technologies for coal and for the large-scale operations necessary. The drying conditions will need to be optimized for specific grades of coal and also the utilization of the product. As coal needs to be ground for many applications a combined grinder-dryer is definitely an attractive option and much effort is being directed in this area as well. Drying in this case a desirable by-product of the grinding operation but supplementary drying may be required for high moisture coals since here electricity is used for the grinding operation.

NOMENCLATURE

a	exposed surface per unit volume dryer, m^{-1}
C_p	heat capacity at constant pressure, $\text{J} \cdot \text{kg}^{-1} \text{K}^{-1}$
d_e	equivalent diameter of particle, m
d_p	particle diameter, m
D	diameter, m
g	acceleration of gravity, $\text{m} \cdot \text{s}^{-2}$
G	gas flow rate, $\text{kg} \cdot \text{s}^{-1}$
h_1	space height upon bed, m
h_c	convective heat transfer coefficient, $\text{W} \cdot \text{m}^{-2} \text{K}^{-1}$
H	height of bed, m
$H_0(h_0)$	initial height of bed, m
J	humid gas enthalpy, $\text{J} \cdot \text{kg}^{-1}$
k	coefficient
\dot{L}	solids flow rate, $\text{kg} \cdot \text{s}^{-1}$
L	mass of solids, kg
L	length, m
n	speed of rotation, s^{-1} ; min^{-1}
N_F	drying rate per unit exposed surface, $\text{kg} \cdot \text{m}^{-2} \text{s}^{-1}$; $\text{kg} \cdot \text{m}^{-2}$
N_v	drying rate per unit volume, $\text{kg} \cdot \text{m}^{-3} \text{s}^{-1}$; $\text{kg} \cdot \text{m}^{-3} \text{h}^{-1}$
Q	heat quantity, J
S	cross-sectional area, m^{-2}
S	surface, m^{-2}
T	temperature, K
T_m	temperature of solids, K
u	velocity, $\text{m} \cdot \text{s}^{-1}$
u_g	gas velocity, $\text{m} \cdot \text{s}^{-1}$
u_{cr}	critical velocity, $\text{m} \cdot \text{s}^{-1}$
u_0	sedimentation velocity, $\text{m} \cdot \text{s}^{-1}$
V	volume, m^3
\dot{V}_g	cubic gas rate, $\text{m}^3 \cdot \text{s}^{-1}$
W	moisture content, kg
X	moisture content
X^*	equilibrium moisture content
X_{cr}	critical moisture content

Greek Symbols

α	angle of inclination, rad, degree
β	loading factor, 1
ΔH	latent heat of vaporization, $\text{J} \cdot \text{kg}^{-1}$
Δp	pressure drop, Pa
θ	temperature difference, K
θ_m	mean temperature difference, K
λ	thermal conductivity, $\text{W} \cdot \text{m}^{-1} \text{K}^{-1}$
μ_g	dynamic viscosity of gas, $\text{Pa} \cdot \text{s}$
ν_g	kinematic viscosity of gas, $\text{m}^2 \cdot \text{s}^{-1}$
ζ_g	gas density, $\text{kg} \cdot \text{m}^{-3}$
ζ_s	density of solids, $\text{kg} \cdot \text{m}^{-3}$

ζ_u	bulk density, $\text{kg} \cdot \text{m}^{-3}$
τ	time, s
ϕ	apex angle, rad, degree
ξ	drying coefficient
ψ	porosity of bed
ψ_0	porosity of solid (filtration) bed

Other

- 1 = inlet
2 = outlet

Commonly Used Dimensionless Group

$$Ar = \frac{gd_p^3}{\nu_g^2} \cdot \frac{s_s - s_g}{s_l} \quad \text{Archimedes number}$$

$$Nu = \frac{h_c d_p}{\lambda} \quad \text{Nusselt number}$$

$$Pr = \frac{C_p \mu_g}{\lambda} \quad \text{Prandtl number}$$

$$Re = \frac{u d_p}{\nu_g} \quad \text{Reynolds number}$$

$$\Pi = \frac{\xi d_e}{\nu_g^a} \quad \text{Drying number}$$

$$K = \frac{\Delta H}{C_p \Theta} \quad \text{Condensation number}$$

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Part IV

Miscellaneous Topics in Industrial Drying

44 Dryer Feeding Systems

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44.1 INTRODUCTION

Feeders are devices that introduce a variety of materials into dryers at a controlled, specified rate. Usually, the feeder is located at the interface stage between material-handling equipment or upstream process and the dryer. The material-handling equipment may be a hopper or a bin whereas the upstream process may be a reactor, crystallizer, filter, hydrocyclone, centrifuge, etc.

Feeding systems of various types and configurations are used for storing, conveying, dosing, and controlling dryer feeds. It should be pointed out that

the bin or the storage tank and feeder should be regarded as an integral unit of the dryer system, and selected and designed as such. Without a properly designed bin to handle the wet solid, the feeder is useless. Conversely, without a proper feeder design the best bin may not produce the desired flow.

Dryer feeds are either wet solids of various physical forms or solids in liquids (solutions, slurries, sludges, emulsions, thixotropic pastes, etc.). Most dryers cannot accept feeds directly unless they are free flowing. A liquid needs the right range of viscosity, surface tension, and discrete size of particles that allow it to be pumped and, if necessary, sprayed.

A wet solid usually needs the right range of consistency, dryness, and particle size and shape that allow it to flow properly; otherwise, it has to be backmixed with enough dry products to make it flowable.

Feeding devices in drying systems for granular feedstocks are identical to those used for the transport of bulk solids; but careful selection must be made because not all transport equipment is suitable for handling wet solids, which may have poor flowability.

This chapter is intended to give a basic review of the feeding systems commonly used in various types of industrial dryers for granular, pasty, or liquid feedstocks. [Section 44.2](#) provides guidelines for the selection of widely used feeders and discusses the basic factors affecting their selection. [Section 44.3](#) describes the most commonly used solid-storage bins and their configurations. [Section 44.4](#) discusses five typical solids feeders, highlighting their configurations, design parameters, advantages, and limitations. Finally, [Section 44.5](#) gives a brief survey of the various feeding techniques employed in selected commercial dryers. In general, vendors of drying equipments also provide the necessary feeding equipment, which is often an integral part of the drying system. It is important to note that physical testing of the feeder for the specific feedstock is essential; small changes in feed consistencies may, in some cases, change feeder characteristics significantly.

44.2 SPECIFICATION AND SELECTION GUIDELINES

Numerous feeding system designs are available to accommodate the diverse properties of the materials handled. It would be difficult, if not impossible, to select a type and capacity without thoroughly defining the flow characteristics of the material. The flow behavior of bulk solids varies with the solids' physical properties and, with the way each acts under different atmospheric, loading, and unloading conditions. Optimum choices can eliminate costly and inconvenient flow stoppages or the uncontrolled flooding of materials—problems that often compromise feeder accuracy. Regardless of what type of feeder is used, it should provide the following [1]:

- Reliable and uninterrupted flow of materials from some upstream device (typically a bin or a hopper)
- The desired degree of discharge flow-rate control for the necessary range
- Uniform withdrawal of materials from the outlet of the upstream device, which is particularly important if a mass-flow pattern is desired, such as to control segregation, provide uniform

residence time, and minimize caking or spoilage in dead regions

- Minimal loads acting on the feeder from the upstream device, minimizing the power required to operate the feeder, as well as particle attrition and abrasive wear of the feeder components

Feeder selection is based on many factors, the most important of which is feed rate or capacity requirements. Capacity requirements can generally be divided into four distinct categories [2]:

1. **Operating capacity.** This is the average required particulate flow, allowing time for normal maintenance, repairs, and process delays that may occur.
2. **Rated capacity.** This is the average required particulate flow expected during ideal conditions, for specific time periods, as indicated by the process.
3. **Design capacity.** This is the particulate flow from which structural and mechanical design calculations are made.
4. **Circulating loads.** This is the amount of material returned from the equipment such as a separation device. This is expressed as a percentage of raw feed.

Once the system capacities have been determined, the process of reviewing preliminary particulate flows is sometimes called preselection guidelines. The most important selection and specification guidelines can be summarized as follows [3–7]:

1. The first step in selecting a feeding system is to classify the materials that will be handled according to its density, abrasiveness, flowability, and other characteristics such as toxicity, corrosiveness, tendency to pack, and so on. Recently, efficient techniques have been developed to predict, with reasonable accuracy, the handling characteristics of a solid as function of temperature, moisture content, size distribution, friability, and flowability. Some useful empirical tests for this purpose are described by Thomson [8]. The important properties measured along with the vital information they give are listed in [Table 44.1](#). [Table 44.2](#) discusses the flow characteristics of several classes of materials.
2. The geometry of the storage bin or hopper must be selected based on the material characteristics. Throat and gate dimensions should be at least two and one-half times the diameter of the largest particle of the material,

TABLE 44.1
Solid Properties That Affect the Selection of Feeding System

Property	Remarks/Applications
Bulk density	The unit weight of the bulk solid; indication of flow; use for feeder and hopper sizing; compressibility determination; possible compressive strength. Usually decreases with increasing moisture content.
Particle size distribution	Indicates flowable and nonflowable fractions. In cases in which the material is extremely fine (10 μm or less), material-handling selection becomes very critical.
Moisture content	Indicates the maximum moisture content a material can have and still be flowable to some degree. If the material has a tendency to stick on metal surfaces, the feed will have to be backmixed.
Hygroscopic	Indicates potential hazards to flowability, need for sealing and inert gas environment.
Particle specific gravity	Useful to determine porosity.
Adhesion	Useful in the design of hopper or bin, type of material for hopper or belt, shows if material will build up.
Corrosiveness	Useful for selecting the material of construction of the feed system. Properties affecting potential corrosiveness are pH and moisture content.
Abrasiveness	Useful for selecting the material of construction and the type of feeder; depends on the particle size and hardness.
Hardness	Useful for feeder design; indicates potential need for wear-resistant materials.
Solubility	Indicates whether a slurry or solution will result; indicates type of mixer needed.
Hazardous nature	Material's potential toxicity defined and precautions needed.
Lump size	Maximum lump size for correct hopper outlet; potential abrasiveness.
Temperature	Useful for determination of storage time, materials of construction. Solids may agglomerate or soften at high temperatures or may undergo phase changes when cooled.
Flowability	Shows the need for feeder accessories such as agitators, vibrators, rotors, special hoppers, and the like.
Effective angle of internal friction, angle of internal friction, kinematic angle of internal friction	These angles of flow can be used to calculate the discharge opening size required on a hopper to prevent arching and ratholing over the entrance to a feeder, predict hopper geometry required to promote gravity flow, estimate flowability and cohesiveness of different solids.
Cohesiveness	Characterizes the degree to which individual particles cling together; used in flow evaluation and assess the effectiveness of vibration in aiding the flow of solids.
Arching and bridging	Indicates type of agitation required.
Compressibility	Expresses the percentage difference between loose and packed density; can signal potential flow problems. Highly compressible solids may compact under head in a bin, making it difficult to initiate flow. Compressibility usually increases with moisture content.
pH	Indicates need for corrosion-resistant materials.

Source: Carr, R.L., *Chemical Engineering*, 76, 7–16, October, 1969; Johnson, J.R., *Chemical Engineering*, 85, 9–17, October, 1978.

and the width of the bin opening must be consistent with the capacity requirements and large enough to minimize bridging.

3. If the flow properties vary (i.e., because of changes in moisture content or temperature), the conditions of lowest flowability should be used to calculate the critical dimensions.
4. The mechanism used to discharge the materials from the feeder must be selected to provide the desired discharge rate without physically damaging the product.
5. The decision to operate the feeder volumetrically (discharging by volume) or gravimetrically

(discharging by weight) determines the type of feeder that is appropriate.

6. Establish the conveying distance and path, and the operating environment.
7. Assess the impact of property data such as lump size and size distribution, bulk density, and the like on feeder design. Hazardous properties or a tendency to dust will also restrict the choice of feeder.
8. Review previous experience with conveying the given feed.
9. Take stock of the geometry and the amount of space available in the plant and other

TABLE 44.2
Material Classification

Material Class	Description
I	Material is granular and free flowing; would normally flow out of storage unassisted, but temperature and moisture changes can cause it to bridge occasionally. Examples are granular salts, sugar, plastic pellets. Size: 1/4 in. to 50 mesh.
II	Material is a sluggish powder; would not normally flow by gravity alone. Examples are flour, limestone. Size: 100 to 300 mesh.
III	Material is a powder that tends to be readily adhesive or becomes easily fluidized. Examples are adhesive materials such as foundry sand, pigments; fluidized materials such as hydrated lime cement, talcum powder, confectionary sugar. Size: –325 mesh.
IV	Material is fibrous or flaky with a relatively low bulk density of 3 to 20 lb/ft ³ . Particle sizes are from 1/8-in. strands to 1 in. or larger chips; has a tendency to interlock and absorb vibration. Examples are wood chips, slivers, shavings, plastic scraps, flaked grains.

Source: Wahl, R., *Powder and Bulk Engineering*, May 1987.

- conveying areas, and note limitations that might affect the size or type of feeding equipments.
- Compare the suitable types with respect to their smoothness of operation and initial, operation, and maintenance costs.
 - The feeding system from the upstream section of the process to the dryer should be as short and direct as possible to prevent possible compaction of a sticky feed and the lodging of materials in bends and corners.
 - The operating range of the feeder should be wide enough to cover the feed-rate range required in the process, with a margin of 10%, so that the feeder can work well under the conditions of unexpected disturbances. Also,

the dynamic response should be as fast as possible [9].

- The normal procedure to size a feeder is to use the manufacturer's rated capacity, usually given in ft³/h (or m³/h) and to provide a design from past experience, rule of thumb, or empirical data.

Table 44.3 presents a brief selection guide for different feeding equipments used in drying based on the material characteristics. The information presented in Table 44.1, Table 44.2, and Table 44.3 is helpful as a general checklist for the preliminary selection of the feed system. Before a final decision is made on the type of equipment, method of operation, initial cost, operating and maintenance costs should be considered.

TABLE 44.3
Feeder Selection Guide Based on Feedstock Properties

Type of Feeder	Feed Size			Flowability			Abrasiveness		
	Fine	Granular	Large Lumps	Very Free	Free	Sluggish	Non	Mildly	Very
Apron		•	•	•	•	•	•	•	•
Belt	•	•		•	•		•	•	•
Reciprocating	•	•	•		•	•	•	•	•
Vibratory	•	•	•	•	•	•	•	•	•
Gravimetric weight	•	•	•	•	•	•	•	•	•
Mechanical vibrating	•	•		•	•		•	•	•
Bar flight	•	•	•	•	•	•	•	•	
Screw	•	•		•	•	•	•	•	
Rotary table	•	•		•	•	•	•	•	•
Rotary plow	•	•		•	•	•	•	•	•
Rotary vane	•	•		•	•		•	•	

Source: Link-Belt catalog no. 1000.

44.3 STORAGE OF PARTICULATE SOLIDS

Storage containers in the dryer feed system are required to smooth out possible discontinuities, variations, and interruptions in the supply of the feed materials. However, long-term storage of feed materials and oversized containers should be avoided to reduce the possibility of settling and caking. Generally, storage bins or hoppers come in three basic configurations [3,8,13–17]: funnel flow, mass flow, and expanded flow (see Figure 44.1) These configurations have the characteristics given below.

The characteristics of funnel-flow bins are:

1. Large storage capacity
2. Inexpensiveness
3. Adequacy when noncaking or nondegradable materials are to be stored
4. Used for materials for which segregation would not present a problem downstream
5. The first-in, last-out flow sequence often leads to particle segregation and degradation
6. Not suited for materials that may spoil or be damaged by long-term storage
7. Formation of dead storage zones reduces the effective capacity and affects signals from solids-level meters

The characteristics of mass-flow bins are:

1. The materials leave in a first-in, first-out manner, thus minimizing the problem of caking, degrading, or segregation during storage
2. Predictable flow velocity
3. Elimination of the formation of dead zones and thus elimination of the risk of material degradation
4. Constant solids density at discharge, making it possible to get good solids control with

volumetric feeders and improve the performance of gravimetric feeders

5. Elimination of erratic flow and ratholing and the associated problems of channeling and flooding of powders
6. The materials in the bin can act as a gas seal

The characteristics of expanded-flow bins are:

1. Combine the geometric characteristics of funnel- and mass-flow bins
2. Used to modify the existing funnel-flow bins in order to reduce or eliminate the flow problems mentioned earlier
3. Large capacity and high discharge rates
4. Remixing of segregated particles is better than in a funnel-flow design, but not as good as in a mass-flow one
5. Discharge flow rates are between the faster funnel-flow and the slower mass-flow bins

It is important to ensure that the maximum feed rate from the storage bin is always greater than the maximum expected operating rate of the feeder. Otherwise, the feeder will become starved and flow-rate control will be lost. This problem is particularly pronounced when handling fine powders because their maximum flow rate through an opening is significantly less than that of coarser-particle bulk solids whenever a mass-flow pattern is used [1].

Flow-aid devices can be used with storage bins in order to enable them to accommodate different materials and to reduce storage-bin problems of classification, ratholing, and bridging. Among them are [16–20]:

1. Sledge hammers
2. Vibrators
3. Pneumatic devices to assist materials flow
4. Fluidized air in the case of cohesive materials
5. Agitators to eliminate particle bridges

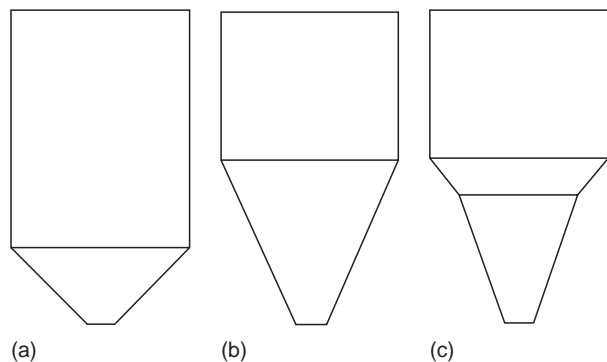


FIGURE 44.1 Basic bin configurations: (a) funnel-flow bin; (b) mass-flow bin; (c) expanded-flow bin.

44.4 BASIC FEEDER TYPES

Among the various feeder types (e.g., vibratory, screw, belt), there are two broad classes of feeders: volumetric and gravimetric. Volumetric feeders rely on known unit volume (as well as motor speed) to meter materials whereas gravimetric feeders rely on weighing and speed-sensing devices to control flow. If the process follows three simple rules, a volumetric feeder can be the most effective and economical choice. The rules are [2]:

1. Maintain a uniform feed bulk density by providing both an adequate level of particulates

- above the feeder inlet and an adequate supply bin residence time
2. Load the positive feed element uniformly and consistently so that the prefeed hopper provides a mass flow into the feeder inlet
 3. Precisely maintain uniform feeder element velocity (i.e., belt speed, screw rpm, etc.)

The motor drive should have tachometer feedback provisions to eliminate velocity fluctuations from screws, rotors, tables, belts, and so forth. Table 44.4 summarizes the major features of both volumetric and gravimetric feeders.

A wide variety of types and configurations of solid feeders have been developed to satisfy the many special requirements found in solid-processing plants [3–5,8,13,22–24]; as all of these cannot be discussed in this chapter, only feeders that are frequently used in industrial dryers are presented. The omission of others does not mean that they may not be suitable for such services. Other specialized dryer feeders, flow-aid devices, and liquid feeders are discussed in [Section 44.5](#).

44.4.1 SCREW FEEDERS

44.4.1.1 Principle of Operation and Equipment Configuration

Screw feeders are made with a long-pitch, steel, helix flight mounted on a shaft, supported by bearings

within a U-shaped trough [5]. As the element rotates, the materials fed into it are moved forward by the thrust of the lower part of the helix and are discharged through openings in the trough bottom or at the end into the downstream equipment. Screw flights near the discharge are typically notched, followed by double-bladed knife sections that cut up lumps and enhance the downward force necessary to handle sticky materials. Some basic types of flights and pitches are shown and described in Table 44.5.

The inlet must be flooded with materials by incorporating changes in the construction of the fighting (diameter, pitch, etc.) and the speed of the feeder screw; the material discharge can be governed to the desired rate. Feeders can be built with variable diameters or stepped pitches or both in units composed of one, two, or a multiple number of screws, depending on the application. Although screw feeders are available in many designs to fit different requirements, commonly used design configurations are shown in [Figure 44.2](#), [Figure 44.3](#), and [Figure 44.4](#).

44.4.1.2 Relevant Design and Operating Characteristics

The basic design of screw feeders depends on the material characteristics, the capacity required, the conveying distance, the weight of the material resting on the feeder screw, the dimensions of the feeder opening, and the power available. Screw feeders are

TABLE 44.4
Volumetric and Gravimetric Feeder Characteristics

Volumetric Feeders

Based on the discharge of a constant volume of the process material per unit time.
Simplest and the least-expensive feeding systems.
Operator must sample feeder output and adjust the speed accordingly.
Feed-rate repeatable accuracy is $\pm 2\%$ at best and is affected by variation in the bulk density of the material.

Used when precise solids control and exact dosages are not critical.

Do not indicate run out of material, bridges formation, or flow interruption.
Feed-rate out-of-tolerance shutdown and alarm are not available.
Can be integrated into a gravimetric system by adding a scale and an appropriate controller.

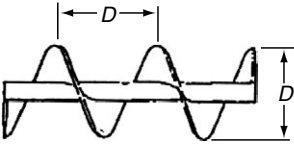
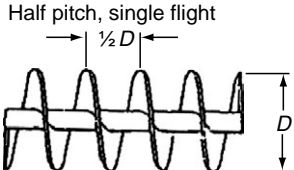
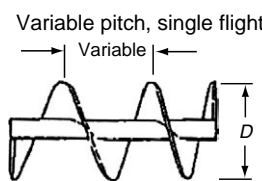
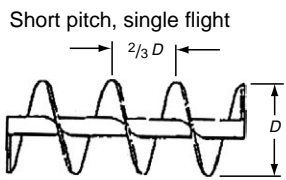
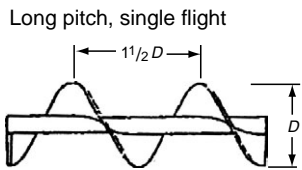
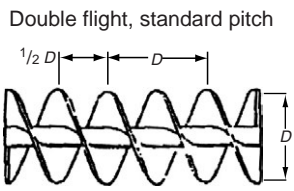
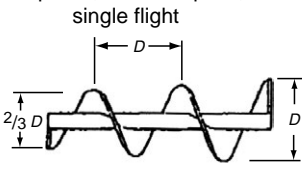
Gravimetric Feeders

Based on weighing materials discharge and control.

More expensive than volumetric units.
Feed-rate control at set point is ensured because there is a feedback controller to assure the correct operational speed.
Feed-rate repeatable accuracy is normally two to five times better than volumetric one irrespective of changes in bulk density or the occurrence of flow problems.
Used when exact dosage and high feed control and accuracy are needed.
Give immediate warning in case of any material interruption from the storage bin.
Feed-rate out-of-tolerance shutdown and alarm are not available.

Source: Wilson, D.H., Dumington, D.L., *Chemical Engineering*, 98, 72–81, August, 1991; Thomson, F.M., *Chemical Engineering*, 85, 77–87, October, 1978; Fahlenbock, T., and Paul, B.O., *Chemical Processing*, April, 1997.

TABLE 44.5
Basic Screw Feeder Flight and Pitch Types

Configuration	Description
<p>Standard pitch, single flight</p> 	<p>Screws with pitch equal to screw diameter are considered standard and are suitable for a wide range of materials.</p>
<p>Half pitch, single flight</p> 	<p>Similar to short pitch, except pitch is reduced to 1/2 standard pitch. Useful for vertical or inclined applications and for extremely fluid materials.</p>
<p>Variable pitch, single flight</p> 	<p>Flights have increasing pitch and are used in feeder screws for uniform withdrawal of fine, free-flowing materials over the full length of the feed opening.</p>
<p>Short pitch, single flight</p> 	<p>Flight pitch is reduced to 2/3 diameter. Recommended for inclined or vertical applications. Shorter pitch retards flushing of fluidlike materials.</p>
<p>Long pitch, single flight</p> 	<p>Pitch is equal to 1/2 diameter. Useful for agitating liquids or for rapid movement of very free-flowing materials.</p>
<p>Double flight, standard pitch</p> 	<p>Double flight, standard pitch screws assure smooth, regular material flow and uniform movement of special types of materials.</p>
<p>Tapered, standard pitch, single flight</p> 	<p>Screw flights increase from 2/3 to full diameter. Used as feeder screws for uniform withdrawal of lumpy materials. Generally equivalent to and more economical than variable pitch flight.</p>

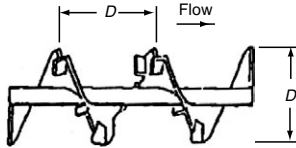
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TABLE 44.5 (continued)
Basic Screw Feeder Flight and Pitch Types

Configuration

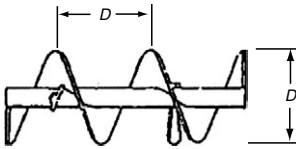
Description

Cut-and-folded flight,
standard pitch



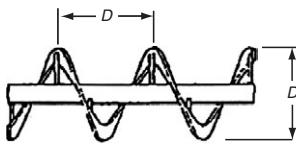
Folded flight segments lift and spill the materials. Partially retarded flow provides thorough mixing action when heating, cooling or aerating light substances.

Standard pitch with paddles



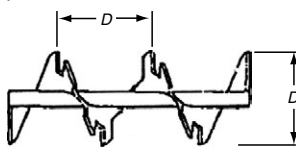
Adjustable paddles positioned between screw flights oppose flow to provide gentle but thorough mixing action. Paddles can be set at any angle to produce desired agitation.

Single-flight ribbon



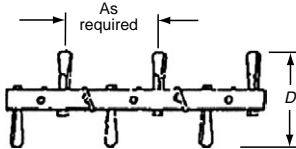
Best for feeding sticky or viscous materials. Open space between flighting and pipe eliminates buildup of the material on the flights.

Single-cut flight, standard
pitch



Screws are notched at regular intervals at outer edge for mixing action and agitation of material in transit which tends to pack.

Paddle



Adjustable paddles provide complete mixing action and controlled material flow.

typically available with a volumetric capacity ranging from 0.0001 to 300 m³/h for single feeder and are normally operated at 2 to 40 rpm. The feed capacity of a horizontal screw feeder submerged in solids is calculated by multiplying the volume contained in one pitch by the rpm, bulk density, and efficiency. Two main factors contributing to the power needs of a screw feeder are: (1) friction of the mechanical parts and (2) power required to move the material, which is proportional to the amount of the material to be moved and the conveying distance. A number of guidelines and rules of thumbs have been developed for designing screw feeders for optimum performance.

These guidelines are well summarized by Marinelli and Carson [22].

Advantages of the screw feeder are:

1. Versatile enough to handle a wide range of materials, from powders to lumps
2. Excellent volumetric flow efficiency
3. Relatively low initial cost, low maintenance cost, and inexpensive to replace
4. Flexible arrangement, can operate at any angle, and has few moving parts
5. Used when the material requires mixing or aeration for breaking up lumpy materials

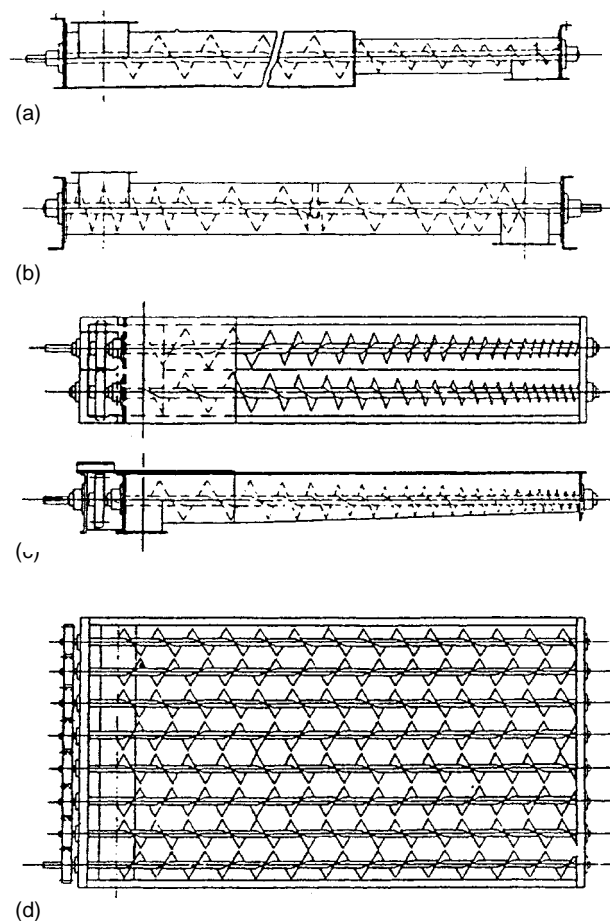


FIGURE 44.2 Screw feeder configurations: (a) multiple-diameter feeder; (b) short-pitch feeder; (c) variable-pitch, twin-tapered feeder; (d) live-bottom feeder. (From Screw Conveyor Corporation.)

6. An enclosed device, which makes it particularly important for safety or prevention of contamination when dealing with fumes or toxic materials
7. Construction with a jacket when preheating of the feed is required

Disadvantages and limitations of the screw feeder are:

1. Screw feeders are limited to lumps that are smaller than the smallest pitch and to bin openings less than 6-screw diameters long.
2. Abrasive solids are not often handled unless the screw speed is slow.
3. The standard full-pitch screw has a pitch-to-diameter ratio of 1; this configuration is satisfactory only when the maximum hopper opening does not exceed 1 to 1.5 pitches. Using larger opening leads to serious flow problems in the bin and unnecessarily high horsepower requirements.

4. Sticky or very cohesive materials can build up in short-pitch sections and conveying will cease. In this situation, special screw configurations should be used.
5. The allowable loading and screw speed are limited by the characteristics of the material. Light, free-flowing, nonabrasive materials fill the trough deeply, permitting a higher rotating speed than with heavier and more abrasive materials.
6. Single-span screws are normally limited to about 6 m in length due to the deflection of longer center shafts.

44.4.2 VIBRATING FEEDERS

44.4.2.1 Principle of Operation and Equipment Configuration

In these feeders, materials are held in a pan or tray. As the pan vibrates, the particles hop down the tray toward the discharge outlet. The motion of a bulk material when subjected to cyclic excitation is extremely

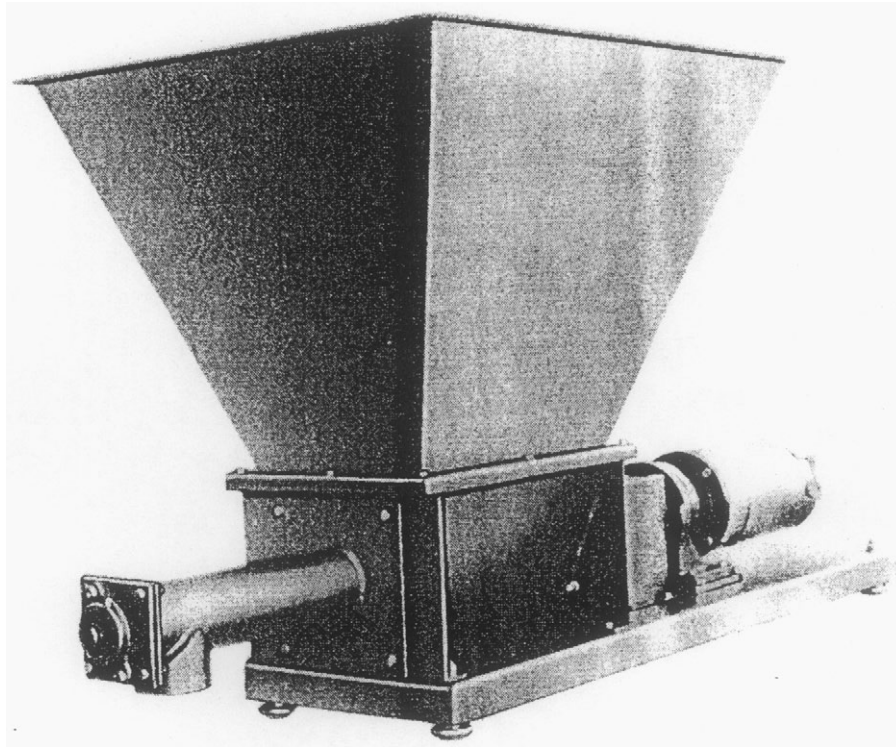


FIGURE 44.3 Volumetric screw feeder. (From Metalfab Inc.)

complex because of the interrelationship of the material properties superimposed on the vibrating characteristics. The equipment may be open-trough design or fully enclosed. The trough that carries the materials rests on a spring system with stabilizing elements. The deck can be a solid or a mesh that allows a cooling medium to move through the materials being conveyed. Two general types of vibrating feeders are available [4,8]: direct force (single mass) and indirect force (tuned two-mass), and two excitation systems are in common use: electromagnetic and electromechanical.

Vibrating feeders are available in both fixed-rate and variable-rate designs. Fixed-rate units operate at full line voltage. The capacity of fixed-rate units can be changed by a mechanical adjustment of the eccentric weight on the exciter drive. Variable rate is obtained by changing the voltage to the drive motor. For maximum control, the feeder is generally mounted horizontally and operated with a reduced material depth. For maximum capacity, the feeder is mounted on a decline and carries a deep material depth. Figure 44.5 shows a typical vibrating feeder.

44.4.2.2 Relevant Design and Operating Characteristics

Factors for determining the flow rate from a vibratory feeder include tray frequency and displacement, head

pressure of the materials on the tray, width and depth of loading on the trough, linear speed along the trough, and material properties such as bulk density, particle size, and angle of repose. These features taken singly or in combination, as well as the hopper design, are important to obtain the required feeder capacity. Vibratory feeders are commonly built with a capacity range of 0.015 to 1200 m³/h and are normally operated at frequencies from 12 to 60 cps.

Advantages of the vibrating feeder are:

1. Gentle handling of materials; hence can be used to move materials (e.g., abrasive or friable solids) that could damage, or be damaged by, screw or belt feeders
2. Precise feed control
3. Cleanliness, no need for moving parts and lubrication
4. For safety reasons, can be enclosed to prevent product contamination; also, toxic materials can be readily handled
5. Robust; can handle a wide range of feed rates, especially if used as a gravimetric feeder
6. Simple construction with low headroom

Disadvantages and limitations of a vibrating feeder are:

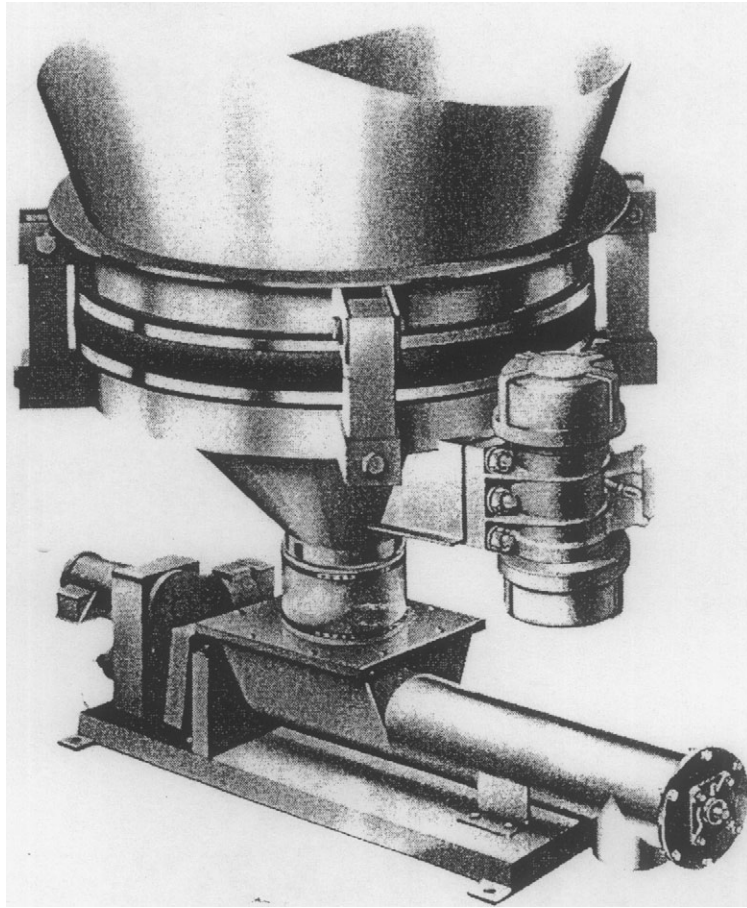


FIGURE 44.4 Screw feeder with bin activator. (From Metalfab Inc.)

1. Limited turndown with conventional controls, and are somewhat nonlinear over the turn-down range. This nonlinearity is particularly problematic during volumetric discharge as frequent calibrations may be required to guarantee accurate feed rates.
2. They are limited for the most part to feeding from round, square, or slightly elongated opening.
3. They should be used with caution for cohesive solids because the vibratory motion may cause the materials to pack in the hopper.

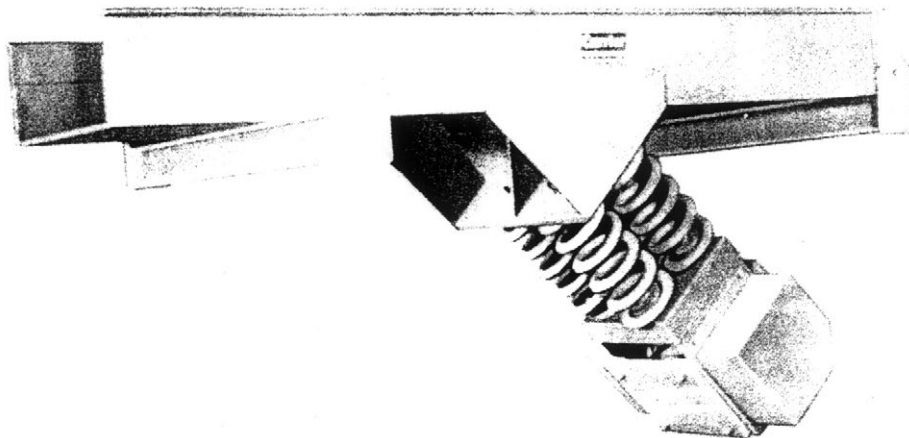


FIGURE 44.5 Vibrating feeder. (From Carrier Vibrating Equipment Inc.)

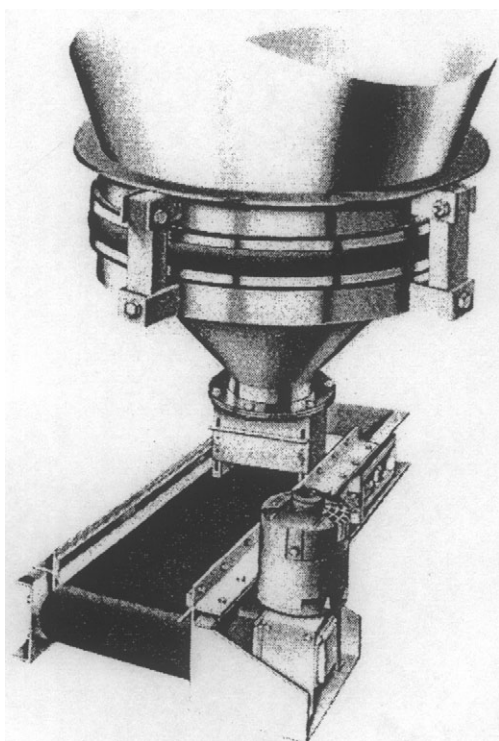


FIGURE 44.6 Belt feeder with bin activator. (From Metalfab Inc.)

4. Vibration can cause consolidation and degradation of materials.

44.4.3 BELT FEEDERS

44.4.3.1 Principle of Operation and Equipment Configuration

Most belt-feeder designs feature a fabric or an elastomer-covered, fabric-reinforced band, supported by idlers or plates. Materials are sheared from the storage hopper above and delivered to the belt. Feed rates are controlled by adjusting the gate that controls the depth of materials on the belt. More precise control can be achieved by measuring the belt speed and the weight impressed on one or several of the belt idlers to compute and control the weight rate of feed.

The main components of the system are the belt, the idlers, the pulleys, the drive, and the structure that supports and maintains the alignment of the idlers and pulleys and support the driving machinery [16]. Belt feeders are available in a gravimetric configuration as well as a weigh-belt feeder. Figure 44.6 shows a belt feeder with a bin activator.

44.4.3.2 Relevant Design and Operating Characteristics

Feed rate may be expressed in terms of bulk density, belt width, depth of loading, and belt speed. Belt widths

range from 0.2 to 2 m and belt speeds run from 0.3 to 40 m/min. The volumetric capacity of belt feeders ranges from 0.03 to 700 m³/h and is determined by the belt speed and by the profile of the materials discharged to belt.

Advantages of the belt feeder are:

1. Versatile; can handle a great variety of materials; can handle very heavy impact and solids load from the hopper above. Particularly adept for applications with high flow rates
2. Can withdraw materials from very long slot openings in bins
3. Low in cost, simple to maintain
4. Low power consumption
5. Ideal for bridging a long gap between an existing hopper and the dryer

Disadvantages and limitations of the belt feeder are:

1. Improperly designed feed hoppers can cause solids compaction, wear of belts, and high horsepower demands.
2. Depending on belt width, lump size can be a limitation.
3. Sticky particulates need special consideration, as any sticking of the wet materials on the belt will increase the variations in the feed rate.
4. Temperatures above 65°C should be approached with caution.
5. Some solids react with rubber in the belt necessitating a special covering material for the belt.
6. Maintenance is needed depending on the materials handled.

44.4.4 ROTARY TABLE FEEDERS

44.4.4.1 Principle of Operation and Equipment Configuration

The rotary table feeder is a circular plate or a disk mounted off-center and sealed against the bottom of the feed hopper (see Figure 44.7). As the motorized table rotates, a groove in it fills with materials from the hopper above and a knife or plough scrapes the materials from the table to the intake of the dryer. Flow can be controlled by changing the height of an adjustable feed collar or by regulating the variable speed drive of the table.

44.4.4.2 Relevant Design and Operating Characteristics

Capacity of rotary table feeders, ranging from 0.000013 to 0.034 m³/h, is a function of material

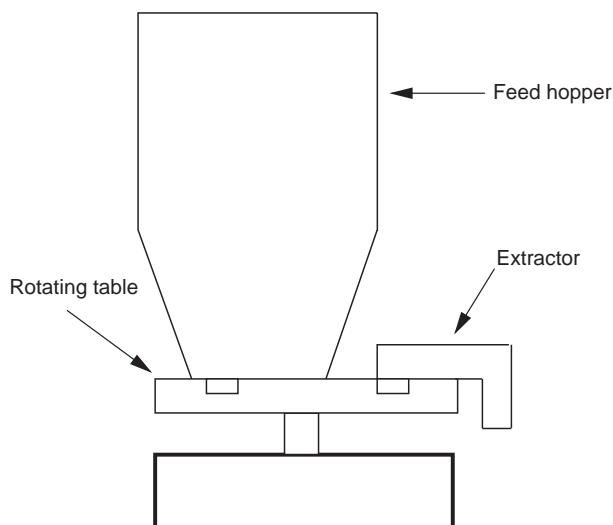


FIGURE 44.7 Rotating disk or table feeder.

density, speed of the disk, and geometry of the groove. The table rotates at about 2 to 10 rpm.

Advantages of the rotary table feeder are:

1. Table feeders can be used for materials that require wide hopper openings to eliminate arching, for sticky granular materials and for abrasive materials
2. Work well with materials that are highly cohesive or have very high shear factors
3. Relatively free of spillage problems

Disadvantages and limitations of the rotary table feeder are:

1. Limited to relatively small particle sizes
2. Imprecise feed control
3. Relatively more expensive than other feeders (e.g., belt feeder)

44.4.5 ROTARY FEEDERS

44.4.5.1 Principle of Operation and Equipment Configuration

Rotary feeders use a motor-driven rotor, enclosed in a cast-iron or forged-steel housing. The blades on the rotor have a star-shaped cross section. They can be machined with close clearances between rotor and housing, ensuring that the materials in the feed hopper remain isolated from the discharge stream. Rotary feeders can be used under a circular, rectangular, or slot openings and are available in four body configurations: drop through, blow through, side entry, and pellet shield [3]. The rotors generally are

of three basic configurations: open, shrouded, and shallow pocket rotor [3,4]. Some rotary feeder configurations are shown in Figure 44.8.

44.4.5.2 Relevant Design and Operating Characteristics

The optimum selection and sizing of rotary feeders depend on the required volumetric rate, the operating temperature, the differential pressure across the air lock, and the material properties. Rotary feeders are commonly available in capacities ranging from 0.015 to 600 m³/h [3] and have been built for pressure differences up to about 150 kPa [25].

Advantages of the rotary feeder are:

1. Suitable for feeding fine powders
2. Suitable for feeding nonabrasive, free-flowing materials
3. Serves as gas pressure seal to pass solids from one pressure environment to another
4. Provides very uniform withdrawal along a slot opening

Disadvantages and limitations of the rotary feeder are:

1. Rotary feeders are not usually considered for feeding very abrasive materials as close-clearance parts can wear rapidly in this case.
2. Sticky or cohesive solids stick and bind within the relatively small ports of the rotary valve, gradually build up in rotor pockets, and significantly reduce capacity.

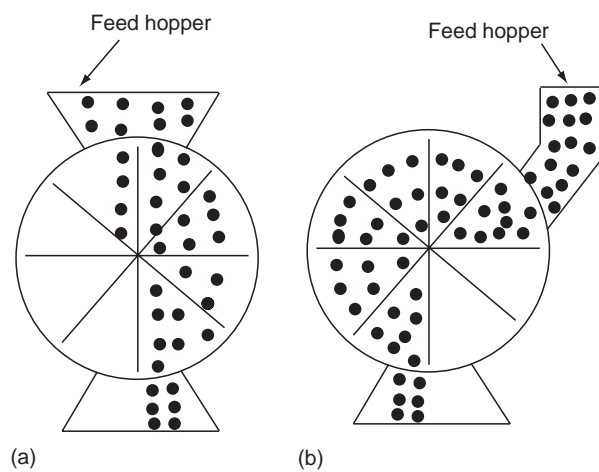


FIGURE 44.8 Some rotary feeder configurations: (a) drop-through rotary feeder; (b) side-entry rotary feeder.

44.5 FEEDERS FOR INDUSTRIAL DRYERS

This section describes the more commonly used feeding systems for selected major industrial dryers. Specialized and less prevalent dryers have been excluded from this coverage; additional information can be obtained from the sources cited at the end of this chapter. Further, we restrict our attention of dryers for particulate, pasty, or liquid feeds. For most types of dryers, the feed-handling equipment is part of the dryer itself and must be designed accordingly.

44.5.1 FLASH DRYERS

Flash or pneumatic dryers are particularly suited for drying powdery, granular, crystalline, or pasty products, which are obtained from centrifuges, rotary filters, or filter press. The selection of the correct feed system for flash dryers to obtain thorough initial dispersion of solids in the gas stream is of major importance. The short residence time (in seconds) in the dryer requires a homogeneous feed material and very rapid, even dispersion of the wet material into the drying airstream. Moreover, the feeding mechanism should supply nonlumpy wet solid into the stream of the drying air with a very controlled flow rate. Otherwise, the adjusted residence time in the dryer will be changed, resulting in off-specification of the product quality.

In most cases, feeds for flash dryers must be granular and free flowing when dispersed in the gas stream. Such products neither stick on the conveyor walls nor agglomerate. Also, particles should not be so heavy as to drop out of the carrier gas. Consequently, the feed system should contain all the necessary elements for even feeding and optimum mixing of the materials and the drying air [26].

Several variants of this feed system exist, each capable of handling different kinds of products. There are specially designed feeders for sticky products and feeders for products where attrition is to be minimized, as well as the more conventional disintegrators feeding the dryer. Finely pulverized materials with proper size distribution are compressed and fed to the dryer by a screw pump. Alternatively, venturi feeders, screw feeders, rotary air locks with rotating table feeders, mixers, dispersers, and disintegrators are used singly or in combination to control the rate of feed and convey the wet material directly into the hot airstream. A number of sophisticated feeders and flow-aid devices have been developed to provide more even dispersion in the drying medium. A typical feed system for a flash dryer is shown in Figure 44.9.

If the material is temperature-sensitive, i.e., sticks to hot metal surfaces, has a tendency to burn, or cannot be easily dispersed in a hot gas stream, then it is a common practice to introduce the material

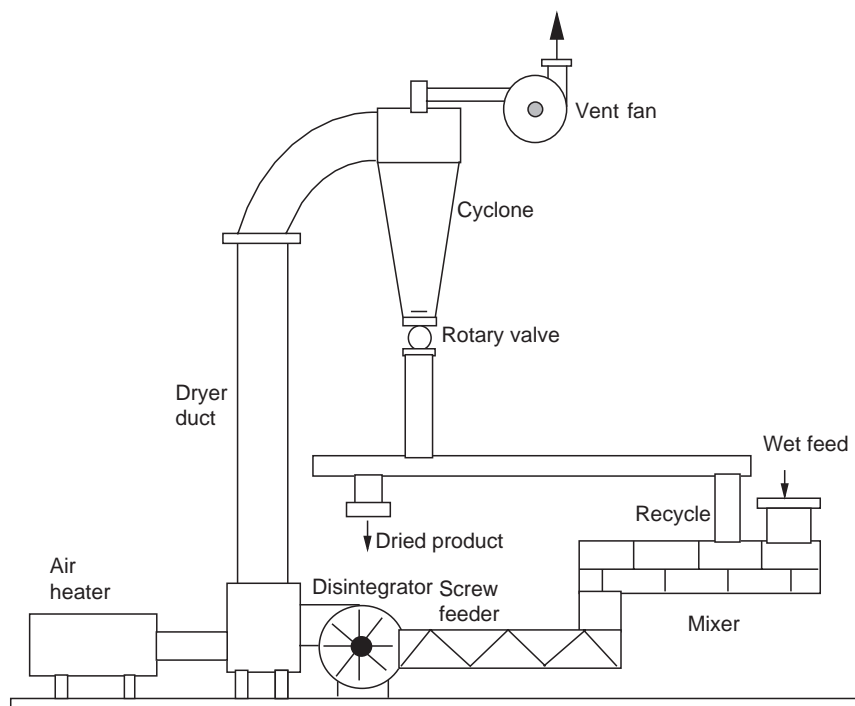


FIGURE 44.9 Flash-dryer feed system.

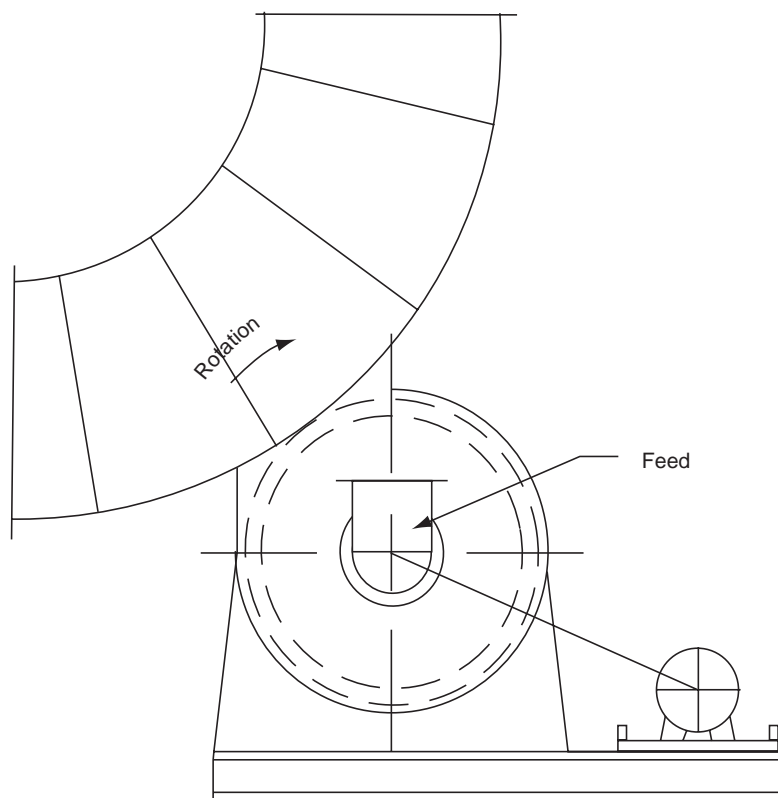


FIGURE 44.10 Sling for a flash dryer. (From Rosin Engineering Company Ltd.)

parallel with the hot gas stream by a dispersion feeder [26–28]. This type of feeder is similar to a material-handling fan with straight blades. The material introduction and discharge points are along the periphery of the fan housing 180° from each other. The centrifugal force generated by the fan wheel propels the material parallel to the gas stream path thus dispersing it [29].

If the feed is likely to agglomerate, form lumps, or is in the form of a sludge or filter-press cake, a mechanical disintegrator like a sling (Figure 44.10), a simple mill, or a cage mill can be included in the circuit after the feed point [28–31]. A cage mill can be placed at the bottom of the flash-drying column to assist in the drying of wet cake feeds containing lumps that cannot be disintegrated by the airstream. The cage mill is a low-shaft-speed machine with a rotor that contains paddles and round bars which break up the lumps and move the materials up the drying column by the centrifugal force [28].

The disintegrator (Figure 44.11) breaks up the wet material and provides a steady stream of finely divided material, flowing in the direction of the drying medium, which disperses quickly in a venturi so that it is spread evenly into the drying gas for maximum efficiency of drying. When these devices do not make the drying sufficiently uniform, a classifier can be

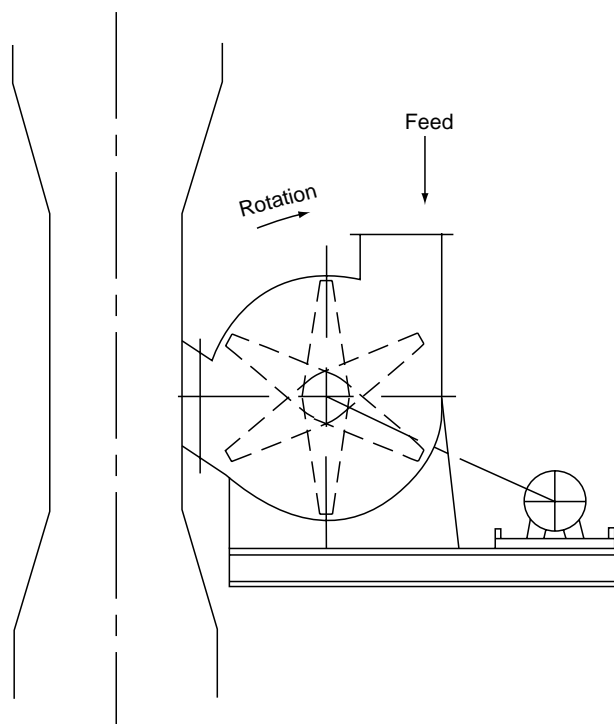


FIGURE 44.11 Disintegrator to break up wet material in a flash dryer. (From Rosin Engineering Company Ltd.)

installed at the top of the dryer tube to remove fines and to recycle the coarse fraction for another pass through the flash dryer.

For some materials that need particularly gentle mechanical handling, such as crystalline materials, Barr has developed the so-called cascading screen [26]. This feeder is similar to the disintegrator but uses a paddle-wheel rotor with a large-aperture screen plate at the outlet. The product is dispersed by the gentle action of a slow-running rotor passing the material through the screen plate, where it is broken up by impact with the edges of the holes.

High-initial-moisture materials can be processed by backmixing some recycled dry products into the fresh feed in order to improve consistency and handling characteristics of the feed. This process is usually carried out in a paddle mixer in which the incoming wet material is mixed with the recycled stream of dry products to obtain a nonsticky feed that is then distributed in the drying gases by the sling. At the same time, the sling breaks up agglomerates. It should be noted that the capital cost of backmixing equipment and the operating costs associated with it can be a substantial fraction of the overall flash-drying costs in some cases.

Another special type of flash dryer that incorporates the characteristics of both flash and fluid-bed dryers is the spin flash dryer [27,29]. This dryer can be used for the drying of high-solids-content feeds that cannot be handled in other conventional dryers. Examples of these feeds are filter cakes, cohesive pastes, and moist granules. Dilatent fluids can also be processed in this dryer. The feed system in spin flash dryers is one of the two types, depending on the feed. For solid feeds, a feed tank (or vat) is fitted with a low-speed center agitator that breaks up the lumps of the cake or paste to uniform consistency and gently forces it down into a screw feeder. Figure 44.12 illustrates this feeding system. In the case of dilatent fluid feed, the agitated tank and screw feeder would be replaced with a pump and several liquid injection ports to the drying chamber.

44.5.2 FLUID-BED DRYERS

Most conventional fluid-bed dryers operate with somewhat granular feeds in the size range from 100 μm to 1 mm, with a reasonably narrow distribution. The particles must not be excessively sticky and they must be resistant to lumping. An essential feature of any

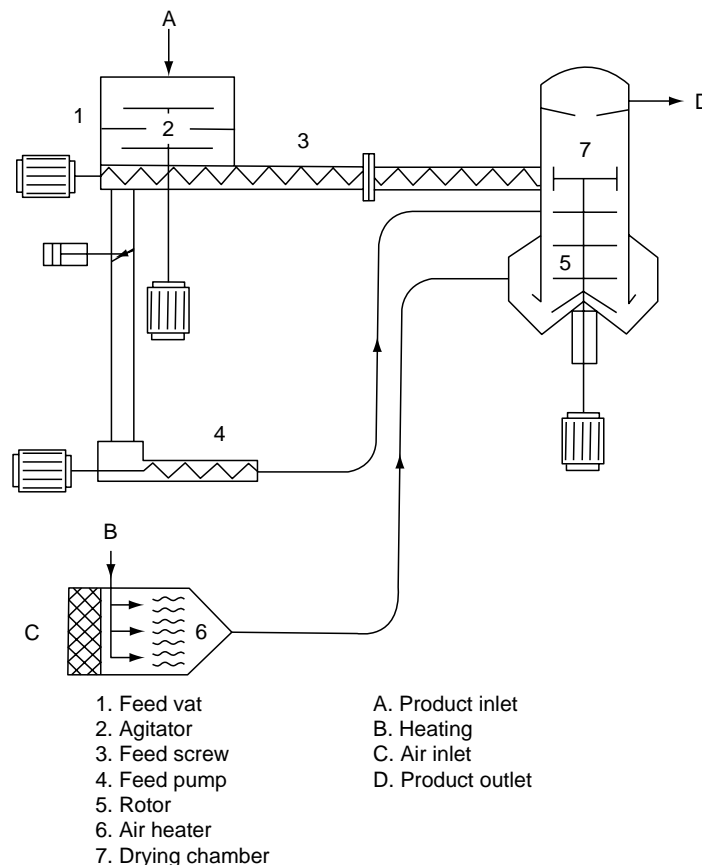


FIGURE 44.12 Spin flash-dryer feed system. (From APV Crepaco Inc.)

arrangement for feeding material is that the material should be in a dispersed form. However, some wet and sticky feeds, or even liquid feeds, can be brought to the right consistency and made suitable for fluidized-bed drying. Wet feeds can be backmixed with enough dry products followed by suitable preforming. Liquid feeds can be processed using a fluid-bed unit of inert particles and spray nozzles within the bed. The feed is introduced into the drying chamber at the top of the fluid bed toward the center. The feed should not be introduced near the wall, i.e., the feeder must extend into the freeboard over the bed and then release the feed material [32]. Often, it is desirable to spread the feed evenly over the top of the bed.

Suitable feed mechanisms for fluid-bed dryers are screw feeders, star feeders, and rotary valves. When the moist material is free flowing, it can be supplied from a steeply sided bunker down a wide pipe terminating near the base of the bed [28]. The bunker can then be equipped by any suitable form of solids conveyor. If the material is in the form of wet cake, it is dispersed with uniform particle distribution by a rotary feeder. A liquid solution can be fed using a rotary feeder or spray nozzle. The nozzles may be located at various positions, including within the bed itself.

Continuous fluid-bed dryers can be broadly classified into two major groups: backmixed and plug flow. Specialized feeding systems have been used for each of these two major types. Screw and rotary lock feeders are commonly used in the circular, well-mixed fluid beds. For deep-bed, rectangular, plug-flow fluid-bed dryers, distribution and acceptance of the feed is relatively easy. However, in the case of shallow beds, the feed must be distributed over the full width of the dryer. For this, it is typical to receive the output from the feeder on a chute that has triangular baffles [32]. If a well-mixed fluid bed precedes a plug-flow unit in a two-stage operation, the well-mixed unit can be located above the plug-flow unit fed directly by gravity.

44.5.3 ROTARY DRYERS

Rotary dryers are used to dry solids that are relatively free flowing and granular when discharged as product. With the right combination of features, even some liquids and sludges can be sprayed into a bed of inert material and dried. Feed materials that are not completely free flowing, such as wet or sticky feeds, are handled by external backmixing with a recycled portion of the dry product to form a free-flowing feed or fed directly into a bed of dry materials.

The most universally applicable mixer suitable for recycle mixing is the double-shaft, pug-mill-type mixer [28]. This mixer should be insulated to prevent excessive heat losses from the hot, dry recycle prod-

uct. Moreover, a surge tank of dry recycle product should be installed to ensure uniformity in the mixing operation and to prevent possible interruptions of the product from the dryer.

The method of feeding rotary dryers depends upon the material characteristics and the location and type of upstream-processing equipments. Preferred feeders are chute, screw, and vibrating units [28,31,32]. A chute can be used when the feed comes from above. To help seal against leaks, or if gravity feed is not convenient, a screw feeder is normally used. Good air sealing is necessary because ingress air reduces heat economy whereas the reverse causes dust problems.

Rotary dryers may operate cocurrently or countercurrently. For direct cocurrent operations, the dryer exhaust gases can be used to convey, mix, and predry the wet feed. Here, the wet material is introduced into the exhaust at a high velocity, separated in a cyclone, and then dropped into the feed point of the dryer. In the cocurrent mode, the feeder may overheat when contacting the inlet hot drying gas. In this case, cooling with cold-water jacket may be required to prevent the risk of overheating the feeder metal wall with the resulting scaling or overheating of heat-sensitive materials. If the dryer operates countercurrently, the heat load of the feeder is limited and normally no special construction is required.

44.5.4 CONVEYOR DRYERS

Conveyor dryers require that the feed materials should be in a state of granular or pelleted subdivision. Further, the feed must cover the conveyor band evenly in order to establish and maintain a structural configuration that will permit the free passage of air through the bed to avoid air channeling and hence nonuniform drying.

Various types of feeding arrangements are available to spread or distribute the wet material over the width of the conveyor. Many feeds can be introduced into the dryer without special preparation. Others require special pretreatment or preforming to make them suitable for thorough circulation of air. Fibrous, flaky, or free-flowing materials are usually loaded directly onto the conveying band without preforming using feeders such as oscillating or vibrating belts, spiked drums, or belts feeding from bins or hoppers (Figure 44.13a) [28].

A comprehensive range of feed-preforming equipment depending on the physical state of the wet solids is commercially available. Steam-heated finned drums have been used as a means of producing a partially dried, preformed feed. However, such equipment is usually much more expensive than many of the mechanical extruders. Pasty materials can often be

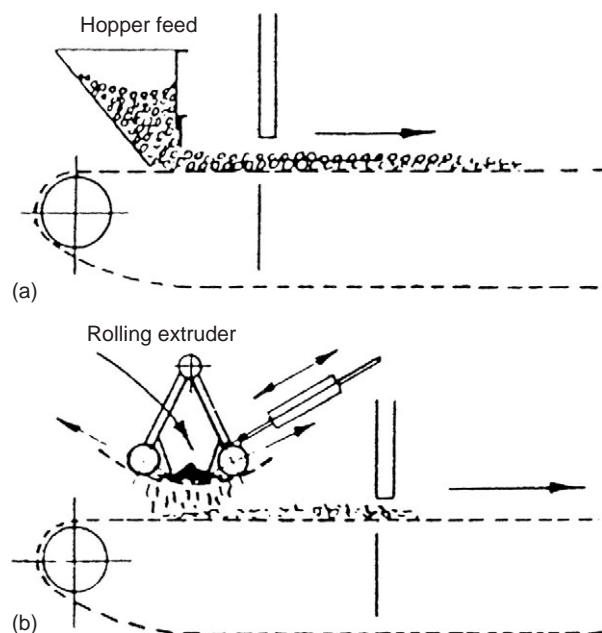


FIGURE 44.13 Typical conveyor feed devices. (From Proctor & Schwartz Ltd.)

performed by extrusion (Figure 44.13b). Generally, extruders operate with rubber-covered rollers moving over a perforated die plate. The rolls are driven back and forth on a curved perforated plate mounted across the full width of the conveyor. The feed in the form of a pressed cake is distributed into a hopper supported between the oscillating rollers from which it is forced by the rollers through holes in the plate [29]. Other feeders of the pressure type employ a gear pump arrangement, with extrusion taking place through a series of individual nozzles, while some use screw feeds, which usually are set up to oscillate in order to obtain effective coverage of the band. Alternate designs include rotating cam blades or conventional bar-type granulators.

Thixotropic filter cakes can often be preformed by scoring by knives. In few cases, powders can be pelleted in briquettes using briquetters or pelletizers to eliminate dustiness and permit drying in conveyor dryers [28]. A detailed description of the performing methods, types of machine used, and laboratory test procedures can be found elsewhere [25,28,29,33].

44.5.5 SPRAY DRYERS

Spray drying is a unit operation involving the conversion of a pumpable liquid feed to a dry particulate solid by means of contact with a hot drying medium. Solutions, emulsions, nonsettling suspensions, and slurries can be spray dried. It must be remembered that to be suitable for spray drying, the feed must be in pumpable condition or should be made so by dilution.

The feed system for spray drying consists of the following components [34]:

1. Product feed tank
2. Feed strainer or filter
3. Water feed tank
4. Feed pump
5. Atomizer

A feed system layout for spray dryers is shown in Figure 44.14.

A feed pump transfers the product to the atomizer either directly or via a constant-head feed tank. The feed tanks are of sufficient volume to permit continuous operation and to smooth out possible interruptions and variations in the supply of the material. It is quite usual for two holding tanks to be used alternately, thus assuring constant product supply to the dryer. The feed strainer or filter is important because all matter likely to jeopardize the performance of the dryer through partial or total blockage of the atomizer must be removed. Demineralized water or clean condensate is usually required for use in cases of feed blockage.

Pumps are required in the feed system either to transfer product within the feed system or to supply product to the atomizer. The pumps in the spray-dryer feed system are sized and selected for each application depending on the type and properties of the feed material and also depending on the atomizer incorporated in the feed system. When rotary atomizers or two-fluid nozzles are used, low-pressure pumps are also usually used. When pressure nozzles are employed, high-pressure pumps are required [34]. Various types of pumps are used because of the variety of feed materials that are encountered (see Table 44.6). A water tank for dryer start-up and shutdown is mounted on the atomizer feed pump.

The atomizer is the most important element in the spray-drying plant. Four types of atomizers are used in industrial drying [34–36]:

1. Rotary, with atomization by centrifugal energy (Figure 44.15)
2. Pressure nozzle, with atomization by pressure energy
3. Pneumatic nozzle, with atomization by kinetic energy
4. Sonic nozzle, with atomization by sonic energy

The classification of atomizers used in spray drying is summarized in Figure 44.16.

The selection of the proper atomizer is governed by the characteristics of the feed material and the dried-product specifications. In cases for which more

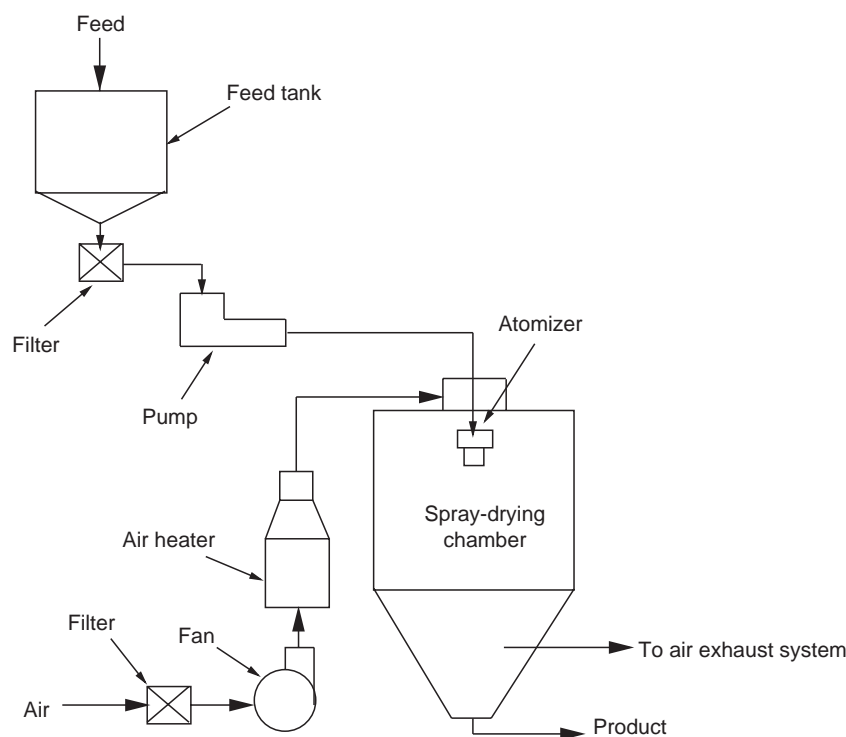


FIGURE 44.14 Spray-dryer feed system. (From Niro Inc.)

than one atomizer type is suitable, the rotary atomizer is generally preferred due to its greater flexibility and ease of operation. For more detailed information on atomizers, other relevant chapters in this handbook can be referred to.

In certain cases, it may be advantageous to preheat or pretreat the feed before spaying it into the dryer in order to increase the atomizer performance [33,34]. Pretreating and preheating equipment is

TABLE 44.6
Pump Selection for Spray Dryers

Pump Type	Feed Material/Atomizer Type
Screw	Low-pressure applications: milk, tomatoes, pharmaceuticals, and clay. Rotary atomizers or two-fluid nozzles.
Gear	Heavy pastes
Diaphragm	Slurries containing large, irregular, insoluble solids.
Centrifugal	Solutions and slurries
High-pressure piston or plunger	Pressure-nozzle atomization arrangements and in-line homogenization

Source: Masters, K., *Spray Drying Handbook*, 5th ed. Longman Scientific and Technical, London, 1991.

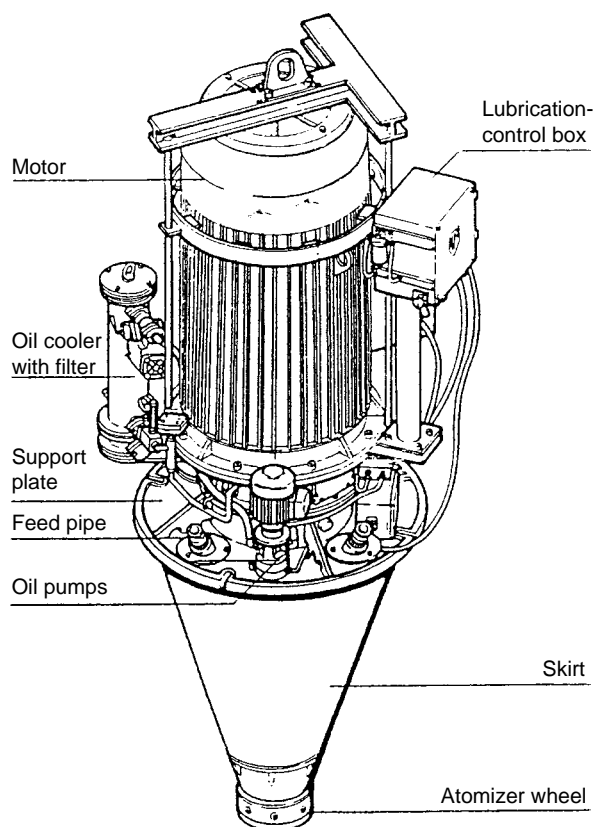


FIGURE 44.15 Rotary atomizer. (From Niro Inc.)

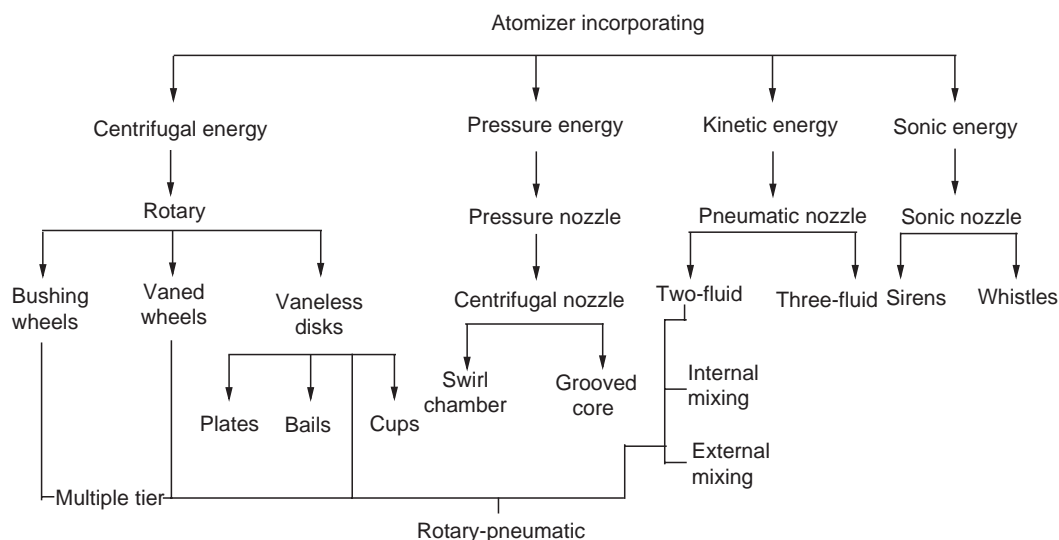


FIGURE 44.16 Classification of atomizers. (From Masters, K., *Spray Drying Handbook*, 5th ed., Longman Scientific & Technical, London, 1991; Barbosa-Cánovas, G.V. and Vega-Mercado, H., *Dehydration of Foods*, Chapman & Hall, New York, 1996.)

connected into the feed system just before the transfer of the product to the atomizer. In other cases, such as in the processing of pharmaceutical products for which any form of powder contamination must be avoided and a high degree of aseptic conditions is required, specially designed cleaning and sterilization systems should be installed in the feed section (see Figure 44.17). The air-cleaning system is based

upon the use of high-efficiency particulate air filters (HEPA) with a removal efficiency of 99.99%.

44.5.6 DRUM DRYERS

Drum dryers are used for drying liquids, liquid suspensions, or paste-like materials, especially those that easily adhere to metal surfaces and are, therefore,

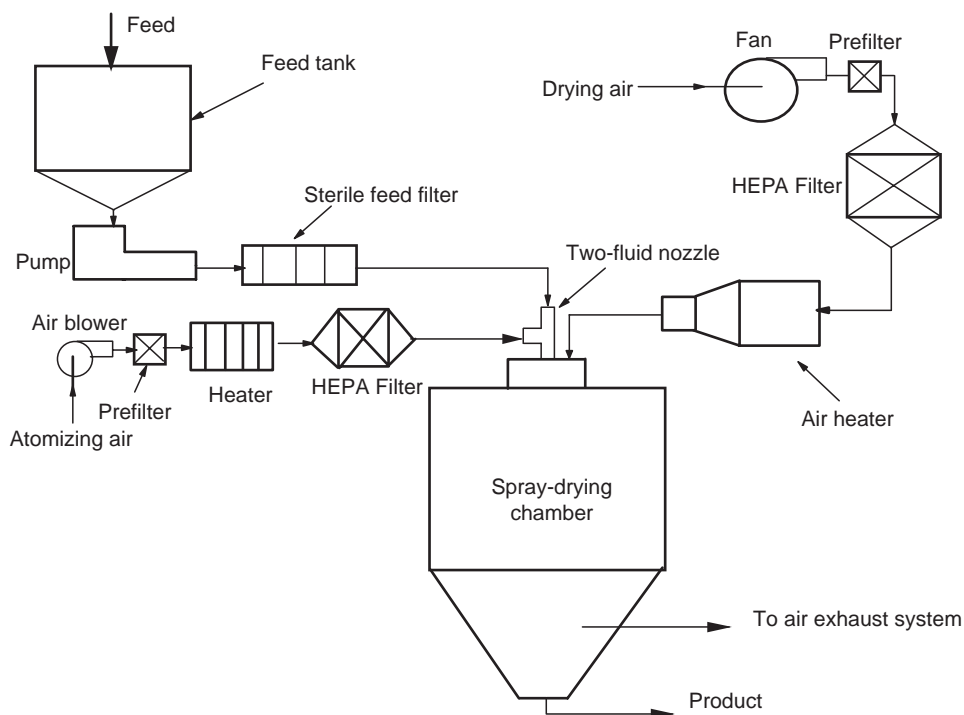


FIGURE 44.17 Aseptic-spray-drying system. (From Niro Inc. With permission.)

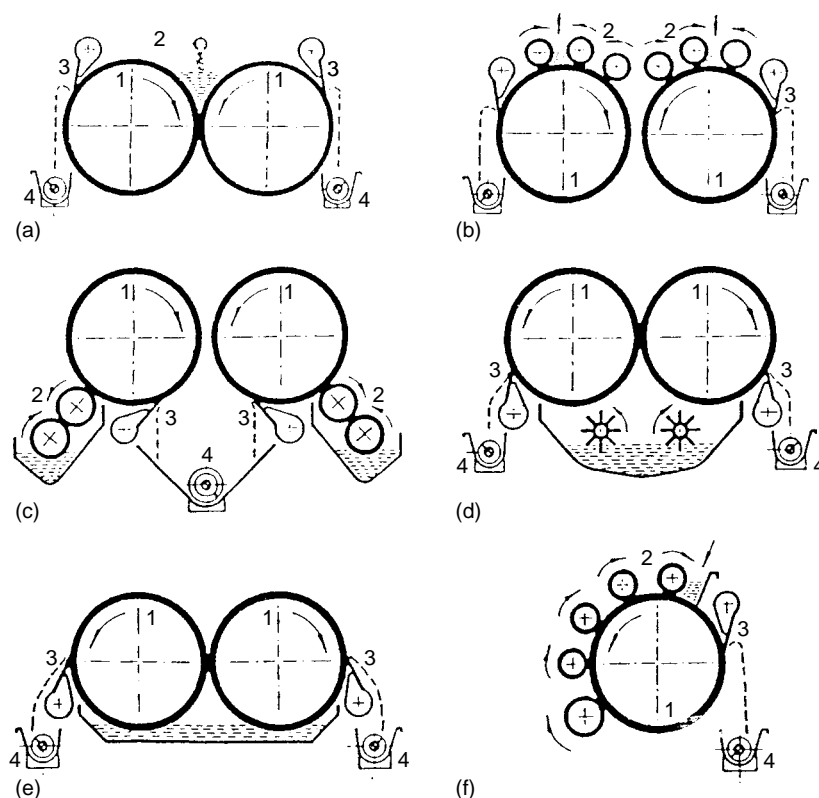


FIGURE 44.18 Standard feeding arrangements for drum dryers: (a) nip feed; (b) feed roll; (c) double-applicator roll; (d) splash feed; (e) dip feed; (f) multiple-applicator roll; (1) drums; (2) applicator rolls; (3) doctor blades; (4) conveyors. (From R. Simon & Sons.)

difficult to dry in other dryers. Drum dryers are divided into two broad types depending on the number of drums: single-drum dryers and double-drum dryers. For proper operation of a drum dryer, the material to be dried is either spread in a thin layer on the drum surface or the drum is placed in a trough filled with the nonsettling liquid slurry. The slurry or paste is fed onto the drums by means of various types of feeding arrangements. Prior experience on the specific product or pilot-plant tests is needed to establish whether a satisfactory film can be formed with a given design. A typical test procedure is outlined by Nonhebel and Moss [33]. Some of the standard feeding arrangements for industrial drum dryers are shown in Figure 44.18.

Double-drum dryers normally employ a nip-feed device with adjustable space between the drums that provides a means of controlling the film thickness [29]. Alternately, and in the case of single-drum types, a variety of feeding methods can be used to apply materials to the drum. The most usual is the simple dip or splash feed [29,37]. With this arrangement, good liquor circulation in the trough is desirable to avoid increasing the concentration of the feed by evaporation. The dip-feed system was the earliest design and

is still used when the liquid can be picked up from a shallow pan. An agitator prevents settling of any particles, and a spreader is sometimes used to produce a uniform coating on the drum. The splash-feed type is used for slurries materials such as calcium arsenate, lead arsenate, and iron oxide [37]. In both splash-feed and dip-feed operations, there should be sufficient overflow from the feed tray to prevent stagnation and deposition of solids at the corners of the tray [33]. For special applications, single-drum dryers use top roller feed in order to prevent possible uneven film formation. While the number of rolls is related to the particular application and the material being handled, in general this method of feeding is used for pasty materials such as starch. When the feed is very mobile, rotating devices such as spray feeders may be used.

44.6 CONCLUSION

This chapter provides a summary of various types of feeders commonly used in conjunction with different dryer types. Appropriate guidelines are also provided for selection. In many cases, a number of alternate feeders can be used for equal performance. Care must be taken to be sure that the full range of moisture

content is tested as moisture-induced stickiness can cause severe operational problems. Reference to the literature cited as well as to the vendors can be done for further information and operational experience.

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45 Dryer Emission Control Systems

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45.1 INTRODUCTION

Dryers are one of the major sources of atmospheric emissions in industrial operations as the majority of

industrial dryers operate in an open-cycle system. A drying installation may cause air pollution by the emission of dust, gases, and odors. Even plumes of clean water vapor are unacceptable in some areas.

The nature of the emissions is determined by the material being dried and the operating conditions. As no such attention is paid to environmental protection, these emissions are no longer permissible in many countries and it is logical that effective means of controlling pollution must be developed. New drying plants must be designed with emission control systems capable of meeting the most severe local requirements so that they can provide the necessary performance; existing plants can be modified and fitted with control systems in order to meet new regulations.

Generally, emission control systems used in the industrial dryers are concerned with the removal or collection of air contaminants for purposes of

1. Air pollution control
2. Safety or health-hazard elimination
3. Recovery of a valuable product or solvent
4. Powdered-product collection (e.g., spray drying)

The objective of this chapter is to briefly present the current technology for dryer emission control systems via flowcharts, tables, design equations, and practical description of various dryer emissions and control systems. Recent developments, trends, and regulations are also discussed. No attempt is made to give full design details or to include all dryer emission control equipment. Discussion will therefore be limited to some of the better-known, well-proven, and widely used devices. The reader is referred to the literature cited for more details and additional information on specific types of emission control systems mentioned in this chapter.

Following a general overview of the air pollution problems encountered in industrial drying plants and the general considerations for selection and design of control equipment, this chapter will provide information on (a) the basic principles of operation and equipment configurations of widely used emission control devices and their advantages and disadvantages, selection and design, and performance characteristics (b) dryer emission control systems in selected industrial processes. We do not address the Control of emissions from fossil fuel combustion used to provide the thermal energy in drying.

45.2 GENERAL CONSIDERATIONS

Dust can be a problem in any convective dryer in which the material is agitated or stirred during the drying process and where direct contact occurs between the material being dried the air and. Dryer types that can be major dust producers are direct-fired rotary dryers, flash dryers, fluid-bed dryers, and

spray dryers. In conveyor dryers, the solids remain stationary and low levels of entrainment are usually expected, whereas indirect dryers produce little or no dust. When an organic liquid is to be removed from a material, the emissions may include vapors, mists, odors, and smoke. In this case, recovery of the solvent is frequently desirable in order to lower costs, prevent a safety hazard, and eliminate air pollution.

More stringent demands governing environmental aspects of the plant operations have resulted in better compliance with antipollution regulations. Modern drying plants incorporate a variety of emission control systems, which constitute a substantial proportion of the total capital investment. Moreover, the strict environmental regulations have led to the use of alternative dryer types and layouts such as indirect dryers and close-cycle layouts instead of the traditional open-cycle convective drying plants.

For industrial dryers, air pollution control equipment may be classified into three groups: (a) equipment for controlling particulate matter (PM); (b) equipment for controlling gaseous emissions; and (c) equipment for controlling odor emissions. In selecting the optimum device for a particular job, it is necessary to consider many factors. In general, they can be grouped into three categories: (a) environmental, (b) engineering, and (c) economic [1,2].

Environmental factors include:

1. Equipment location
2. Available space
3. Ambient conditions
4. Availability of adequate utilities (e.g., power, water) and ancillary system facilities (e.g., waste treatment and disposal)
5. Maximum allowable emissions (air pollution regulations)
6. Aesthetic considerations (e.g., visible steam, water vapor plume)

Engineering factors include:

1. Contaminant characteristics (e.g., physical and chemical properties, concentration, particulate shape, and size distribution—in the case of particulates, chemical reactivity, corrosivity, abrasiveness, toxicity)
2. Gas stream characteristics (e.g., volume flow rate, temperature, pressure, humidity, composition, viscosity, density, reactivity, combustibility, corrosivity, toxicity)
3. Design and performance characteristics of the particulate control system (e.g., size and weight, fractional efficiency curves in the case of particulates, mass transfer in the case of

TABLE 45.1
Types of Equipment Applicable to Control Various
Classes of Dryer Emissions

Equipment Type	Emission Classification		
	Particulate	Gas	Odor
Cyclones	•		
Fabric filters	•		
Wet scrubbers	•		
Electrostatic precipitators	•		
Incineration	•	•	•
Adsorption		•	•
Absorption		•	•
Condensation		•	•
Wet scrubbing with chemical reaction			•
Biofiltration			•
Vapor neutralization			•
Air dilution			•

gases or vapors, pressure drop, reliability and dependability, power requirements, utility requirements, temperature limitations, maintenance requirements)

Economic factors include:

1. Capital cost (e.g., equipment, installation, engineering)
2. Operating cost (e.g., utilities, maintenance)
3. Expected equipment lifetime and salvage value

Table 45.1 lists the types of devices that can be used to control emissions from process dryers. The selection procedure is always dictated by experience, reviewing the advantages and disadvantages of each type of air pollution control equipment, and capability of the equipment to meet the technical requirements of the process and to achieve compliance with

the regulations at the lowest overall cost. In some drying installations, the cost of the equipment needed to comply with regulations can exceed that of the dryer itself.

Many types of particulate collection devices are available commercially (see Table 45.2). Each operates on a different principle for accomplishing removal of particulates from the gas stream. Four basic types are common in drying systems: (a) the drying vessel itself (in the case of vessel dryers), (b) cyclones, (c) bag filters, and (d) wet scrubbers. Electrostatic precipitators (ESPs) are not used widely in drying installations in spite of their low-pressure drop and high collection performance. The initial cost of purchase and construction is high. For this reason, the emphasis in [Section 45.3](#) will be on the three most widely used devices, e.g., cyclones, fabric filters, and wet scrubbers. For a concise discussion of various types of solid–gas separation equipment and guidelines for selection of dust collectors, the reader is referred to Hanly and Petchonka [3], Croom [4], Morgan [5], Constance [6], Amrein [7], Bielostradek [8], Morgan and Walters [9], and Kaff [10].

The removal of gaseous pollutants from dryer exhaust may be accomplished by several possible processes. Among these are absorption, adsorption, condensation, and incineration [11–16]. The choice of a given process is usually determined by physical and chemical characteristics of the dried product and by economic and environmental considerations. [Table 45.3](#) summarizes some of the basic characteristics of the gaseous emission control equipment.

Odor removal may be obtained by physical or chemical means (Table 45.1). It is worth noting that moisture-laden or organic vapor emissions should first be passed through a condenser that is capable of removing a high percentage of the water vapor or condensable volatile organic compounds (VOCs). This minimizes the odorous gas stream to be treated or allows recovery of a costly solvent.

TABLE 45.2
Summary of Particulate Control Equipment Characteristics

Type of Equipment	Primary Collecting Forces	Collecting Surface	Optimum Particle Size, microns	Efficiency, %	Collected Pollutant
Cyclones	Centrifugal	Cylindrical	10–100 μm	50–90	Dry dust
Bag filters	Direct interception	Particulate layer	Down to 1 μm or less	>99	Dry dust
Wet scrubbers	Inertial, diffusional, direct interception	Spherical or irregular	Down to 5 μm or less	80–99	Liquid
Electrostatic precipitators	Electrostatic	Plane or cylindrical	Down to 1 μm or less	95–99	Dry or wet dust

TABLE 45.3
Summary of Gaseous Pollutant Control Equipment Characteristics

Type of Equipment	Efficiency, %	Remarks
Incinerators	90–95% at low gaseous pollutant concentrations (<100 ppmv) 95–99% at higher concentrations (>100 ppmv)	Produce products of incomplete combustion or otherwise undesirable by-product that may require additional controls
Adsorption	90% at gaseous pollutant concentrations in the order of few hundreds ppmv 95% at higher concentrations (>1000 ppmv)	Less favored if the mixture of recovered organics cannot be returned to the process with maximum additional treatment
Absorption	90–95% at low inlet concentrations 95–99% at high inlet concentrations	Less favored if a liquid blowdown stream cannot be accommodated
Condensation	80+% at high inlet concentrations (> few thousands ppmv)	Cannot meet high efficiency without the use of very low temperature or high pressure

Source: From A. Buonicore, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992. With permission.

45.3 EMISSION CONTROL EQUIPMENT

45.3.1 CYCLONES

45.3.1.1 Principle of Operation and Equipment Configurations

The cyclone is a mechanically simple, reliable device for the separation of PM from an air stream by the action of centrifugal forces. The centrifugal forces, resulting from the tangential velocity given to the dust-laden gas at the top of the cyclone, eject the particles in a circular, vortex motion toward the cyclone wall. These particles, because of their inertia forces, attempt to move toward the outside wall, from which they are led to a receiver or hopper. The clarified gas exits as an axial inner vortex through the top by way of the gas exit duct.

Figure 45.1 shows a typical reverse-flow cyclone in which the necessary elements consist of a gas inlet that produces the vortex; an axial outlet for cleaned gas, and a dust-discharge opening. There are a number of different arrangements and modifications that offer variations in performance and overcome some of the limitations of the conventional reverse-flow cyclone. These modifications can be summarized according to the following classifications [18]:

1. Design mode, which can be (a) a scroll or swirl vane inlet instead of the tangential duct, (b) straight-through flow, without the conical portion, (c) insertion of various kinds of internal vanes, or (d) mechanical centrifugal separators

2. Operation mechanism, which can be (a) two or more cyclones in parallel, (b) two or more cyclones in series, (c) recycling portion of the gas outlet stream back to the feed, (d) the use of secondary air flow, (e) skimming off portion of uncollected dust in the emission stream, (f) base purge bleeding, or (g) use of electrostatic cyclones

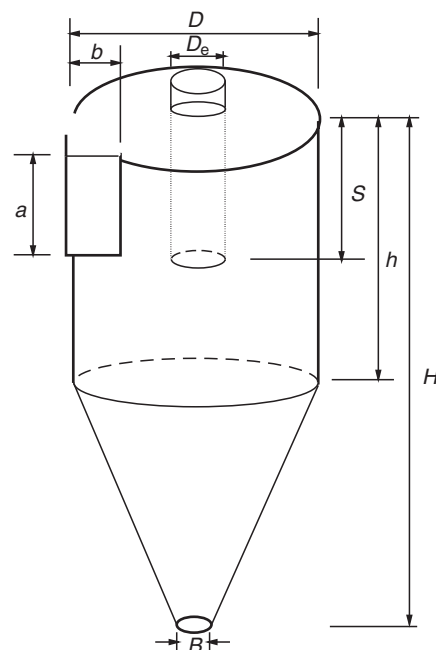


FIGURE 45.1 Shape and dimensions of reverse-flow cyclone.

45.3.1.2 Advantages and Disadvantages [1,19,20]

Advantages of cyclones are:

1. Simple equipment with no moving parts and few maintenance problems
2. Low operating and capital costs
3. Temperature and pressure limitations imposed only by the materials of construction; special heat-resisting alloys or refractory linings used for high temperatures
4. Design to handle a wider range of chemical and physical operation conditions than most other types of collection equipment can handle
5. Relatively low operating pressure drop
6. Dry collection and disposal
7. Relatively small space requirements

Disadvantages of cyclones are:

1. Inefficiency in collecting particles smaller than 5 to 10 μm in diameter
2. Large gas flow rates require multiple-unit or other designs
3. Not suitable for collecting sticky materials that would tend to cling in the interior wall, or other particles that might tend to agglomerate and clog the exit duct
4. Normally require secondary collector to meet emission control regulations

45.3.1.3 Selection and Design

Cyclone design consists of selecting a configuration, and then determining the size, total efficiency, pressure drop, power requirement, and cost of each cyclone to be used. The main parameters are design configuration, inlet velocities, grade and overall efficiency, and pressure drop.

Design Configuration—The overall size of the cyclone is characterized by the cyclone body diameter D , and the seven geometric ratios: $K_a = \frac{a}{D}$, $K_b = \frac{b}{D}$, $K_e = \frac{D_c}{D}$, $K_s = \frac{S}{D}$, $K_h = \frac{h}{D}$, $K_B = \frac{B}{D}$, $K_H = \frac{H}{D}$. Licht [18], Koch and Licht [21], and Leith [22] have summarized a number of cyclone configurations that have been proposed and studied sufficiently to be accepted as standard designs. Table 45.4 shows some of the configurations that can be selected based on whether high-efficiency or general purpose use is desired.

Inlet Velocities—Inlet velocities are normally in the range of 10 to 30 m/s. Cyclone efficiency increases with inlet velocity up to a certain value at which saltation occurs and particles bounce along the cyclone wall and cannot be separated effectively. The saltation velocity is given, empirically, by [22]

$$u_s = 2400 \frac{\mu \rho_p}{\rho_g^2} D^{0.2} \frac{(b/D)^{1/2}}{1 - b/D}$$

Grade and Overall Efficiency—Collection efficiency is a complex exponential function of inlet velocity; air density and viscosity; particle mass, size, shape, and roughness; and cyclone design configuration. A number of attempts have been made to predict the grade efficiency of conventional reverse-flow cyclones. Several modeling approaches have been reviewed by Licht [18] and the results of these models are summarized in Table 45.5.

The overall efficiency η is defined as the mass of particles removed divided by the mass entering the cyclone, per unit time. It may be calculated by integrating the grade efficiency over the particle size distribution:

$$\eta = \int_0^1 \eta_i df$$

TABLE 45.4
Cyclone Design Configurations (Reverse-Flow Type)

Source	Recommended Duty	D	K_a	K_b	K_e	K_s	K_h	K_H	K_B
Stairmand [18,22,23]	High efficiency	1.0	0.50	0.20	0.50	0.50	1.50	4.0	0.375
Swift [18,22,24]	High efficiency	1.0	0.44	0.21	0.40	0.50	1.40	3.9	0.40
Lapple [18,22,25]	General purpose	1.0	0.50	0.25	0.50	0.625	2.00	4.0	0.25
Swift [18,22,24]	General purpose	1.0	0.50	0.25	0.50	0.60	1.75	3.75	0.40
Dirgo and Leith [18,22,26]	Experimental	1.0	0.50	0.30	0.333	0.558	3.50	6.0	0.375
Peterson and Whitby [18,27]	Experimental	1.0	0.583	0.208	0.50	0.583	1.333	3.17	0.50
Stern et al. [22,28]	Consensus	1.0	0.45	0.20	0.50	0.625	0.75	2.0	
Stairmand [18,22,23]	High throughput (scroll-type entry)	1.0	0.75	0.375	0.75	0.875	1.50	4.0	0.375
Swift [18,22,24]	High throughput (scroll-type entry)	1.0	0.80	0.35	0.75	0.85	1.70	3.7	0.4

TABLE 45.5
Models for Calculating Grade Efficiency

1. Ranz [18,31]

$$\eta_i = 2\pi N_e S t_{kc} D / b$$
 where

$$S t_{kc} = \frac{\rho_p d_p^2 u_T}{18\mu D}$$

$$N_e = [h + (H - h)/2]/a$$
2. Leith and Licht [18,32]

$$\eta_i = 1 - \exp\left(-2(S t'_{kc} K)^{\frac{1}{2n+2}}\right)$$
 where

$$S t'_{kc} = (n + 1)K_a K_b S t_{kc}$$

$$n = 1 - [1 - 0.67D^{0.14}](T/283)^{0.3}, T \text{ in K}$$

$$K = 8K_c/K_a^2 K_b^2$$

$$K_c = (V_s + V_{nl}/2)/D^3$$

$$V_s = \pi(S - a/2)(D^2 - D_e^2)/4$$

$$V_{nl} = \frac{\pi D^2}{4}(h - S) + \frac{\pi(\ell + S - h)D^2}{4}\left(1 + \frac{d}{D} + \frac{d^2}{D^2}\right) - \frac{\pi D_e^2 \ell}{4}$$

$$d = D - (D - B)(S + l - h)/(H - h)$$

$$\ell = 2.3D_e\left(\frac{D^2}{ab}\right)^{\frac{1}{3}}$$
3. Dietz [18,33]

$$\eta_i = 1 - [K_o - (K_1^2 + K_2)^{0.5}] \exp\left(-\left(\frac{\pi(2K_s - K_a)S t_{kc}}{K_a K_b}\right)\right)$$
 where:

$$K_o = \frac{1}{2}\left[1 + K_e^{2n}\left(1 + \frac{K_a K_b}{2\pi K_\ell S t_{kc}}\right)\right]$$

$$K_1 = \frac{1}{2}\left[1 - K_e^{2n}\left(1 + \frac{K_a K_b}{2\pi K_\ell S t_{kc}}\right)\right]$$

$$K_2 = (K_e)^{2n}$$

$$K_\ell = \frac{\ell}{D}$$
4. Ranz [18,31]

$$\eta_i = 1 - \frac{1}{1 + \frac{(2\pi N_e)}{K_b} S t_{kc}}$$

where f is the mass fraction of solids less than that size in the feed stream.

Pressure Drop—A knowledge of cyclone pressure drop is required so that power consumption can be estimated in order to size and select the exhaust air fan. Cyclone pressure drop is expressed as the number of inlet velocity heads N_H as follows:

$$\Delta P = \rho_g u_i^2 N_H / 2$$

Empirical, theoretical, and semitheoretical expressions for determining N_H have been presented in the literature. Reviews of these expressions are given by Leith [22] and Strauss [29]. Several well-known correlations are listed in Table 45.6.

It is worth noting that increasing grain loading has a pronounced effect on cyclone pressure drop. Briggs quantified this effect in terms of a correction factor as [30]

TABLE 45.6
Equations for Predicting the Number of Inlet Velocity Heads (N_H) for Cyclones

Shepherd and Lapple [22,34]

$$N_H = 16 \frac{ab}{D_e^2}$$

Stairmand [22,35]

$$N_H = 1 + 2\phi^2 \left(\frac{2(D - b)}{D_e} - 1 \right) + 2 \left(\frac{4ab}{4\pi D_e^2} \right)^2$$

$$\phi = \frac{-\left(\frac{D_e}{2(D-b)}\right)^2 + \left(\frac{D_e}{2(D-b)} + \frac{4G^*A}{ab}\right)^{\frac{1}{2}}}{(2G^*A/ab)}$$

$$A = \frac{\pi}{4}(D^2 - D_e^2) + \pi Dh + \pi D_e S$$

$$+ \frac{\pi}{2}(D + B) \times \left((H - h)^2 + \left(\frac{D - B}{2} \right)^2 \right)^{\frac{1}{2}}$$

$$G^* = 0.005$$

$$\Delta P_{\text{dusty}} = \frac{\Delta P_{\text{clean}}}{1 + 0.0086(c_i)^{\frac{1}{2}}}$$

where c_i is in g/m^3 .

45.3.1.4 Performance Characteristics and Operational Problems

A properly designed cyclone can effectively process dusts in very high concentrations and high loadings. Higher grain loadings simultaneously bring about an increase in pressure drop for a given volumetric gas flow rate. Cyclones are commercially available in sizes to process 50 to 50,000 m^3/h [22]. Cyclone performance is commonly described by its collection efficiency. Typical ranges of collection efficiencies as functions of particle size and cyclone type are shown in Table 45.7.

Cyclone collectors are normally selected for pressure drops in the following ranges [1]:

Low-efficiency cyclones	500 to 1000 N/m^2
Medium-efficiency cyclones	1000 to 1500 N/m^2
High-efficiency cyclones	2000 to 2500 N/m^2

Cyclones, like other types of process equipment, have a number of potential operational problems and certain precautions need to be taken to obtain the desired performance from them [20,36–39]. These problems are:

1. Reduced flow rates.
2. Increased loadings in submicrometer particle sizes.

TABLE 45.7
Cyclone Collection Efficiency

Particle Size (μm)	Efficiency (% by wt)	
	Conventional Cyclone	High-Efficiency Cyclone
<5	—	50–80
5–20	50–80	80–90
15–50	80–95	90–99
>40	95–99	95–99

Source: From L. Theodore and A. Buonicore, *Industrial Air Pollution Control Equipment for Particulates*, CRC Press, Cleveland Thiessen, DB, 1976. With permission.

3. Insufficient capacity for collected material.
4. Irregular flow patterns resulting from erosion damage.
5. Cyclone overloading (i.e., where a multiunit installation is involved, partial blockage in one of the inlet ducts may be causing uneven distribution of gas flow).
6. Partial or total blockage of the cyclone base.
7. Precautions should be taken to avoid reaching the dew point since moisture condensation on the internal surfaces of the cyclone will start corrosion and initiate solid buildup. Condensation can be reduced by preheating the system before the feed of material is started, gradual shut down until the system has been completely drained of dust and water vapor, proper insulation, and heating the system.
8. Fire and explosion hazards may occur due to buildup of solids on the internal surfaces of ducts and cyclone.
9. Precautions are necessary in the prevention of leakage of air into the cyclone and leakage of gas plus dust out of the cyclone; both will reduce the collection efficiency.

For intermittent withdrawal of dust, an airtight receiver is used; for continuous operation a star valve, screw conveyor, or special types of locking gate valves are used.

45.3.2 FABRIC FILTERS (BAGHOUSES)

45.3.2.1 Principle of Operation and Equipment Configurations

Fabric filter systems (e.g., baghouses) are among the oldest and most widely applied particulate emission control devices. Dust filtration in baghouses is

principally accomplished by passing the dust-laden gas through a fabric filter. As the gas passes through the fabric, the dust is collected on the upstream side by the filtering action of the fabric, building up a filter cake that also acts as a filter medium. Because of this accumulated load, the pressure drop increases and the gas velocity decreases to a prescribed lower limit. At this point, the operation must be stopped and the dust collected on the fabrics must be removed and collected in a hopper.

A combination of mechanisms results in the collection of particles from a gas stream by means of a fabric filter. These mechanisms include [40,41]:

1. Interception
2. Inertial impaction
3. Diffusion
4. Electrical precipitation
5. Thermal precipitation
6. Gravitational settling

Fabric filtration equipment may be characterized and classified according to the following criteria [2]:

1. Equipment geometry
2. Mode of operation
3. Capacity
4. Type of fabric material
5. Cleaning method

Industrial fabric filters usually consist of a number of fabric filter elements. These elements are constructed as tubes, envelopes, a pocket slipped over a wire frame, or cartridges. The most commonly employed fabric filter element is tubular shaped. This system is frequently referred as baghouses, as the fabric is usually configured in cylindrical bags. In a baghouse collector, a number of tubular fabric filters are suspended vertically and operate in parallel. If the required number of bags is very large, they may be grouped into compartments of several hundred bags each. Bags are typically 0.1 to 0.3 m in diameter and 2 to 10 m in length.

In terms of mode of operation, fabric filter equipment follows one of the three systems [2]: bottom-feed units; top-feed units; and exterior filtration units. The capacity provides another way to categorize and view fabric filters. Generally, the groupings are small volumes ($<300 \text{ m}^3/\text{min}$), medium volumes (300 to $3000 \text{ m}^3/\text{min}$), and large volumes ($>3000 \text{ m}^3/\text{min}$) [42].

The choice of filter medium and fabric construction is important to baghouse performance. The fabric filters are constructed from both natural and synthetic materials, and the selection is usually made on the basis of the following factors [2]:

1. Maximum temperature of the gas stream
2. Physical and chemical characteristics of the particles to be collected
3. Chemical composition of both the carrier gas and the collected dust
4. Cost

Table 45.8 is a guide that can be used in the selection of the fabric filters based on the above-mentioned factors.

The temperature of exhausts discharged from high-temperature drying operations must be adjusted to conditions that can be tolerated by the fabric filter. This requires that the exhaust be properly precooled prior to filtration. Lowering the temperature will protect the filter media and ensure economic filter life. Furthermore, lowering the temperature will reduce the flow volume, which, in turn, decreases the fabric filter area requirement. Three methods (Figure 45.2) are generally considered when precooling is required:

1. Dilution by mixing with outside air
2. Radiation and convection cooling using heat exchangers or waste-heat boilers
3. Spray cooling with water

Baghouses can best be characterized and categorized according to the method used to remove collected materials from the bags. Three common methods have been employed in baghouses (Figure 45.3): shake cleaning; reverse-flow cleaning; and pulse-jet cleaning. Table 45.9 gives a summary of the general characteristics associated with the three cleaning types.

For dispersion dryers such as flash, fluid-bed, spray, and some others, the most practical bag filter collector is the jet pulse type [3]. It causes the least fluctuation of air flow and so allows these dryers perform more efficiently. Collectors that periodically shut down one of the several compartments for cleaning are unable to hold a steady pressure drop. But

TABLE 45.8
Fiber Selection Guide

Fabric	Generic Name	Maximum Temperature, °C		Acid Resistance	Alkali Resistance	Abrasion Resistance	Tensile Strength
		Continuous	Surges				
Cotton	Natural fiber cellulose	80	110	Poor	Good	Good	Good
Polypropylene	Polyolefin	90	120	Excellent	Excellent	V. good	Excellent
Dacron	Polyester	135	160	Good	Good	Excellent	Excellent
Nomex	Nylon aromatic	200	260	Fair	Excellent	Excellent	V. good
Teflon	Fluorocarbon	230	260	Excellent	Excellent	Fair	Fair
Fiberglass	Glass	260	315	Good	Fair	Fair	Excellent
Wool	Natural fiber protein	90	120	Fair	Poor	Fair	Poor
Nylon	Nylon polyamide	90	120	Fair	Excellent	Excellent	Excellent
Dynel	Modacrylic	80	115	Good	Good	Good	—
Orlon	Acrylic	115	135	V. good	Fair	Good	Fair
Polyethylene	Polyolefin	90	—	V. good	Excellent	Good	—
Goretex	Polytetrafluoroethylene	230	260	Excellent	Excellent	Poor	Fair
Microtain	Acrylic	125	140	V. good	Fair	Fair	Fair
PBI	Polybenzimidazole	540	650	Good	Fair	Excellent	Excellent
Ryton	Polyphenylene sulfide	170	—	Good	Good	Good	Fair
Dralon-T	Homopolymer acrylic	120	—	Good	Excellent	Good	Good
Bekipor	Stainless steel 316L	450	540	Good	Good	Excellent	Excellent

Source: From L. Theodore and A. Buonicore, *Industrial Air Pollution Control Equipment for Particulates*, CRC Press, Cleveland Thiessen, DB, 1976; W. Licht, *Air Pollution Control Engineering*, Marcel Dekker, New York, 1988; J. Turner and J. McKenna, in *Handbook of Air Pollution Technology*, S. Calvert and H. Englund (Eds.), John Wiley & Sons, New York, 1984; J. McKenna and G. Greiner, in *Air Pollution Control Equipment*, L. Theodore and A. Buonicore (Eds.), Prentice-Hall, Englewood Cliffs, NJ, 1982; J. McKenna and Furlong, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992; K. Iinoya and R. Dennis, in *Filtration Principles and Practice*, 2nd ed., M. Matteson and C. Orr, Jr. (Eds.), Marcel Dekker, New York, 1987; K. Iinoya and C. Orr, Jr., in *Air Pollution*, 2nd ed., Vol. 3, A. Stern (Ed.), Academic Press, New York, 1968; and K. Iinoya and C. Orr, Jr., in *Air Pollution*, 3rd ed., Vol. 4, A. Stern (Ed.), Academic Press, New York, 1977. With permission.

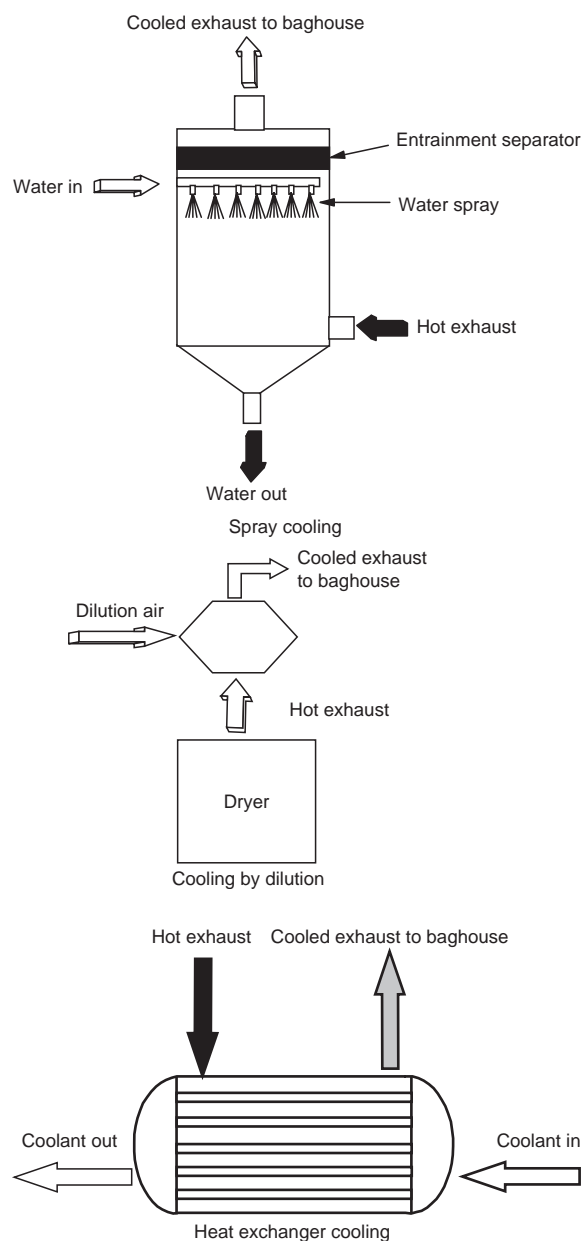


FIGURE 45.2 Dryer exhaust cooling methods.

these collectors have other advantages and are in common use in other dryer types as well.

45.3.2.2 Advantages and Disadvantages [1,2,37,38,39,40,42,47–49]

Advantages of fabric filters are:

1. Reliable and highly efficient on both coarse and fine particulates.
2. Capable of meeting particulate emission standards in most countries.
3. Efficiency and pressure drop are relatively insensitive to gas stream fluctuations as a result

of large changes in inlet dust loadings for continuously cleaned filters.

4. Filter outlet air may be recirculated within the plant in many cases, which leads to energy conservation.
5. No problems with liquid waste disposal, water pollution, or liquid freezing.
6. Corrosion and rusting of components are usually not problems.
7. There is no hazard of high voltage, thus simplifying maintenance and repair and permitting collection of flammable dusts.
8. Use of selected fibrous or granular filter aids permits the high-efficiency collection of submicron smokes and gaseous contaminants.
9. Filter collectors are available in a large number of configurations, resulting in a range of dimensions and inlet and outlet flange locations to suit installation requirements.
10. Relatively simple operation.

Disadvantages of fabric filters are:

1. Temperatures greater than 250 to 300°C require special refractory mineral or metallic fabrics.
2. Fabric life may be shortened at elevated temperatures and in the presence of acid or alkali particulate or gas constituents.
3. Certain dusts may require fabric treatments to reduce dust seeping or, in other cases, assist in the removal of the collected dust.
4. Concentrations of some dusts in the collector ($\sim 50 \text{ g/m}^3$) may represent a fire or explosion hazard if a spark of flame is admitted by accident. Fabrics can burn if readily oxidizable dust is being collected.
5. Relatively high maintenance requirements (e.g., bag replacement).
6. Hygroscopic materials, condensation of moisture, or tarry adhesive components may cause crusty caking or plugging of the fabric or require special additives.
7. Replacement of fabric may require respiratory protection of maintenance personnel.
8. Not recommended for handling air of high moisture loading, hygroscopic or deliquescent airborne particles, products requiring high standards of hygiene in handling, particles/agglomerates in stringy (fibrous) form, and in multiproduct drying where product contamination must be prevented.

45.3.2.3 Selection and Design

The sizing and selection of fabric filters (Figure 45.3) are based on [18,50]: (a) past experience; (b) actual

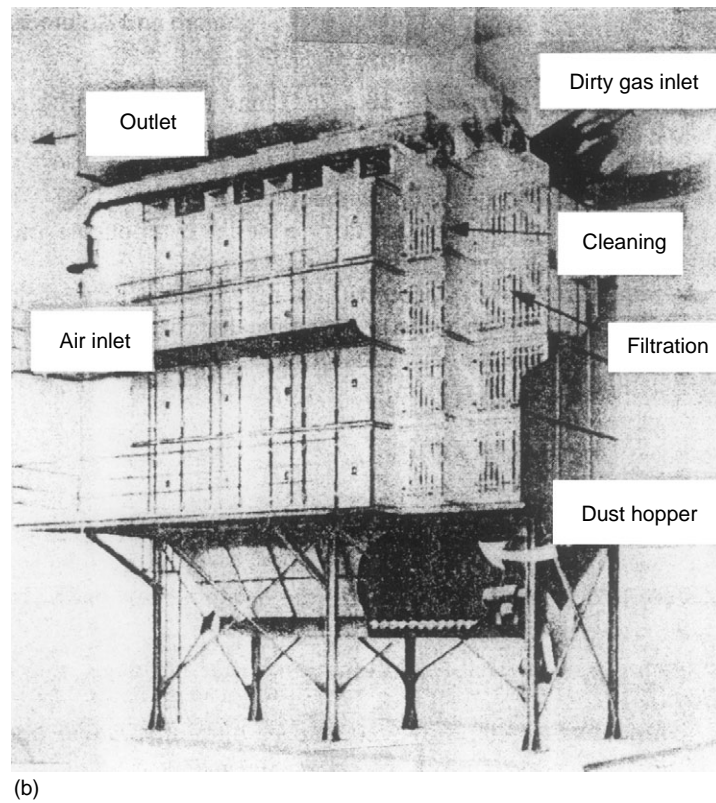
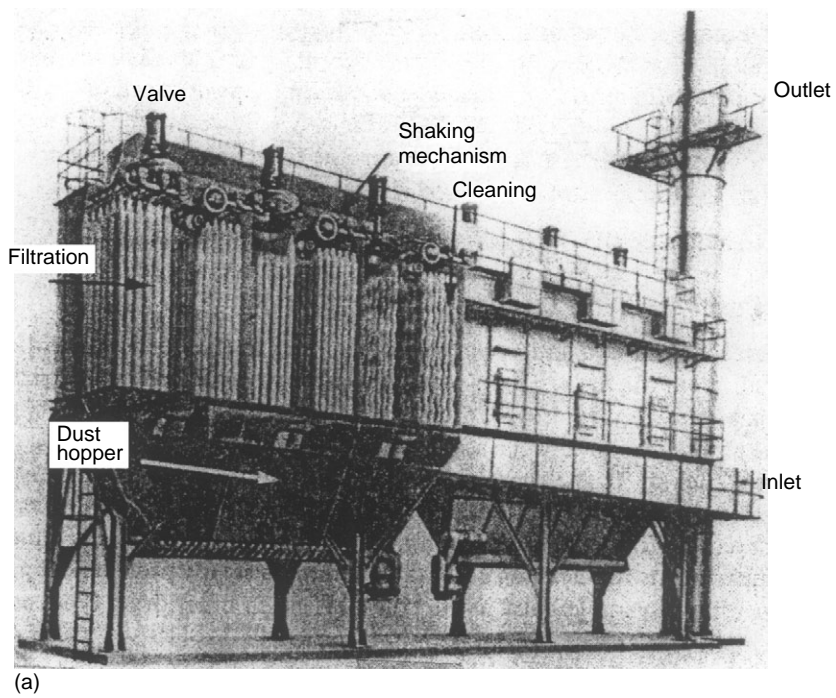


FIGURE 45.3 Fabric filters: (a) large, multicompartiment fabric filter cleaned by mechanical shaking. (Courtesy of Wheelabrator-Shint.) (b) Envelope-type fabric filter cleaned by reverse flow. (Courtesy of Shinto Dust Collector.)

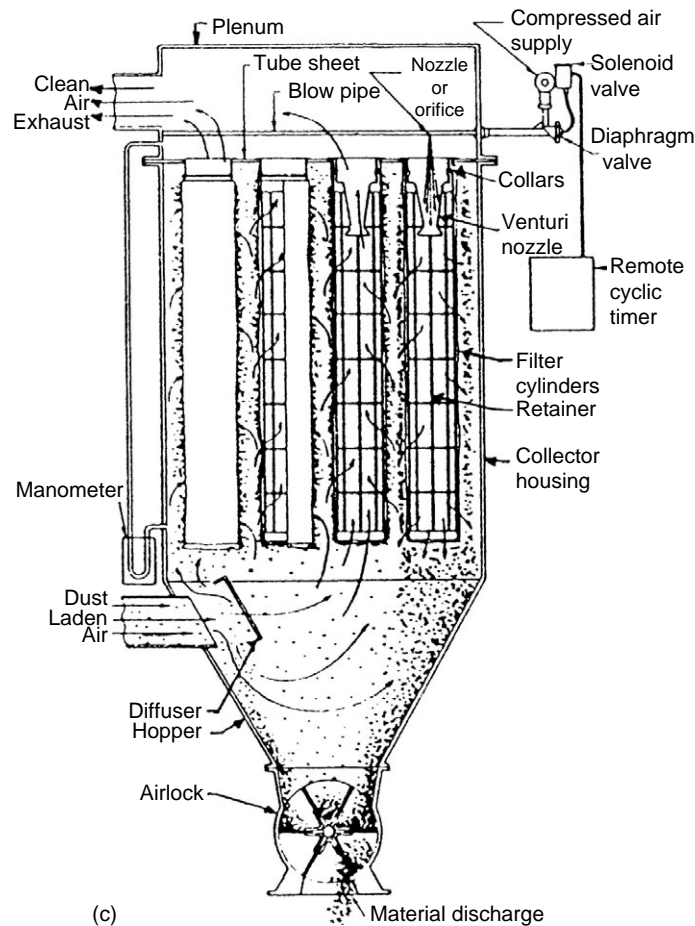


FIGURE 45.3 (continued) (c) Pulse-jet fabric filter. (Courtesy of Mikropul Corp.)

tests that use specific fabrics at specific dust-to-air loadings; and (c) empiricism and experience blended with theory. The steps to be taken in the design or selection of a baghouse system are listed in the order recommended by Turner and McKenna [41]:

1. Specify dust properties: type, density, size distribution, concentration range, stickiness, and other characteristics
2. Specify gas properties: temperature range, flow rate range, humidity range, and chemical constituents
3. Specify exhaust requirements: concentration, allowable pressure drop, and opacity
4. Design the collector:
 - (a) Choose filter medium (Table 45.8)
 - (b) Choose cleaning method (Table 45.9)
 - (c) Estimate gas-to-cloth ratio
 - (d) Calculate penetration
 - (e) Calculate required cloth area and number of compartments
 - (f) Specify fan and duct requirements

Gas-to-Cloth Ratio—One of the major factors in the design and operation of a fabric filter is the gas-to-cloth ratio. This parameter is the apparent velocity of the gas approaching the cloth. It can be calculated by dividing the total gas volume rate reported as ft^3/min or m^3/min by the total area in ft^2 or m^2 of cloth.

The selection of the gas-to-cloth ratio is very much an art rather than science. Factors influencing the gas-to-cloth ratio selection include cleaning method, filter media, dust size, dust density, dust loading, and other factors [42]. The gas-to-cloth ratio may be as low as $1 (\text{ft}^3/\text{min})/\text{ft}^2$ up to as high as $30 (\text{ft}^3/\text{min})/\text{ft}^2$ [40]. In general, a lower gas-to-cloth ratio will be used for gases containing smaller particles or particles that may otherwise be difficult to collect. A higher gas-to-cloth ratio will be used for gas streams containing larger particles or those particles that are easier to collect. Examples of typical gas-to-cloth ratios are given in the literature [2,18,4–42]. The usual design factor for drying is a gas-to-cloth ratio of 4:1, i.e., $4 (\text{ft}^3/\text{min})/\text{ft}^2$ of cloth area ($1.2 (\text{m}^3/\text{min})/\text{m}^2$ of cloth area) [50].

TABLE 45.9
Comparison of Fabric Filters Cleaning Methods

Cleaning Method	Fabric Type	Frequency	Motion	Mode	Duration	Bag Diameters	Gas Velocity (m/s)	Flow Direction	Dust Accumulation
Shaking	Woven	Several cycles/seconds; adjustable	Simple harmonic or sinusoidal	Off stream	10–100 cycles; 30 s to few minutes	13–30 cm	0.01–0.03	Outward	Inside the bag
Reverse-air	Woven	Clean one compartment at a time, sequencing one compartment after another; can be continuous or initiated by a maximum pressure drop switch	Gentle collapse of bag upon deflation; slowly repressurize a compartment after completion of a back-flush	Off stream	10–30 s or 1–2 min including valve opening and closing and dust settling periods	20–30 cm; length: 55–75 cm	0.01–0.03	Outward	Inside the bag
Pulse-jet	Felted	Usually a row of bags at a time, sequence one row after another; can sequence such that no adjacent rows clean one after another	Shock wave passes down bag; bag distends from cage momentarily	On stream	Compressed air (690 kPa) pulse duration 0.1 s	13–15 cm	0.02–0.06	Inward	Outside the bag

Source: From J. McKenna and G. Greiner, in *Air Pollution Control Equipment*, L. Theodore and A. Buonicore (Eds.), Prentice-Hall, Englewood Cliffs, NJ, 1982; J. McKenna and Furlong, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992; C. Billings and J. Wilder, *Handbook of Fabric Filter Technology*, Contract No. CPA-22-69-38, Vol. 1, GCA Corporation, December, 1970; and C. van't Land, *Industrial Drying Equipment*, Marcel Dekker, New York, 1991. With permission.

Pressure Drop—Another factor of major importance in a fabric filter is the pressure drop caused by the combined resistances of the fabric and the accumulated dust layer that builds up on the surface of the fabric. The pressure drop for an operating fabric filter with dust cake accumulated on the fabric can be estimated using the following equation [2,18,41]:

$$\Delta P = S_E + k_2 c_i V^2 t$$

where

ΔP = pressure drop, N/m²

S_E = effective residual drag of the fabric, N s/m³

k_2 = specific resistance coefficient of the cake
N s/g.m

c_i = concentration of the dust in the inlet gas
stream g/m³

t = filtration time, s

V = velocity, m/s

S_E is a measure of drag for the cleaned fabric and k_2 , the specific resistance coefficient, is primarily a characteristic of the dust, varies for different dusts, and is a measure of how rapidly pressure drop will build up in the system. Values of S_E and k_2 should be obtained experimentally from a plot of ΔP vs. t . The slope of the plot is proportional to k_2 while extrapolation to $t = 0$ gives an intercept from which S_E may be calculated.

Typical values of k_2 reported by various investigators for recorded case studies can be found in the literature cited. If no experimental values of S_E are available, a default value of 350 N min/m³ at 25°C for a residual fabric loading of 50 g/m² has been suggested [41].

The total pressure drop in multicompartment baghouses operated in parallel, in terms of drag, is analogous to a set of electrical resistances in parallel [43]:

$$\frac{N}{S_T} = \frac{1}{S_1} + \frac{1}{S_2} + \cdots + \frac{1}{S_N}$$

where

$$S_T = \text{total baghouse drag} = \frac{\Delta P}{V_T}$$

$$V = \text{total gas volume} = V_1 + V_2 + \cdots + V_N$$

$$N = \text{number of compartments.}$$

If one knows the drag in each compartment at a given time, the instantaneous drag (and, therefore, pressure drop) could be easily calculated for the entire system.

Penetration—Penetration (1-efficiency) is used as a measure of performance and may be measured or

calculated for specific particle size or size ranges. A series of empirical equations developed by Dennis and Klemm may be used for predicting the outlet concentration and penetration through a filter [18,41].

$$c_o = [P_{ns} + (0.1 - P_{ns})e^{-aw}]c_i + C_R$$

$$P_{ns} = 1.5 \times 10^{-7} \exp[12.7(1 - e^{1.03V})]$$

$$a = 3.6 \times 10^{-3} V^{-4} + 0.094$$

where

c_o = outlet dust concentration, g/m³

P_{ns} = dimensionless constant

V = local face velocity, m/min

C_R = residual outlet concentration due to
seepage = 5 mg/m³

w = dust loading on fabric, g/m²

These equations were developed for woven glass fabrics and fly ash. Their use to other systems would require experimentally determined values for P_{ns} , a , and C_R .

45.3.2.4 Performance Characteristics and Problems

The major performance characteristics of primary concern for fabric filters are collection efficiency, pressure drop, and lifetime. These characteristics depend upon (a) gas-handling capacity, (b) the type of product handled, (c) bag fabric, and (d) bag cleaning method.

Collection efficiency is probably the most important factor in the baghouse performance; it is usually expressed in terms of the mass percentage of dust retained. Filter collection efficiencies range from the minimum levels with new or freshly cleaned filters to the maximum values attainable with dust-laden filters immediately before cleaning. The weighted average efficiency is the most important parameter; it normally is 99% or more. Generally, the highest efficiencies and the lowest pressure losses are associated with low-velocity filters.

In spite of the wide ranges of the baghouse operation and performance parameters, the following typical values can be used [44]:

Collection efficiency	99 to 99.9%
Clean fabric pressure drop	60 to 200 Pa
Operating baghouse pressure drop	500 to 2500 Pa
Inlet dust concentrations	0.1 to 20 g/m ³
Gas-to-cloth ratio	0.5 to 5 (m ³ /min)/m ² of cloth area

Average fabric dust holdings	200 to 2000 g/m ²
Specific resistance coefficient (k_2)	0.5 to 50 N min/g · m
Bag life	1 to 4 y

The following key parameters should be considered during baghouse operation [41]:

1. *Gas volume*—A too high gas-to-cloth ratio can blind the bags; too low values can cause dust dropout in the ducts.
2. *Temperature*—Too high can destroy the bags and/or gasketing; too low might cause dew point excursions.
3. *Dust load*—Too high may exceed the unit's capacity to convey the dust from the baghouse; too low may cause extended emissions after each cleaning cycle.
4. *Particle size*—Too fine can cause blinding of the bags or excessive emissions at the design gas-to-cloth ratio.

The key to baghouse maintenance is the frequent and routine inspection of the filter bags. All bag units have a finite lifetime. Typical causes of bag failure are [42]:

1. High gas-to-cloth abrasion
2. Metal-to-metal abrasion
3. Bag-to-bag abrasion
4. Inlet velocity abrasion
5. Chemical attack
6. Upset conditions (e.g., temperature)
7. Accidents

Other major operation problems in baghouse collection system can be identified as follows [38,39,41,44,46,48,51]:

1. Abnormal high-pressure drop across the collector due to ineffective cleaning, wetting of bags, reentrainment of dust, or change in inlet loading or particle distribution.
2. If the dust is sticky or hygroscopic, it can lead to inadequate cleaning, eventually causing blinding of the gas.
3. Lower than normal dust discharge due to improper system fan damper position, inleakage at discharge points, or reentrainment of dust within collector.
4. High moisture in the air will foul the filters, causing excessive pressure drop and eventual blockage of flow.
5. Corrosion of material due to improper material selection, improper insulation, or dew point excursions.

6. Inadequate dust layer formation.
7. Bypassing due to failure of flow control.
8. When baghouses operate below the dew point of the gas or with dryer discharge gas with a high water vapor content, possible problems of condensation are most likely to occur during the winter, especially if filter units are located outdoors. In these cases, the baghouses should be thermally insulated from the atmosphere.
9. Baghouses must be safeguarded against dust explosions because considerable dust holdup in the baghouse can occur. Explosion panels that relieve directly or through short lines to the atmosphere are normal features of baghouses.
10. Filtration of oxidizable materials can constitute a serious fire hazard. The ignition source can be a burning particle from the dryer or the static charge carried by the dust particle. Because of this, baghouse systems should be equipped with automatic devices for closing off the airflow in case of fire, should be provided with adequate sprinklers or chemical fire control apparatus, and vents should be provided on the side containing the dust particles. To dissipate the accumulated charges, the filter cloth is woven with a few carbon or stainless steel wires that are properly grounded.

45.3.3 WET SCRUBBERS

45.3.3.1 Principle of Operation and Equipment Configuration

Emission control by wet scrubbing involves the separation of particles and gases from a contaminated gas stream by means of contact with a liquid. Usually, the liquid used is water, occasionally with a surface active agent added. They may also separate or neutralize gaseous pollutants by absorption if suitable chemicals are added to the scrubbing liquid (e.g., KMnO_4 or NaCl).

For PM collection, material transfer between the gas and the liquid phases can be done through different mechanisms. For gas removal, it is basically by diffusion. The particulate collection mechanisms involved in wet scrubbing can be one or more of the following basic mechanisms [2,52,53]:

1. Inertial impaction
2. Gravitational sedimentation
3. Direct interception
4. Brownian diffusion
5. Thermophoresis
6. Electrostatic precipitation
7. Diffusiophoresis

8. Condensation on particles
9. Coagulation

Scrubber types most frequently used for drying processes in general terms of geometric type, collection mechanism, equipment configuration, and method of operation are presented in [Table 45.10](#) and [Figure 45.4](#).

45.3.3.2 Advantages and Disadvantages [1,20,29,60]

Advantages of wet scrubbers are:

1. Can collect gases and particles simultaneously.
2. Ability to handle high-temperature, high-humidity gas streams.
3. Relatively small space requirements.
4. Ability to achieve high collection efficiencies on fine particulates.
5. No fire or explosion hazard if suitable scrubbing liquid is used (usually water).
6. Corrosive gases and mists can be recovered and neutralized.
7. Recovers soluble material, and the material can be pumped to another process for further treatment.
8. No secondary dust sources.
9. Low capital cost (if wastewater treatment system is not required).
10. They are generally more efficient than the dry collector; many separate particles less than 1 μ m in size.
11. Possible condensation of water vapor is a positive advantage because it improves the separation through diffusiophoresis and thermophoresis.

Disadvantages of wet scrubbers are:

1. May create water disposal problem.
2. Product is collected wet.
3. Corrosion problems.
4. Very small particles (submicrometer size) are difficult to wet, and so will pass through the plant.
5. Relatively high maintenance costs.
6. Relatively high operation energy costs, especially for high-efficiency systems that have large pressure drops.
7. Mist and vapors may be entrained in effluent gas streams.

45.3.3.3 Selection and Design

The design and selection of wet scrubbers usually focus on the estimation of two primary performance

parameters, the efficiency of removal (or the penetration) and the pressure drop. For a given scrubber, performance will depend on the particle diameter, collection mechanism(s), collecting element dimensions, gas flow rate, and liquid flow rate. The system pressure drop dictates the power requirements and the size of auxiliary equipment such as fans and pumps.

Detailed mathematical models and design equations for predicting the grade penetration and system pressure drop have been developed [2,18,29,52,56,57,59]. Summaries of some of these models are shown in [Table 45.11](#) and [Table 45.12](#). Pressure drop equations for other scrubbers can be found in standard Chemical and Environmental Engineering texts or handbooks [11–16].

45.3.3.4 Performance Characteristics and Operational Problems

In a general sense, wet scrubbers may be broadly classified as low-, medium-, and high-energy scrubbers. Low-energy scrubbers having 1 to 5 in. of water (250 to 1250 Pa) pressure drop include spray towers, which correspondingly have the lowest collection efficiency. The medium-energy group includes centrifugal scrubbers, atomizing impingement scrubbers, and certain packed bed scrubbers with typical pressure drops from 5 to 15 in. of water (1250 to 3750 Pa). High-energy scrubbers, such as the venturi type, have collection efficiencies up to 99.5% and greater pressure drops within the range of 15 to 35 in. of water (3750 to 8750 Pa) and collect particles as small as 0.5 μ m. However, their running costs are high.

Several general types of wet scrubbers are discussed in [Table 45.13](#), with a summary of their performance and ranges of operation.

Wet scrubbers display some problems as either high utility costs or low removal efficiencies. Most of the scrubber problems usually involve spray nozzle plugging, liquid circuit restrictions, and/or entrainment of droplets from the vessel. Such problems may include [20,53,61]:

1. Wet/dry zone buildup.
2. Nozzle plugging.
3. Scaling on the scrubber surface.
4. Localized corrosion and erosion that may develop leakage and deteriorate scrubber internals so that the power is not being used effectively.
5. Excessive entrainment that shows high particulate loading in the scrubber outlet and the presence of moisture in the duct.

TABLE 45.10
Wet Scrubber Configurations

Scrubber Type	Collection Mechanism	Equipment Configuration and Method of Operation
Plate	Inertial impaction	Consists of a vertical tower with one or more perforated plates that are irrigated with water and through which gas travels and is scrubbed. Usually, the scrubbing liquid is introduced at the top plate and runs down from stage to stage via downcomers or by trickling through the holes in the perforated plate. Scrubbers that use perforated plates are typically called <i>sieve tray towers</i> ; those that place rigid baffles opposite each perforation are called <i>impingement plate scrubbers</i>
Preformed spray	Inertial impaction	Consists essentially of a round or rectangular chamber into which scrubbing liquid is introduced through one or more sprays. Gases and particles flow through a chamber with sprays directed co-current, cross-current, or countercurrent to the flow, with the last being advantageous if gases are also to be removed in the scrubbing process. In this scrubber type, the particles are collected on liquid drops that have been atomized by spray nozzles
Packed bed	Centrifugal deposition, inertial impaction	Cylindrical and rectangular towers packed with materials such as Raschig rings and Berl saddles are normally used for the removal of pollutant gases and vapors. However, such systems have also been considered for the removal of PM. The scrubbing liquid is fed to the top of the tower. A distributor plate with a pattern of overflow pipes provides an even distribution of liquid to the top of the bed. The liquid flows down through the bed, wetting the packing, and thus provides interfacial surface area for mass transfer of the pollutant with the gas stream. The scrubbing liquid serves to wet, collect, and wash PM from the bed. There are three general ways in which the scrubber may be operated: crossflow, co-current flow, and countercurrent flow
Gas-atomized spray	Inertial impaction	Gas-atomized spray scrubbers use a high-velocity flow gas across a liquid surface to first atomize liquid into droplets and then accelerate the droplets, which can then be used as collectors of particles in the gas stream. A variety of atomizing scrubbers work this way: venturi, flooded disc, orifice, and similar types
Centrifugal	Centrifugal deposition	These are scrubbers in which liquid is sprayed into the unit and mixed with the rising vortex of gas to be cleaned. Centrifugal scrubbers impart a spinning motion to the gas passing through them. The spin may come from introducing gases to the scrubber tangentially or by directing the gas stream against stationary swirl vanes. These scrubbers serve as their own demister and are resistant to plugging
Moving bed		Moving bed scrubbers incorporate a zone of light mobile packing, usually plastic or glass spheres, in which gas and liquid can mix intimately. The gas stream passes upward through the packing, while liquid is sprayed up from the bottom and/or flows down over the top of the moving bed. Particle collection may be enhanced by using several moving bed stages in series. Continuous movement and agitation cleans the packing and reduces solid deposition

Source: From W. Licht, *Air Pollution Control Engineering*, Marcel Dekker, New York, 1988; S. Calvert, in *Handbook of Air Pollution Technology*, S. Calvert and H. Englund (Eds.), John Wiley & Sons, New York, 1984; A. Buonicore, in *Air Pollution Control Equipment*, L. Theodore and A. Buonicore (Eds.), Prentice-Hall, Englewood Cliffs, NJ, 1982; S. Calvert, J. Goldschmid, D. Leith, and D. Mehta, *Scrubber Handbook*, US EPA, NTIS PB 213 016, 1972; S. Calvert, *Chemical Engineering*, 54–68, August 29, 1977; D. Cooper, in *Handbook of Powder Science and Technology*, M. Fayed and L. Otten (Eds.), Van Nostrand Reinhold, New York, 1984; S. Calvert, in *Air Pollution*, 3rd ed., Vol. 4, A. Stern (Ed.), Academic Press, New York 1977; K. Schiffner and H. Hesketh, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992; and J. Kelly, in *Gas Cleaning for Air Quality Control*, J. Marchello and J. Kelly (Eds.), Marcel Dekker, New York, 1975. With permission.

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|---|--|
| 6. Air leaks into the scrubber or ductwork can lead to an increase in fan horsepower and/or a decrease in the amount of gas drawn through the scrubber. | include provisions for preventing damage due to freezing when the scrubber is not in operation. |
| 7. Wet scrubbers are subject to possible freezing in cold weather; hence cold weather operation must | Detailed information about wet scrubber problems, troubleshooting, and upgrading existing scrubbers in |

order to obtain improved product recovery and pollution control and better performance are discussed by Buonicore [53], Calvert [61], and Gilbert [62].

45.4 DRYER EMISSION CONTROL IN SELECTED INDUSTRIAL PROCESSES

This section provides a brief description of the emission characteristics and air pollution control measures employed in selected industrial processes. The reader is referred to the literature cited for details.

45.4.1 MINERAL PRODUCTS INDUSTRY

Drying is an important operation carried out in many mineral processing plants to improve product quality, facilitate handling and conveying, improve subsequent drying processes, and/or to meet customer specifications. The main atmospheric emission from mineral process dryers is PM; other emissions include gaseous and VOCs, depending on the composition of the raw materials and the fuel used.

The following sections describe the new performance standards for dryers in mineral industries and

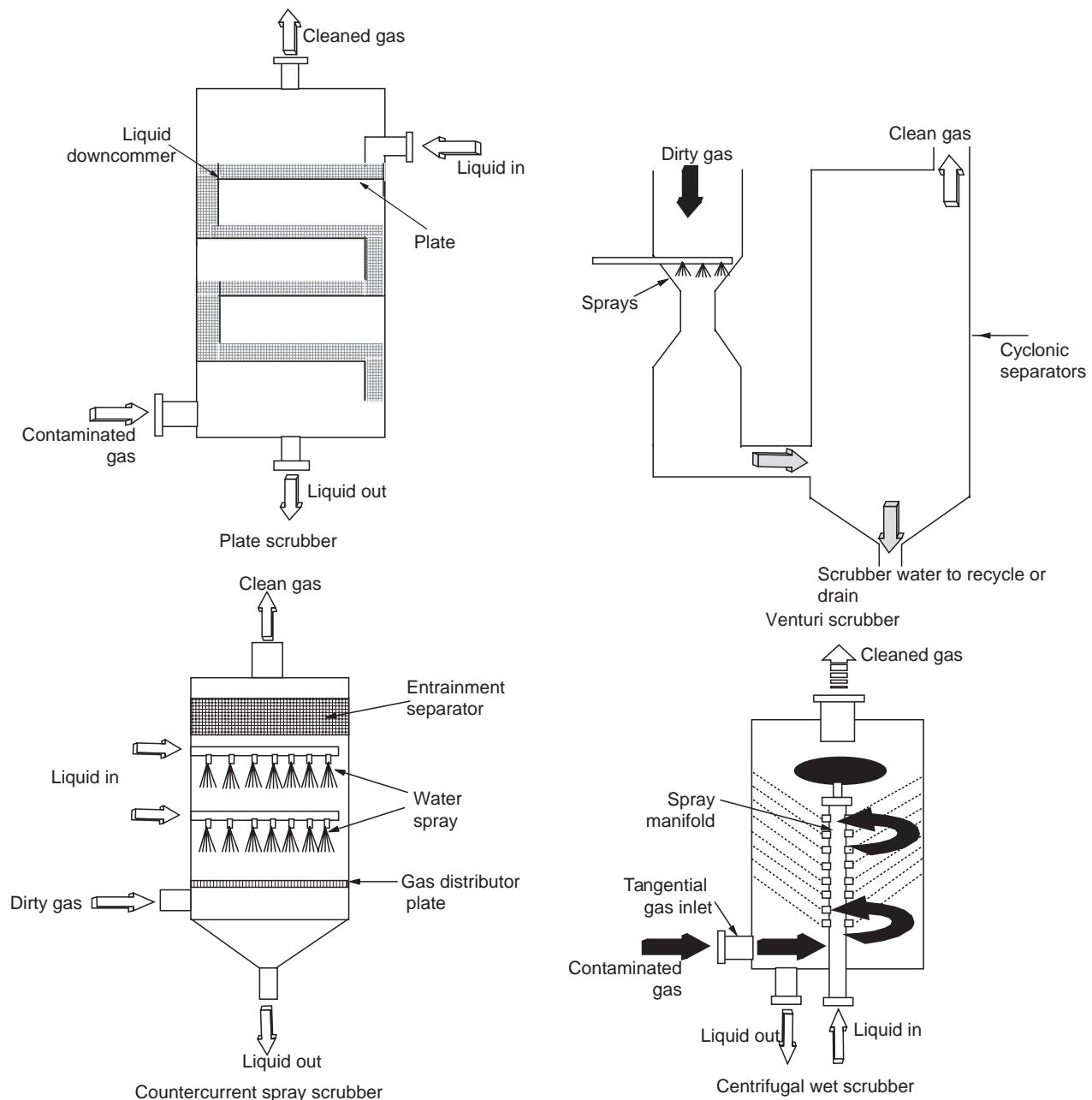


FIGURE 45.4 Wet scrubber configurations.

(continued)

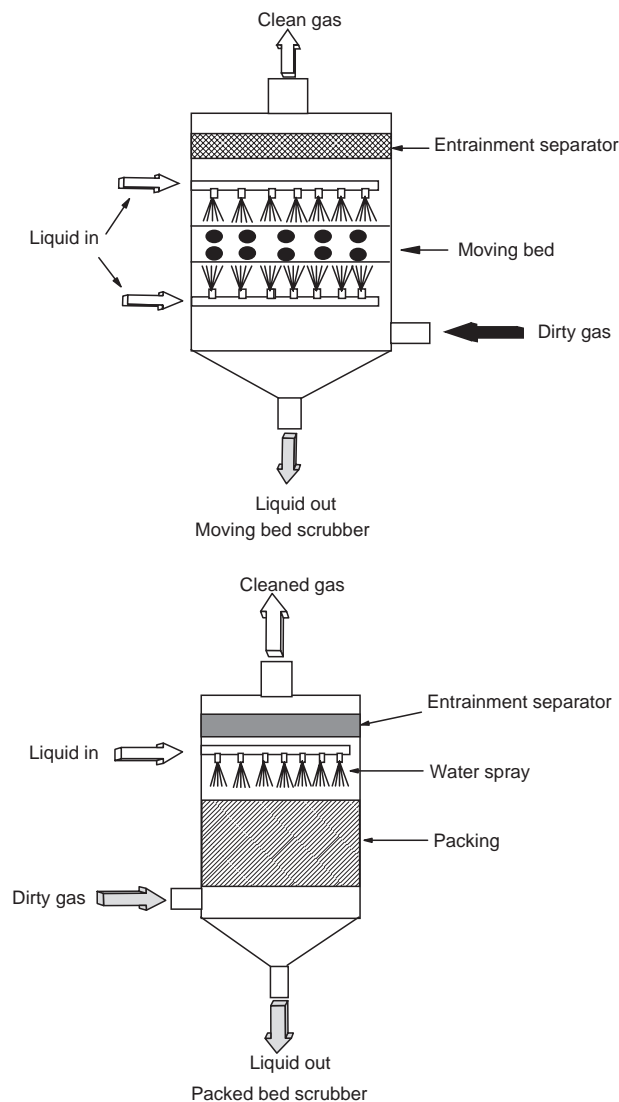


FIGURE 45.4 (continued)

then describes, in more detail, selected mineral processes and their emission characteristics and control techniques commonly, but not exclusively, used.

45.4.1.1 Environmental Protection Agency New Performance Standards

According to the U.S. Environmental Protection Agency (EPA) standards [63], no emissions shall be discharged into the atmosphere from any facilities that

- (a) contain PM in excess of 0.092 grams per dry standard cubic meter (g/dscm) (0.04 grain per dry standard cubic foot [gr/dscf]) for calciners and for calciners and dryers installed in series

and in excess of 0.057 g/dscm (0.025 gr/dscf) for dryers; and

- (b) exhibit greater than 10% opacity, unless the emissions are discharged from an affected facility using a wet scrubbing control device.

In these standards, a *mineral processing* plant is any facility that processes or produces any of the following minerals, their concentrates, or any mixture, or a combination of their minerals: alumina, ball clay, bentonite, diatomite, feldspar, fire clay, fuller's earth, lightweight aggregate, gypsum, industrial sand, coal, magnesium compounds, perlite, roofing granules, talc, titanium dioxide, and vermiculate. Calciners and dryers installed in series means a calciner and dryer installed such that the exhaust gases from one flow through the other and then the combined exhaust gases are discharged to the atmosphere. It should be noted that emission standards are different in different countries. The types of dryers to which the standards apply include rotary (direct), rotary (indirect), fluid-bed, vibrating tray, flash, and spray dryers.

The standards are based on the emission levels achievable using well-designed and operated fabric filters, wet scrubbers, or ESPs. All of these systems are considered the best demonstrated technology for controlling emissions from calciners and dryers in this source category, depending on the type of calciner, dryer, and mineral industry.

45.4.1.2 Coal Preparation [64–69]

Coal preparation is a collective term for the physical and mechanical processes such as breaking and crushing, screening, wet and dry concentration, and dewatering. The primary purposes of a coal preparation plant are to improve the coal quality, meet customer specifications, reduce the amount of ash and the sulfur content of coal, and reduce transportation expenses.

Wet cleaning systems involved in coal preparation are generally wet and do not, in themselves, cause air pollution problems. However, the auxiliary processes, such as drying, can be major emission sources. Drying of coal is carried out to increase its calorific value, facilitate its transport, and to prevent freezing in cold climates. Several types of thermal dryers are used for coal processing (e.g., fluidized bed, flash, and multilouvered). The most prevalent type of dryer in use today is the fluidized bed dryer. Particulate emissions occur predominately from ultrafine (200 mesh) coal particles that are entrained and carried from the dryer by the combustion gases used as the drying medium. A flash dryer is the second most used type of dryer.

TABLE 45.11
Grade Penetration Models for Wet Scrubbers

Scrubber Type	Penetration, Fraction
Sieve plate	$p_t = \exp(-40FK_p)$ $K_p = \frac{u_h dp_a^2}{9\mu_G d_h}$ $-\ln F = 0.184u_{Gs}\rho_G^{0.5} + 0.45$ $d_{pa} = d_{ps} \left[C' \rho_l \right]^{\frac{1}{2}}$ $C' = 1 + \left[1.257 + 0.4 \exp\left(-1.1 \frac{d_{ps}}{2\lambda_g}\right) \right] \frac{2\lambda_g}{d_{ps}}$ $\lambda_g = \frac{kT}{2\pi P d_{mo}^2}$ $d_{pc} = 0.4 \left(\frac{\mu_G d_h}{u_h F} \right)^{\frac{1}{2}}$
Impingement plate	$p_t = \exp(-0.693dp_a^2/dp_c^2)$ $d_{pc} = 0.4 \left(\frac{1.37\mu_G n_h d_h^3}{Q_G} \right)^{\frac{1}{2}}$
Massive packing	$p_t = \exp\left(-7.0 \frac{ZK_p}{d_c}\right)$ $d_{pc} = \left(\frac{f_v d c^2 \mu_G}{u_{Gs}} \right)^{\frac{1}{2}}$
Preformed spray	<p>Vertical counterflow:</p> $p_t = \exp\left(\frac{-3Q_L u_t Z \eta_d}{4Q_G r_d (u_t - u_G)}\right) = \exp\left(-0.25 \left(\frac{A_d u_t \eta_d}{Q_G}\right)\right)$ $A_d = \frac{3Q_L Z}{r_d (u_t - u_G)}$ $\eta_d = \left(\frac{K_p}{K_p + 0.7}\right)^2$ <p>Cross-flow:</p> $p_t = \exp\left(\frac{-3Q_L Z \eta_d}{4Q_G r_d}\right) = \exp\left(\frac{-0.25 A_d u_r \eta_d}{Q_G}\right)$ $p_t = \exp\left(\frac{Q_L u_G \rho_L d_d}{55 Q_G \mu_G} F(K_{po}, f)\right)$ $F(K_{po}, f) = \left[-0.7 - K_{po} f + 1.4 \ln\left(\frac{K_{po} f + 0.7}{0.7}\right) + \frac{0.49}{0.7 + K_{po} f} \right] \frac{1}{K_{po}}$ <p>For large-scale, gas-atomized scrubbers, $f = 0.5$</p> $d_d = \frac{0.0585}{u_G} \left(\frac{\sigma}{\rho_L}\right)^{0.5} + 1884 \left(\frac{\mu_L}{(\sigma \rho_L)^{0.5}}\right)^{0.45} \left(\frac{Q_L}{Q_G}\right)^{1.5}$
Gas atomized	

Source: From S. Calvert, in *Handbook of Air Pollution Technology*, S. Calvert and H. Englund (Eds.), John Wiley & Sons, New York, 1984; S. Calvert, J. Goldschmid, D. Leith, and D. Mehta, *Scrubber Handbook*, US EPA, NTIS PB 213 016, 1972; D. Cooper, in *Handbook of Powder Science and Technology*, M. Fayed and L. Otten (Eds.), Van Nostrand Reinhold, New York, 1984; and S. Calvert, in *Air Pollution*, 3rd ed., Vol. 4, A. Stern (Ed.), Academic Press, New York, 1977. With permission.

The size consistency of the coal being dried and the velocity of the gases through the bed are the major factors determining the air pollution potential of the plant. Emissions from dryers in coal-cleaning plants consist mainly of entrained coal fines and VOCs released from the coal, in addition to the standard combustion products [69]. A typical composition of the combustion products emitted from dryers is shown in Table 45.14. Sulfur oxides vary with the

sulfur content of the coal, while the other emissions are a function of the burner design and excess air. Total emissions depend on the amount of energy required to dry the coal to the desired level.

Typical emission control systems in coal dryers are cyclone collectors and high-efficiency wet scrubbers or packed towers in series. Fabric collectors are not usually used because of fire hazard. High-efficiency cyclone, a venturi scrubber, and a demister

TABLE 45.12
Pressure Drop Equations for Wet Scrubbers

Scrubber Type	Pressure Drop
Plate	$\Delta P = \Delta P_{\text{dry}} + \Delta P_{\text{wet}}$ $\Delta P_{\text{dry}} = 588 \left(\frac{d_h}{d_p^2} \right)^2$ $\Delta P_{\text{wet}} = \rho_L g H_{\text{weir}}$
Preformed spray	Cross-current and co-current flow: $\Delta P = 1.03 \times 10^{-3} \left(\frac{Q_L}{Q_G} \right) u_G^2$
Venturi	$\Delta P = - \left(\frac{2\rho_L u_G^2 Q_L}{g_c Q_G} \right) [1 - x^2 + (x^4 - x^2)^{0.5}]$ $x = \left(\frac{3I_{\text{C,Do}} \rho_G}{16d_d \rho_L} \right) + 1$

Source: From S. Calvert, in *Handbook of Air Pollution Technology*, S. Calvert and H. Englund (Eds.), John Wiley & Sons, New York, 1984; S. Calvert, J. Goldschmid, D. Leith, and D. Mehta, *Scrubber Handbook*, US EPA, NTIS PB 213 016, 1972; D. Cooper, in *Handbook of Powder Science and Technology*, M. Fayed and L. Otten (Eds.), Van Nostrand Reinhold, New York, 1984; and S. Calvert, in *Air Pollution*, 3rd ed., Vol. 4, A. Stern (Ed.), Academic Press, New York, 1977. With permission.

installed on a thermal dryer can reduce PM from 4.5 to 0.1 g/m³ [64]. The estimated emission factors for coal drying plants expressed as kg/t of coal dried are summarized in Table 45.15.

45.4.1.3 Portland Cement Production [64–66,70,71]

Portland cement is a fine powder that mainly consists of calcium- and silica-based materials such as hydraulic cement minerals, dicalcium silicate, and tricalcium silicate, with smaller amounts of alumina and iron compounds such as tricalcium aluminates and tricalcium aluminoferrite. The Portland cement production process involves quarrying and crushing, grinding and blending, clinker production, drying and calcination, and finish grinding and packaging. Generally, there are wet process and dry process Portland plants. In cement production using the dry process, the kiln feed may be dried in a rotary dryer or in combined drying and grinding units.

The dryer is the second major source of emissions after the kiln in cement production, especially if kiln exit gases are utilized for drying. PM is the primary emission from cement dryers. Emissions also include the combustion products from the fuel used to supply heat for the direct drying operation. Typical emission factor for cement dryers without control is 48 kg/t of cement [65].

Emission control systems for cement dryers must be designed for higher temperatures. Typical systems include multicyclones or other types of mechanical collectors, ESPs, or a combination of both. Where fabric filters are applied to drying operation, they must be fiber type suitable for temperature above 120°C (for more information on the selection of fabric filters see Table 45.8).

45.4.1.4 Asphalt Concrete [64–66,72]

Asphalt concrete paving material is a mixture of graded aggregates and asphalt cement. The aggregates, which include stone, sand, and mineral dust or filler, and include reclaimed asphalt pavement, make up about 92 to 96% of the total mixture by weight. In a typical asphalt concrete plant, the aggregate is proportioned and charged to a thermal dryer, typically a rotary dryer, and heated to temperatures ranging from 135 to 165°C. The hot, dried aggregate is then screened, transferred to storage bins, and introduced into a mill where it is mixed with hot asphalt. Both countercurrent and co-current direct-fired rotary dryers are used commonly.

Particulate emissions from the rotary dryer are by far the largest source of emissions in asphalt concrete plants. Dryer dust emissions increase with air mass velocity, increasing rate of rotation, and feed rate. Rotary dryer emissions vary with the aggregate size and type of fuel. Uncontrolled particulate emissions average approximately 22.5 kg/t. The new EPA regulations for asphalt concrete processes prohibit emissions of particulate in excess of 0.0915 g/m³ of dry exhaust gas.

Other emissions include gaseous emissions such as carbon monoxide, hydrocarbons, sulfur oxides, nitrogen oxides, hydrogen oxides, VOCs, and aldehydes. Gaseous emissions are much lower than the particulate, generally mounting to less than 0.5 kg/t. The EPA emission factors for various gaseous pollutants in asphalt concrete manufacture are listed in Table 45.16.

Dust control systems usually include hoods and enclosures connected to a local exhaust ventilation system that discharges to cyclone collectors and scrubbers or a baghouse in series. The most common type of control equipment is the baghouse; these are increasing in use as more stringent pollution control codes are adopted. These filters provide excellent collection efficiency with little or no visible emissions and the collected dry fines are sometimes usable in concrete mixes. However, if the exhaust gas temperature is at or near the dew point of the gas, condensation of moisture on the fabric filters is always a possibility and poses a serious fire hazard. The other type of the commonly used control equipment is the venturi wet

TABLE 45.13
Summary of Wet Scrubber Types, Performance, and Ranges of Operations

Design	Efficiency	Cut Diameter, μm	ΔP	Liquid Consumption	Gas Velocity
Plate	90–98% on 1- μm particle	1 μm for 3.2-mm diameter holes	250–1,000 Pa	0.4–0.8 m ³ /1,000 m ³ of gas	Superficial: 3 m/s
Packed bed	99% on 5- μm particle	1.5 μm with 2.5 cm packing size	Cross flow: 80200 Pa/m depth Counter flow: 160–1,250 Pa/m depth	Cross flow: 0.133–0.53 m ³ /1,000 m ³ of gas Counter flow: 1.35–2.7 m ³ /1,000 m ³ of gas	Superficial: 12 m/s
Preformed spray	70–80%	2 μm	100–200 Pa/m depth	0.4–2.7 m ³ /1,000 m ³ of gas	15–45 cm/s
Centrifugal	87% on 5- μm particle	4–5 μm without spray	500–1,500 Pa	0.27–0.67 m ³ /1,000 m ³ of gas	Cyclone inlet velocity: 15–25 m/s
Venturi	99% on 5- μm particle	—	1,450–17,500 Pa	0.27–1.33 m ³ /1,000 m ³ of gas	Through venturi throat: 60–120 m/s
Moving bed	99% on 2- μm particle	—	750–1,250 Pa per stage	—	
Orifice type	90–95%	—	750–2,500 Pa	—	

Source: From S. Calvert, in *Handbook of Air Pollution Technology*, S. Calvert and H. Englund (Eds.), John Wiley & Sons, New York, 1984; A. Buonicore, in *Air Pollution Control Equipment*, L. Theodore and A. Buonicore (Eds.), Prentice-Hall, Englewood Cliffs, NJ, 1982; and J. Kelly, in *Gas Cleaning for Air Quality Control*, J. Marchello and J. Kelly (Eds.), Marcel Dekker, New York, 1975. With permission.

TABLE 45.14
Gaseous Emission from Coal Dryers

Compound	Concentration, ppm	Emission Rate, lb/10 ⁶ BTU ^a
NO _x	40–70	0.39–0.68
SO _x	0–11	0–0.09
HC ^b	20–100	0.07–0.35
CO	<50	<0.30

^aBased on heat input to dryer.

^bExpressed as methane.

Source: From R. Gerstle, in *Handbook of Air Pollution Technology*, S. Calvert and H. Englund (Eds.), John Wiley & Sons, New York, 1984. With permission.

scrubber. Generally, a high-pressure (6-kPa gauge) venturi scrubber is required to meet the regulated requirements. In addition to controlling particulate emissions, the venturi scrubber is likely to remove some of the process hydrocarbon emissions from the exhaust gas.

45.4.2 CHEMICAL PROCESS INDUSTRY

The chemical process industry can be defined as a process industry in which crude raw materials of a mineral or petroleum origin are connected by way of chemical intermediates into semifinished or final products. For this section, consideration of the chemical industry has been restricted to the dryers involved

TABLE 45.15
Emission Factors for Coal Dryers

Pollutant	Operation	
	Fluidized Bed Dryer, kg/MT	Flash Dryer, kg/MT
<i>Particulates</i>		
Before cyclone	10	8
After cyclone	6	5
After scrubber	0.05	0.02
SO ₂		
After cyclone	0.22	—
After scrubber	0.13	—
NO _x		
After scrubber	0.07	—
<i>Volatile organic compounds</i>		
After scrubber	0.05	—

Source: From L. Simmons and L. Lambert, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992. With permission.

TABLE 45.16
Emission Factors for Selected Gaseous Pollutants from a Conventional Asphalt Concrete Plant Stack

Material Emitted	Emission Factor	
	g/mg	lb/t
Sulfur oxides	146 × S ^a	0.292 × S ^a
Nitrogen oxides	18.000	0.036
Volatile organic compounds	14.000	0.028
Carbon monoxide	19.000	0.038
Polycyclic organic material	0.013	0.000026
Aldehydes	10.000	0.02
Formaldehyde	0.075	0.00015
2-Methyldehyde (isobutyraldehyde)	0.650	0.0013
1-Butanal (<i>n</i> -butylaldehyde)	1.200	0.0024
3-Methylbutanal (isovaleraldehyde)	8.00	0.016

^aS = sulfur content of fuel by weight.

Source: From K. Gunkel, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992; and U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 8.1—Asphaltic Concrete Plants, EPA, Research Triangle Park, NC, October, 1986. With permission.

in the manufacture of fertilizers, sodium carbonate, and detergents. These processes are selected to illustrate the main emission control techniques and air pollution problems encountered with chemical process dryers.

45.4.2.1 Phosphate Fertilizers [74–80]

Phosphate Rock Preparation—Domestic phosphate rocks are essentially fluorapatite admixed with various proportions of other compounds of calcium, fluorine, iron, aluminum, and silicon. Phosphate rock preparation involves beneficiation to remove impurities, drying to remove moisture, and grinding to improve reactivity. Phosphate rock, when very finely pulverized, has limited direct use as a fertilizer. However, it is mainly used as a raw material for the manufacture of phosphate acid, superphosphate, phosphorus, and phosphorus compounds.

After beneficiation, the washed rock may contain 7 to 20% moisture, which is reduced to between 1 and 2% moisture using direct-fired rotary dryers. Emissions expected from the dryer consist primarily of fine rock dust. Some sulfur dioxide may also be present in the dryer exhaust from the combustion of sulfur in the fuel. Phosphate rock dryers are usually equipped with dry cyclones, followed by wet scrubber systems for control of rock dust. The scrubber

systems will operate at typical particulate collection efficiencies of 97 to 98% with emissions ranging from 0.1 to 0.3 kg/t of phosphoric anhydride (P_2O_5) [77].

Normal Superphosphate Production—Normal superphosphate is the material resulting from the reaction of ground phosphate rock and 65 to 75% sulfuric acid. This fertilizer material contains from 15 to 21% P_2O_5 . The process includes mixing and reaction, dropping into a den, curing, granulation, drying, cooling, and screening.

If the normal superphosphate is granulated, the material passes through a direct-fired rotary dryer. Emissions from the dryer will include gaseous fluorides, fertilizer dust, and sulfur oxides, especially if the dryer is fired with high-sulfur oil. At a typical normal superphosphate plant, the emissions from the dryer are controlled by a cyclone collector as the primary collector, followed by a wet scrubber or baghouse, with efficiencies ranging from 90 to 99% [74,77]. In addition to controlling the particulate emissions, the wet scrubber serves to remove the fluoride emissions by scrubbing the off-gases with recycled water. Scrubber types that can be used are cyclonic, venturi, impingement, jet ejector, and crossflow-packed spray scrubbers [76].

Triple Superphosphate Production—Triple superphosphate (TSP) is a fertilizer material with a phosphorus content of over 40%, measured as phosphorus pentoxide (P_2O_5). Two processes have been used to produce triple superphosphate: run-of-the-pile (ROP-TSP) and granular (GTSP). These processes involve reaction, granulation, drying, cooling, screening, and conveying to the storage building. In the drying stage, the slurry-wetted granules are fed to a rotary gas- or oil-fired rotary dryer in which excess water is removed. Emissions from the dryer include particulates, sulfur oxides, and fluorides such as silicon tetrafluoride and hydrogen fluoride. For emission control, the plant is usually equipped with cyclone separators for removal of a portion of the dust before using wet scrubbers to remove fluorides. Typically, two scrubbers are placed in series. The first is generally a venturi scrubber that utilizes either gypsum pond water or dilute phosphoric acid as the scrubbing liquor. The final cleaning device is typically a packed scrubber that uses water from the gypsum pond to control gaseous emissions. The effectiveness of the abatement systems for the removal of fluorides and particulates varies from plant to plant, depending on a number of factors [79]: (1) inlet fluoride concentration; (2) outlet or saturated gas temperature; (3) composition and temperature of the scrubbing liquid; (4) scrubber type and transfer units; and (5) effectiveness of the entrainment separation. Control efficiency is enhanced by increasing the number of scrubbing stages

and by using a freshwater scrub in the final stage. Reported efficiencies for fluoride control range from <90 to >99%, depending on inlet fluoride concentration and the system employed. The maximum allowable fluoride emission from triple phosphate plants is 0.1 kg/t P_2O_5 fed [76]. A schematic diagram of the process with its emission control system is shown in Figure 45.5.

Ammonium Phosphates—Ammonium phosphates are produced mainly as either monoammonium phosphate or diammonium phosphate by reacting phosphoric acid with anhydrous ammonia. In many plants, 93 to 98% sulfuric acid is used to control the composition of the final product. After the reactor and the granulator, the moist ammonium phosphate granules are transferred to a rotary dryer.

Exhaust gases from the dryer contain ammonia, fluorides, and particulates. These streams are commonly combined and passed through cyclones for particulate recovery and then to primary scrubbers. The primary scrubbers collect ammonia at efficiencies greater than 99% using a solution of 20 to 30% phosphoric acid in water. Materials collected in the cyclone and primary scrubbers are returned to the process. The exhaust is sent to secondary scrubbers in which recycled gypsum pond water is used as a scrubbing liquid to control fluoride emissions; the scrubber effluent is returned to the gypsum pond. Primary scrubbing equipment commonly includes venturi and cyclonic spray towers. Impingement scrubbers and crossflow-packed bed spray scrubbers are used as secondary controls. Emission control efficiencies for ammonium phosphate plant control equipment are reported as 94 to 99% for ammonia, 75 to 99.8% for particulates, and 74 to 94% for fluorides [80]. Average controlled emission factors for the dryer are 0.75 and 0.02 kg/t P_2O_5 for particulates and fluorides, respectively [76]. Figure 45.6 shows a schematic diagram of the process, illustrating the dryer and its emission control system.

45.4.2.2 Muriate of Potash [78]

The term *potash* is generally used to describe compounds that contain the element potassium. In the fertilizer industry, all potassium values are, by convention, measured in equivalent weight of potassium oxide. The most popular potassium-based fertilizer is potassium chloride (KCl); in the fertilizer trade it is referred to as muriate of potash. The operations involved in the production of muriate of potash from sylvinite (crude KCl) ore are crushing, scrubbing, separation, conditioning, flotation, crystallization, centrifuging, drying, and screening.

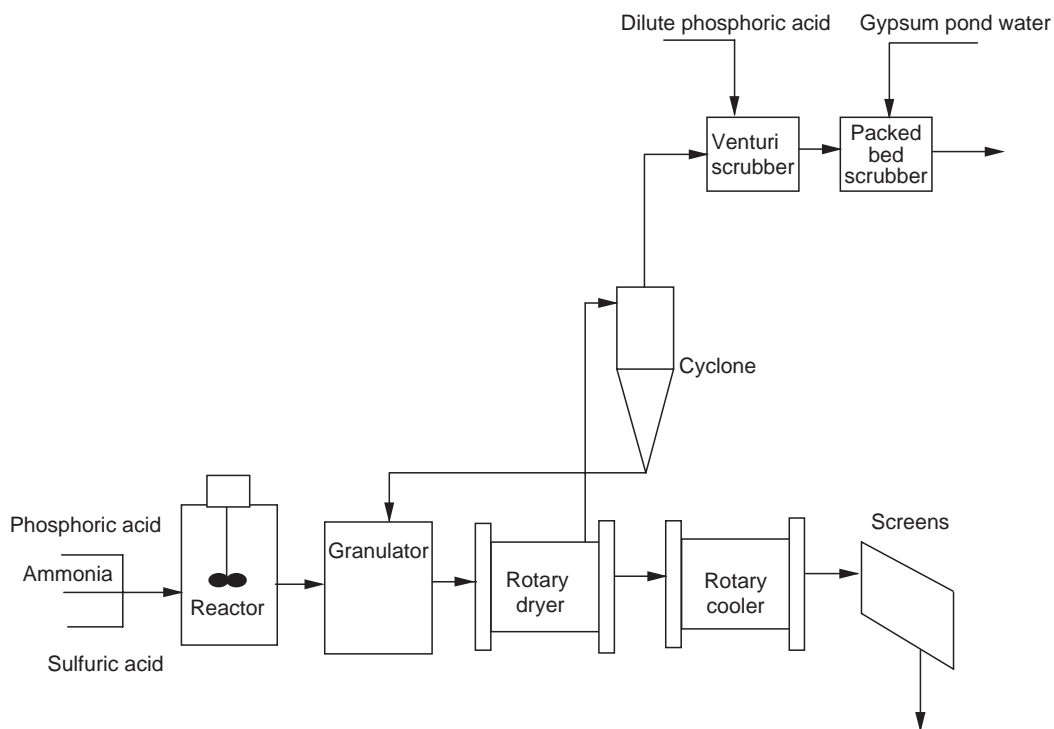


FIGURE 45.5 Granular triple superphosphate process flow diagram with rotary dryer emission control system.

Dryers in the potash industry are the major source of atmospheric emissions. Direct-fired rotary dryers are commonly used and fines are entrained with the high exit gas velocity. Particulate control equipment

used on potash dryers includes cyclones, wet scrubbers, and baghouses. A comparison of the effectiveness of these collectors for controlling particulate emissions from potash dryers is illustrated in [Table 45.17](#).

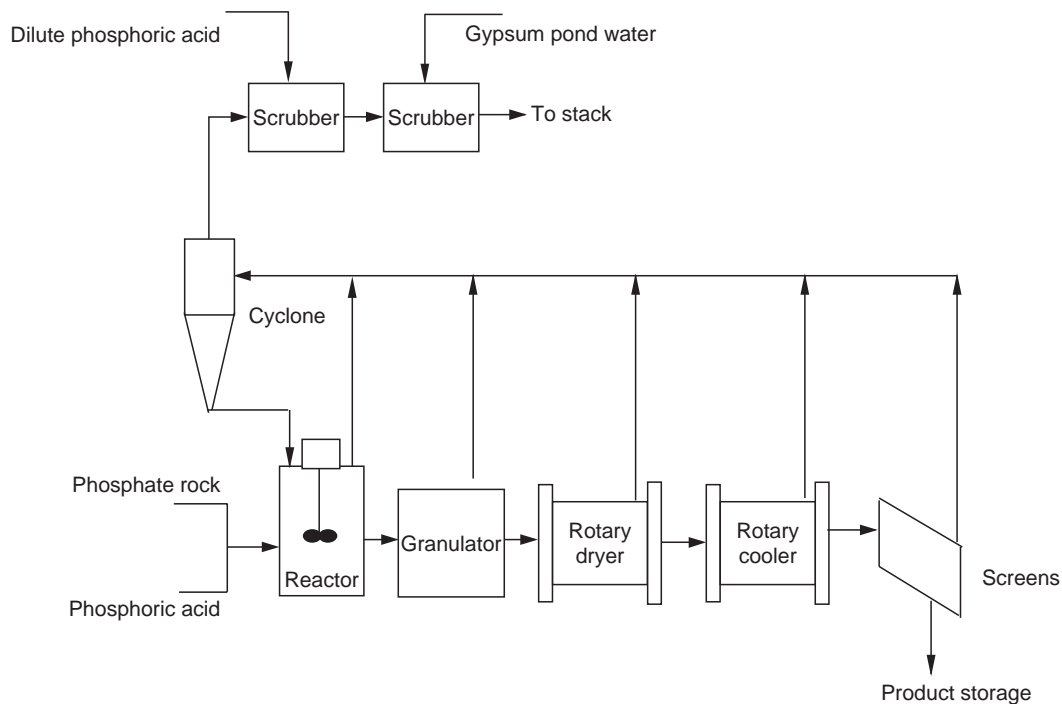


FIGURE 45.6 Ammonium phosphate process flow diagram and emission control system.

TABLE 45.17
Controlled Emissions from Gas-Fired Rotary
Potash Dryers

Dryer Capacity t/y product	Control Equipment	Controlled Emissions t/y	Uncontrolled Emissions t/y
350,000	Dry cyclone	450	7000
200,000	Venturi scrubber	14	—
120,000	Baghouse	0.2	—

Source: From L. Beck, in *Handbook of Air Pollution Technology*, S. Calvert and H. Englund (Eds.), John Wiley & Sons, New York, 1984. With permission.

45.4.2.3 Soap and Detergents [37,75,81,82]

The term “soap” refers to a particular type of detergent in which the water-soluble group is carboxylate and the positive ion is usually sodium or potassium. The term “synthetic detergents” applies broadly to cleaning and laundering compounds containing surface-active (surfactant) compounds along with other ingredients. Synthetic detergents are used in different cleaning operations containing surfactants to remove dirt, builders to treat water to improve the efficiency of the surfactants by softening the water, and additives to improve cleaning or physical properties. Additives to detergents include bleaches, anticorrosion agents, antisoil redeposition agents, foam boosters, perfume, optical brighteners, and fillers.

The main atmospheric pollution problem in soap manufacturing is odor. Odors emanating from the spray dryer may be controlled by scrubbing with an acid solution. The production of soap powder by spray drying is the single largest source of dust in the manufacture of synthetic detergents. The large sizes of the particulates from synthetic detergent drying means that high-efficiency cyclones installed in series can achieve satisfactory control. Dust emissions from other finishing operations can be controlled by dry filters such as baghouses. Currently, no emission factors are available for soap manufacturing. No information on hazardous air pollutants, VOCs, ozone depleters, or heavy metal emissions information were found for soap manufacturing [82].

The manufacture of the granular detergents by spray drying is of paramount interest. This process includes three steps: slurry preparation, spray drying, and granular handling (cooling, additive blending, and packaging). The product is dried in either concurrent-flow or countercurrent spray drying towers with drying air temperatures in the range of 250 to

400°C for concurrent operation and from 250 to 350°C for countercurrent operation. The nozzle atomizing pressure is in the range of 30 to 60 atm.

The exhaust air from the spray dryer contains two main contaminants. The first is fine detergent particles. These particles consist mainly of detergent compounds in addition to some uncombined phosphates, sulfates, and other mineral compounds. The second consists of organic compounds vaporized in the spray dryer. These organic compounds originate primarily from the surfactants used in the feed slurry. The amount vaporized depends on many variables such as tower temperature and the volatility of organics used in the slurry. These vaporized organic materials condense in the tower exhaust airstream into droplets or particles. Paraffin alcohols and amides in the exhaust stream can result in a highly visible plume that persists after the condensed water vapor plume has dissipated.

Opacity and organic emissions are influenced by granular temperature and moisture at the end of drying, temperature profiles in the dryer, and formulation of the slurry. A method for controlling visible emissions would be to remove offending organic compounds (i.e., by substitution) from the slurry. Otherwise, tower production rate may be reduced thereby reducing air inlet temperatures and exhaust temperatures. Lowering production rate will also reduce organic emissions.

Several separation and control approaches have been developed to collect the fine dust and organic compounds in the spray dryer effluent. For the control of particulate emissions, cyclones and cyclonic impingement scrubbers are frequently employed as the primary collectors. Secondary collection equipment is used to collect fine dust and organic aerosols that pass through the primary collectors. Cyclonic impingement scrubbers are usually followed by mist eliminators, while cyclones are followed by wet scrubbers, scrubber precipitators, or fabric filters. Spray chambers, packed scrubbers, as well as venturi scrubbers have been used. Problems may be encountered when filters are used as control devices for detergent drying towers as the exhaust gas is very humid (near the dew point) and the product particles are sticky at high temperatures. These conditions result in the condensation of water vapor and organic aerosols and thus binding the filter fabrics. Typical control efficiencies and emission factors for various control schemes for a detergent spray dryer are shown in Table 45.18.

45.4.2.4 Sodium Carbonate (Soda Ash) [83–85]

Sodium carbonate (Na_2CO_3), commonly referred to as soda ash is a lightweight crystalline solid, moderately soluble in water, usually containing 99.3% Na_2CO_3 . Soda ash is made commercially by four

TABLE 45.18
Particulate Emissions Factors for Detergent Spray Dryer

Control Device	Efficiency, %	Emission Factor, kg/mg of Product
Uncontrolled	NA	45
Cyclone	85	7
Cyclone with		
spray chamber	92	3.5
packed scrubber	95	2.5
venturi scrubber	97	1.5
wet scrubber	99	0.544
scrubber/ESP	99.9	0.023
Fabric filter	99	0.544

Source: From R. Scherr, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992 and U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 6.8—*Soap and Detergents*, EPA, Research Triangle Park, NC, 1995. With permission.

different processes: monohydrate; sesquicarbonate; direct carbonation; and Solvay process. Environmental issues and increasing fuel costs have been contributed to the decline in the production of soda ash by the Solvay process and hence, new soda ash plants are likely to use the monohydrate or the direct carbonation processes.

Three types of dryers are used for drying in the monohydrate and direct carbonation processes: rotary steam tube, rotary gas-fired, and steam tube fluid bed. Sodium carbonate fines are emitted from all of these dryers. Estimated uncontrolled particulate emission rates, particulate concentrations, and exit

gas flow rates extrapolated from EPA test data are presented in Table 45.19 [83,84].

Particulate emission control equipment applicable to dryers in soda ash plants includes cyclones, wet scrubbers, ESPs, and baghouses. Venturi scrubbers are used to control emissions from rotary steam tube dryers. Cyclones in series with venturi scrubbers are used to control emissions from fluid-bed steam tube dryers. Both venturi scrubbers and ESPs have been used to control emissions from gas-fired dryers. However, the use of ESPs or baghouses can result in operation problems because of the high humidity of the dryer exit gas and the fact that the collected soda ash particles are quite soluble and hygroscopic. This wet, sticky dust adheres to the electrodes and hoppers of the ESP or blinds and cakes the fabrics in the baghouse. Controlled emission factors of total PM from rotary soda ash dryers and fluid-bed soda ash dryers are 0.25 and 0.019 kg/mg of product, respectively [85].

45.4.3 PHARMACEUTICAL INDUSTRY [37,86–90]

The processes and unit operations that are commonly used in the pharmaceutical industry include chemical synthesis, fermentation, product recovery and purification, extraction, formulation, filtration, crystallization, and drying. Drying is applied to pharmaceutical products to remove excess solvents, as these heat-sensitive materials are more stable in dry than in wet form.

Because of the importance of product purity and the presence of toxic and biologically active constituents, pharmaceutical manufacturing operations must be very clean and closely controlled. Controlling air emissions is considered important to avoid the inadvertent contamination of products and to prevent occupational and community exposure to harmful substances. Using emission controls is also important

TABLE 45.19
Uncontrolled Particulate Emissions from Steam Tube Dryers in the Monohydrate and Direct Carbonation Processes

Steam Tube Dryer Type	Production Rate, mg/h	Particulate Emission Factor, kg/mg Feed	Particulate Emission Rates, 1000 kg/h	Exit Gas Particulate Concentration, g/dNm ³
Rotary	23	25.5–33.9	0.581–0.767	69–77
Rotary	63	25.5–33.9	1.62–2.15	69–77
Fluid bed	63	51.5–116	3.27–7.35	42–91
Fluid bed	113	51.5–116	6.08–13.6	42–91

Source: From M. Barboza and N. Haymes, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992 and U.S. Environmental protection Agency (EPA), *Sodium Carbonate Industry*—Background Information for Proposed Standards—Draft EIS, EPA-450/3-80-029 a, U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards, Emission Standards Division, Research Triangle Park, NC, August, 1980. With permission.

in recovering high-priced solvents often removed during the drying operation.

Dryer emissions in the pharmaceutical industry are similar to those in many chemical process dryers in that both gaseous and particulate emissions can be significant. However, of greatest concern are VOCs. Many of these compounds are photochemically reactive and contribute to tropospheric ozone formation. An added concern is that some of these compounds are toxic; both short-term and long-term exposure to them are undesirable and can lead to adverse health effects.

Typical control measures used in pharmaceutical industry to reduce VOCs (including odors) are condensation, absorption (spray towers, venturi scrubbers, packed columns, and plate columns), adsorption (using activated carbon beds), thermal destruction (flares, boilers and process heaters, thermal incinerators, and catalytic incinerators), and vapor containment. Water- or brine-cooled condensers are the most common control devices, with carbon adsorbers in occasional use. Where the main objective is not solvent reuse but is the control of an odorous or toxic vapor, scrubbers or incinerators are used [86]. Typical spray dryer systems equipped with thermal and catalytic incinerators are shown in Figure 45.7.

Cyclones, bag filters, and wet scrubbers are commonly used to control particulate emissions as in other industries. However, since the permitted levels of particulate emissions from dryers in the pharmaceutical industries have been greatly reduced in line with the increasing demands for environmentally clean operations, new design layouts have been developed to meet these requirements and legislations. These systems use high-efficiency cyclones and bag filters in series. Expected particulate emission levels from spray dryers for various system layouts are listed in Table 45.20.

Specialized designs have been developed to solve emission control problems connected with conventional methods in the pharmaceutical industry. Among these is closed-cycle drying, which employs recycle of the drying medium. Closed-cycle drying is recommended if

1. Flammable, toxic, or valuable solvents are used
2. No atmospheric pollution due to odor, solvent vapor, or powder emissions is permitted
3. Powder-air explosion/fire risks exist

Figure 45.8 and Figure 45.9 show flow sheets for closed-cycle spray and flash dryers.

Niro A/S have developed a special design for spray drying aqueous feeds that generate either unpleasant odors during drying, or consist of toxic materials, and/or form potential explosive powder-air mixtures within the dryer [90]. This drying system features

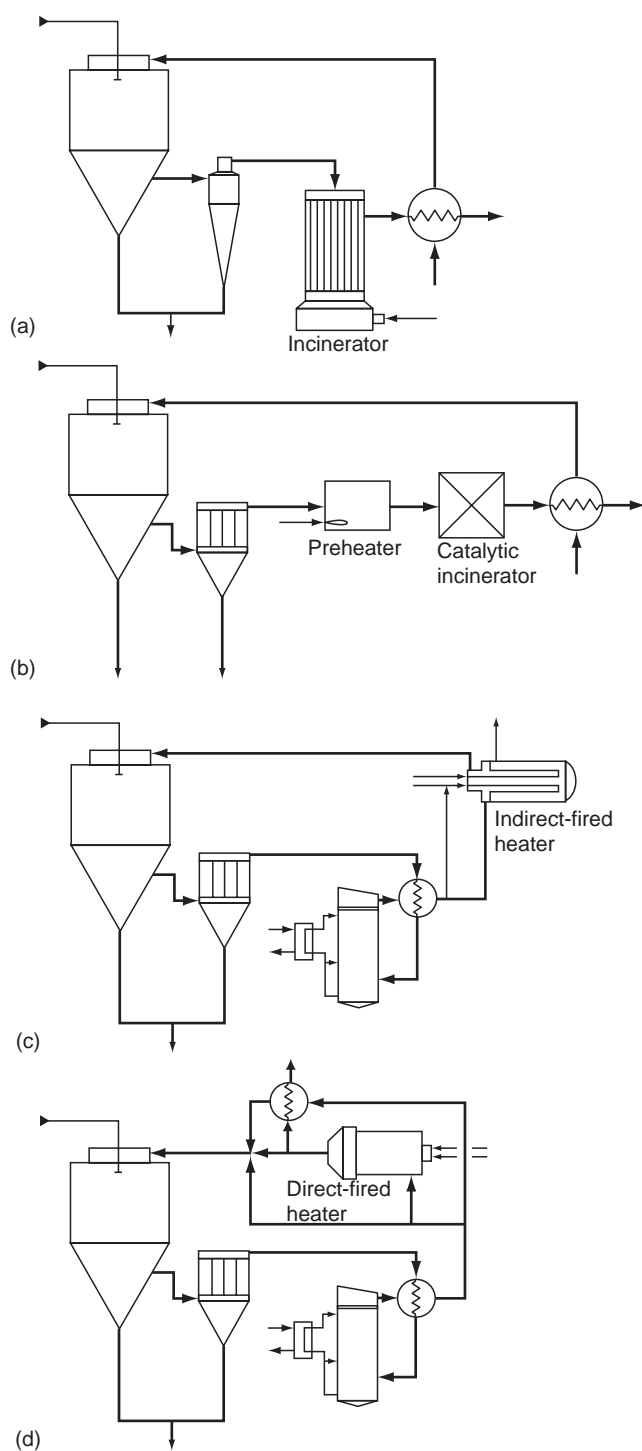


FIGURE 45.7 Spray dryer systems for toxic and odorous materials emission control: (a) open-cycle thermal incineration mode; (b) open-cycle catalytic incineration mode; (c) recycle thermal incineration mode where indirect dryer air heating is essential; and (d) recycle thermal incineration mode where direct dryer air heating is acceptable. (Courtesy of Niro, Inc.)

built-in deactivation (incineration) and heat recovery stages to eliminate residual active/toxic material emissions, while recovering the necessary incineration heat

TABLE 45.20
Particulate Emissions for Various Spray
Dryer Layouts

Plant Layout	Particulate Collector	Particulate Emission Levels, mg/Nm ³
Open-cycle system	Cyclone	>250
	Special cyclone or vortex flow separator	50–250
Open-cycle system	Cyclone + scrubber	20–100
	Special cyclone + scrubber	5–50
Open-cycle system	Bag filter	<20
Open-cycle system	Cyclone + bag filter	0.1–5
Open-cycle system	Bag filter + absolute filter	0.1
Recycled systems	All types	Negligible

Source: From K. Master, *Spray Drying Handbook*, Longman Scientific & Technical, New York, 1991 and A/S Niro Atomizer, *Spray Drying Systems for Meeting Environmental Standards*, Bulletin No. 56, A/S Niro Atomizer, Copenhagen, Denmark. With permission.

for reuse in the spray dryer. A flow sheet is shown in Figure 45.10 to illustrate this drying system.

45.4.4 FOOD AND AGRICULTURE INDUSTRY

45.4.4.1 Grain Handling, Food, and Feed Processing [75,91–97]

Grain handling and processing facilities move grain from the farm through storage and transfer locations (grain elevators), to the mills and plants that generate

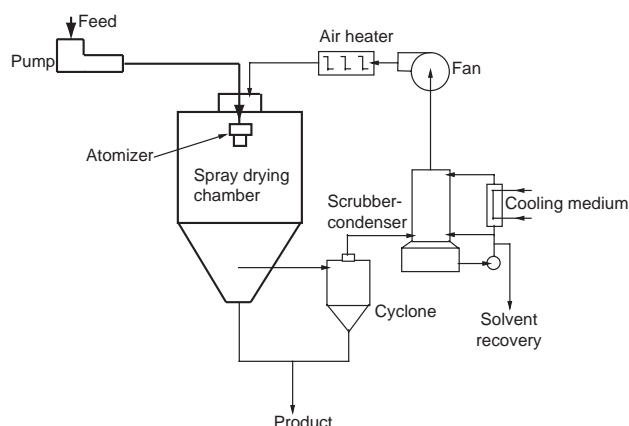


FIGURE 45.8 Closed-cycle spray drying system. (Reproduced by permission of Niro Inc.)

a variety of products (e.g., flour, starch, oil, animal feed) that can be used as food for the humans and feed for the animals.

Many grain elevators, grain mills, and feed manufacturing plants use grain dryers to reduce the moisture content of the grain. In the food and feed industries, whenever a powder or dry crystalline product is desired as a result of a wet process, the final operation is drying. Dried foods are easy to transport and store since microbial growth is controlled because free water is not available for growth. Thus, quality and nutritional values are retained during storage. A wide variety of dryers are used in the food and feed industries (e.g., spray, flash, fluid-bed, rotary).

The main pollutant of concern in grain handling and processing facilities is PM. Also, direct-fired grain drying operation may emit small quantities of

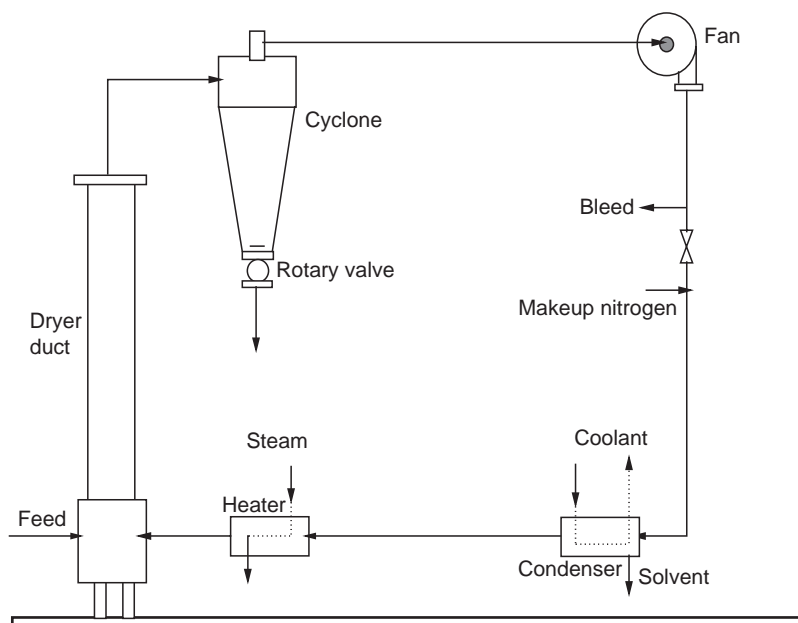


FIGURE 45.9 Closed-cycle flash drying system.

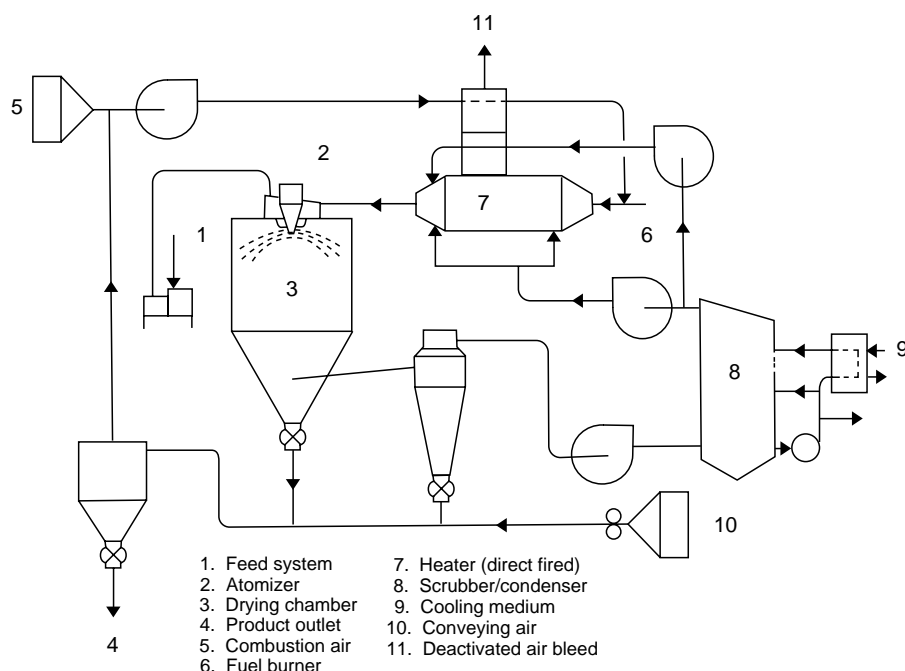


FIGURE 45.10 Spray drying plant, environmental layout. (Courtesy of Niro, Inc.)

VOCs and other combustion products. Grain dryers present a difficult problem for air pollution control because of the large volumes of air exhausted from the dryer, the large cross-sectional area of the exhaust, the low specific gravity of the emitted dust, and the high moisture content of the exhaust stream. The emission of PM from grain dryers is primarily dependent upon the type of grain, the dustiness of the grain, and the dryer configuration (emission factor = 3.0 lb/t in rack dryers and 0.22 lb/t in column dryers). Cross-flow column dryers have a lower emission rate than rack dryers because some of the dust is trapped by the column of grain. In rack dryers, the emission rate is higher because the turning motion of the grain generates more “bees wings” (a light flaky material that breaks off from the corn kernel during drying and handling). The EPA’s New Source Performance Standards (NSPS) for grain elevators established visible emission limits for grain dryers by requiring 0% opacity for emissions from column dryers with column plate perforations not to exceed 2.4 mm diameter or rack dryers with a screen filter not to exceed 50 mesh openings [94].

The PM control devices typically used in the grain dryers are screens, cyclones, and fabric filters. In the food and feed industries, cyclones, baghouses, and wet scrubbers are used to control solid particulate emissions. However, as with any organic dust, food processing dusts also have a high explosion tendency. Hence, special care is required to select, size, and operate the control devices. Emission factors for various dryer types utilized in alfalfa dehydration plants

are given in Table 45.21. Table 45.22 shows the filterable PM emission factors developed from available data on several source/control combinations for corn wet milling operations.

Flash drying systems, as applied to the feed industry, are usually designed to accomplish drying with disintegration. During drying, the airborne contaminants that may be emitted are dusts, vapors, and odors. The nature of the emissions is determined by the material being dried and the operating conditions of the drying system. Specifications were written to

TABLE 45.21
Emission Factors for Alfalfa Dehydration Plants

Source	Particulate Matter (PM)	
	Filterable PM, lb/t	Condensible PM, lb/t
Triple-pass dryer cyclone:		
Gas-fired	4.8	1.0
Coal-fired	7.5	ND
Single-pass dryer cyclone:		
Gas-fired	4.1	0.65
Wood-fired	3.1	1.3

ND = No data.

Source: From U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 9.9.4—Alfalfa Dehydrating, EPA, Research Triangle Park, NC, 1996. With permission.

TABLE 45.22
Particulate Matter (PM) Emission Factors for Corn Wet Milling Operations

Source	Type of Control	Filterable PM Emission Factor, lb/t
Gluten feed drying:		
Direct-fired rotary dryers ^a	Cyclone	0.27
Indirect-fired rotary dryers ^a	Cyclone	0.49
Starch drying:		
Flash dryers ^b	Wet scrubber	0.59
Spray dryers ^b	Fabric filter	0.16
Gluten meal drying:		
Direct-fired rotary dryers ^a	Cyclone	0.27
Indirect-fired rotary dryers ^a	Cyclone	0.49
Fiber, germ, and dextrose drying	ND	ND

^aEmission factors based on weight of PM, regardless of size, per unit weight of gluten meal or gluten feed produced.

^bEmission factors based on weight of PM, regardless of size, per unit weight of starch produced.

ND = no data

Source: From U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 9.9.7—*Corn Wet Milling*, EPA, Research Triangle Park, NC, 1995. With permission.

deal primarily with the odor problem. Both scrubber systems and thermal combustion systems are used [75]. In general, the higher the air velocities used in the dryer, the greater is the level of particulate carryover in exhaust.

Food products such as milk powder, coffee, tea, corn syrup solids, starches, potatoes, eggs, cheese, fruit and vegetable powders, and so on are usually spray dried. All dryers can be fitted with cyclone collectors to collect the fine particles that do not settle in the dryer chamber. If necessary, the cyclone can be backed up with a fabric collector unless the powder is hygroscopic, in which case a wet scrubber may be used. Emission factors for cheese drying and whey drying in natural and processed cheese manufacture are shown in Table 45.23.

45.4.4.2 Fish Processing [98,99]

The processing of fish is divided into two main categories. Freshly caught fish such as salmon and tuna are either frozen or canned for human consumption. The by-products from these operations are processed by fish meal plants and the products from these plants are used for pet food and animal feed supplements.

The fish meal process typically is a continuous operating system including cooking, pressing, drying,

TABLE 45.23
Particulate Emissions for Natural and Processed Cheese Manufacture^a

Source	Pollutant	Average Emission Factor, lb/t
Cheese dryer	Filterable PM	2.5
	Condensable inorganic PM	0.29
	Condensable organic PM	0.44
Whey dryer	Filterable PM	1.24
	Condensable PM	0.31

^aEmission factors for cheese dryers represent average values for controlled emissions based on wet scrubbers or venturi scrubbers. Factors for whey dryers are average values for controlled emissions based on cyclones, wet scrubbers, or fabric filters. PM = particulate matter.

Source: From U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 9.6.1—*Natural and Processed Cheese*, EPA, Research Triangle Park, NC, 1997. With permission.

cooling, storage, and shipping. Drying is used to reduce the moisture content (or water activity) of the fish meal in order to prevent microbial decomposition. Two main types of dryers are used: direct-fired rotary units and indirect rotary dryers.

The largest odor emission source in fish meal plants normally is the dryer, especially direct-fired dryers that require significant volumes of air for moisture removal. These dryers contribute about 60 to 80% of the total odor emission from the fish meal process. Table 45.24 illustrates the odor emissions from several rotary direct-fired dryers without odor control. Direct-fired dryers also emit smoke and particulate. It should be noted that the exhaust from direct-fired dryers normally averages about 95°C and its moisture content ranges between 15 and 25% by volume. PM emission factors for steam tube dryers and direct-fired dryers are 5.0 and 8.0 lb/t, respectively [99].

Typical dryer exhaust emissions are sulfur compounds such as hydrogen sulfide, carbon disulfide, carbonyl sulfide, and methyl and *n*-propyl mercaptans. In addition to ammonia, the only amine present is trimethyl amine. Since the emissions from the dryers contain considerable moisture at temperature of about 95°C, necessary means should be provided to remove most of this moisture and to cool the air before further odor treatment. Also, there may be dust particles in the cyclone exhaust that should be removed before effective odor measures can be applied. This is normally accomplished by either direct or indirect contact (e.g., shell and tube) water-cooled condensers. The direct-contact type includes co-current flow venturi scrubbers and countercurrent

TABLE 45.24
Typical Odor Emissions from Rotary Fish Meal Dryers

Feed Rate, t/h	Type of Fish	Temperature of Dryer Discharge, °C	Exhaust Gas Volume, (sm ³ /min)	Odor Units/sm ³
10	Tuna	105	524	52 972
15	Mackerel	105	524	52 972
70	Tuna	105	255	24 720
10	Tuna	115	283	52 972
14	Tuna	150	227	141 258
9	Tuna	95	480	88 286

Source: From W. Prokop, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992 and U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 9.13.1—Fish Processing, EPA, Research Triangle Park, NC, 1995. With permission.

flow spray-type scrubbers. In addition to condensing water vapor and cooling the gas, a specific degree of odor reduction is obtained with direct-contact condensers since certain odorous compounds are quite water soluble (Table 45.25).

Two basic methods of odor control are applied to emissions from dryers after condensing the water vapor: boiler incineration by direct-flame oxidation and wet scrubbing by chemical oxidation or the use of other scrubbing agents. Incineration provides the most positive control of nuisance-causing odorous compounds. Chlorinator scrubbers have been found to be 95 to 99% effective in controlling odors from fish cookers and dryers [99]. Table 45.26 shows the odor reduction obtained by boiler incineration.

Recently, new drying technologies have been developed for the drying of cooked and pressed fish (e.g., superheated steam fluid-bed dryer) (Figure 45.11) [100], which leads to nonpolluting and safe drying at low energy consumption. In this, the fluid-bed dryer operates under pressure by circulating steam in a closed vessel. Water evaporated from the feed material gener-

ates steam. As the drying takes place in a closed system, no pollution by fine particles or odor emission to atmosphere will occur. Steam-dried products are easier to digest and in many cases have a higher nutritive value.

45.5 NOMENCLATURE

45.5.1 CYCLONES

<i>a</i>	cyclone gas entry height, m
<i>A</i>	inside surface area of cyclone, m ²
<i>b</i>	cyclone gas entry width, m
<i>B</i>	cyclone dust outlet diameter, m
<i>c_i</i>	inlet dust concentration, g/m ³
<i>c_o</i>	outlet dust concentration, g/m ³
<i>d_p</i>	particle diameter, m
<i>d</i>	diameter of cyclone cone at the natural length, m
<i>D</i>	cyclone cylinder diameter, m

TABLE 45.25
Odor Removal Ability of Direct Condensers

System	Odor Units/1000 sm ³		Percent Removal	Air Temperature, °C	
	Inlet	Outlet		Inlet	Outlet
1	4875	670	86	—	—
	635	127	80	92	35
2	1835	635	65	79	28

Source: From W. Prokop, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992 and U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 9.13.1—Fish Processing, EPA, Research Triangle Park, NC, 1995. With permission.

TABLE 45.26
Odor Removal Ability of Boiler Incineration

System	Odor Units/1000 sm ³		Percent Removal
	Inlet	Outlet	
1	5085	200	96
	635	30	95
2	168	30	82
	127	21	83
3	635	42	93
	318	35	89

Source: From W. Prokop, in *Air Pollution Engineering Manual*, A. Buonicore and W. Davis (Eds.), Van Nostrand Reinhold, New York, 1992 and U.S. Environmental Protection Agency (EPA), *Compilation of Air Pollution Emission Factors (AP-42)*, Section 9.13.1—Fish Processing, EPA, Research Triangle Park, NC, 1995. With permission.

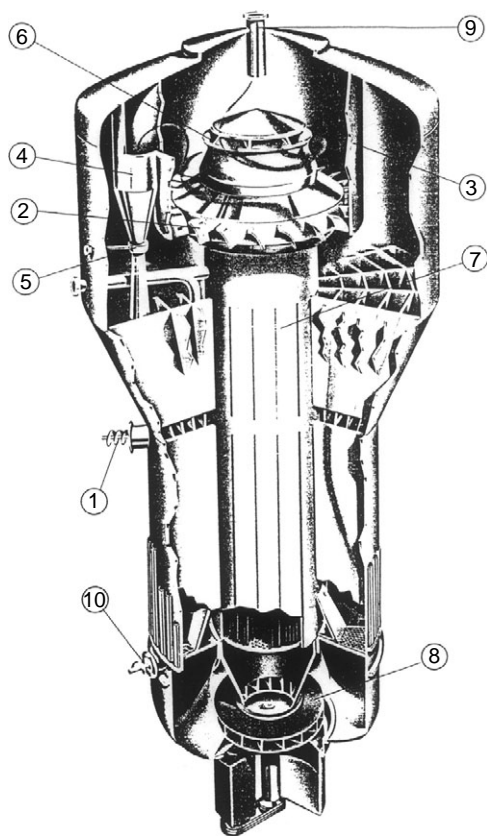


FIGURE 45.11 Superheated steam fluid-bed dryer: (1) screw feeder; (2) stationary blades; (3) cylinder; (4) cyclone; (5) ejector; (6) vanes; (7) heat exchanger; (8) impeller; (9) top outlet; and (10) screw conveyor. (Courtesy of Niro, Inc.)

D_e	cyclone gas outlet diameter, m
g	acceleration of gravity, m/s^2
G^*	friction factor, 0.005
h	cyclone cylinder height, m
H	cyclone overall height, m
K	cyclone geometry parameter
l	natural length, m
n	vortex exponent
N_e	number of turns gas makes within cyclone
N_H	number of inlet velocity heads
ΔP	pressure drop, N/m^2
S	cyclone gas outlet height, m
St_{kc}	stokes number
St'_{kc}	stokes number
u_i	cyclone gas inlet velocity, m/s
u_s	saltation velocity, m/s
u_T	tangential gas velocity, m/s
V_s	annular volume around the exit duct from S up to half the inlet height $a/2$, m^3
V_{nl}	annular volume around the central core from S down to the natural length l , m^3
η_i	grade efficiency
η	overall efficiency
μ	gas viscosity, $\text{Pa} \cdot \text{s}$
ρ_g	gas density, kg/m^3
ρ_p	particle density, kg/m^3

ϕ ratio of maximum tangential gas velocity to velocity within gas entry

45.5.2 WET SCRUBBERS

C'	Cunningham slip correction factor, dimensionless
d_c	nominal packing diameter, cm
d_h	diameter of sieve plate hole, cm
d_{mo}	molecular diameter, cm
d_{ps}	physical particle diameter, m
d_{pa}	aerodynamic diameter of particle, cm
d_{pc}	performance cut diameter, $\text{m}\mu$
f	empirical factor
f_v	volume fraction voids
F	foam density, g/cm^3
k	Boltzmann constant = 1.38062×10^{-23} J/K
K_p	inertial impaction parameter, dimensionless
K_{po}	inertial impaction parameter, dimensionless
l_t	venturi throat length, cm
n_h	number of holes
P	gas pressure, kPa
ΔP	pressure drop, N/m^2 , or cm water column (WC)
ΔP_{dry}	dry plate pressure drop, cm WC
ΔP_{wet}	wet plate pressure drop, cm WC
p_t	penetration, fraction
Q_G	gas volumetric flow rate, m^3/s
Q_L	liquid volumetric flow rate, m^3/s
r_d	drop radius, cm
T	gas temperature, $^\circ\text{C}$
u_G	gas velocity relative to duct, cm/s
u_{Gs}	superficial gas velocity, cm/s
u_h	gas velocity through sieve plate hole, cm/s
u_r	drop velocity relative to gas, cm/s
u_t	drop terminal settling velocity, cm/s
Z	height of packing
η_d	efficiency for single drop
λ_g	mean free path of gas molecules, cm
μ_G	gas viscosity, $\text{g/cm} \cdot \text{s}$
μ_L	liquid viscosity, $\text{g/cm} \cdot \text{s}$
ρ_L	liquid density, g/cm^3
ρ_p	particle density, g/cm^3
σ	surface tension, dyne/cm

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46 Energy Aspects in Drying

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46.1 INTRODUCTION

It has become apparent in recent years that energy resources, especially natural gas and oil, are limited. Consequently, all industrial sectors in all parts of the world need to identify more efficient methods of energy utilization. Despite periodic fluctuations, there is a tendency for energy costs to increase and, consequently, in many cases energy should become an important element of innovations to drying practice and changes in equipment technology. Many studies, publication, and monographs are concerned with

optimized energy utilization; most of these refer to a specific industry (e.g., chemical, ceramic, and metallurgical). However, a process-oriented approach would be more justified, as it would be based on a particular equipment or operation regardless of the branch of industry.

In the case of drying, it is worth noting that in many technologically feasible cases the initial moisture content of a product can be reduced by using other methods that are more energy efficient. For example, water removal from solids by mechanical devices such as presses or centrifuges and, in the

case of liquids, the reuse of the heat in a multistage evaporator results in energy consumption being appreciably lower overall. During the past decades, the rise of energy prices was accompanied by increasingly stringent legislation on pollution, working conditions, and safety. To meet these requirements as well as to optimize energy consumption, new ideas in drying methods and dryer design are required. Considering the problems encountered in drying (Table 46.1), it is obvious that this is not an easy task and, because energy aspects in drying is a broad subject, this chapter is limited to the following points:

- General considerations
- Methods of heat recovery
- Nonconventional energy sources (e.g., microwaves, radiofrequency [RF], and induction)
- New drying techniques
- Control and monitoring of the drying process
- Environmental problems.

46.2 GENERAL CONSIDERATIONS

The potential for energy reduction can be considered from several points of view, the most important being:

- *Methodological*: Development of general methods for reducing energy consumed
- *Technological*: Utilization of secondary energy sources, reduction of heat losses to the atmosphere, and the like
- *Socioeconomic*: Stimulation of search for potential energy sources
- *Organizational*: Coordination of processes aimed at an economical use of energy

- *Ecological*: Focusing attention on solutions safe for the natural environment:

The general principles and methods of energy conservation have been outlined in a number of comprehensive monographs [1–5]. In the past few years, many articles have been published concerning the potential for energy conservation in industrial drying. The importance of energy saving is so great that this topic needs not only an up-to-date, comprehensive literature review, but also a new and generalized approach to the dryer as energy-consuming equipment.

Industrial dryers consume a significant part of the total energy used in manufacturing processes, 12% on the average. For example, the energy used for drying in selected U.K. industries was estimated to be 128×10^9 MJ/y [6].

Table 46.2 presents the overall pattern of energy usage for drying in France [7] and the United Kingdom. The estimated total energy used for drying in the United States amounts to about 1600×10^9 MJ/y [8]. Hallström [9] reported a similar result for the Swedish chemical industry, for which he found that 15% of the total energy was used for drying.

Per capita water evaporation per year is estimated to be [10]:

United States	1090 kg
Great Britain	800 kg
Czechoslovakia	630 kg

Many industrial users of dryers ignore energy efficiency on the grounds that product quality and throughput are paramount. However, a proper balance between these parameters should be maintained.

The process of dryer design and improvement must be constantly reviewed since many external

TABLE 46.1
Problems Encountered in Drying

Mode of Heating	Conduction, Convection, Dielectric, and Radiation
Energy sources	Coal, oil, natural gas, electric energy, waste materials, and solar energy
Pressure	High vacuum (freeze drying), vacuum, atmospheric pressure, and high pressure
Scale	From 10 kg/h to >100 t/h
Type of apparatus	About 100 types of identified dryers are used in the world at present
Process rate limitation	Boundary layer, inner diffusion, and boundary layer with inner diffusion
Size and shape of material	Powder, granules, foil, film, and plate
Material properties	Nonporous, capillary-porous, and hygroscopic; solution, syrup, mud, gel, extruded material, crystalline, fabric, cardboard, and fiber
Initial moisture content	From almost dry (<1% kg/kg dry material) to full saturation (>100% kg/kg dry material)
Thermal resistance of the material	From material very sensitive to temperature (<30°C) to thermally resistant materials (>200°C)
Value of the product	From bulk chemical products (<\$70/t of dry material) to pharmaceutical products (>\$150,000/t)

TABLE 46.2
Overall Pattern of Energy Usage for Drying

Subsector	French Industry Drying (10 ⁹ MJ/y)	British Industry		% Due to Drying
		Drying (10 ⁹ MJ/y)	Total (10 ⁹ MJ/y)	
Food and agriculture	46.3 ^a	35	286	12
Chemicals	8.6	23	390	6
Textiles	1.9 ^a	7	128	5
Paper	38.8	45	137	33
Ceramic and building materials	15.7	14	127	11
Timber	7.9 ^a	4	35	11
Others	50.3	No data	—	—
Total	≈168	128	1103	12

^aAdded thermal energy (tons of oil equivalent) and electricity (GWh), extracted from original data.

Source: From A. Larreture and M. Laniau, The State of Drying in French Industry, *Drying Technology*, 9(1):263–275 (1991); C.G.J. Baker and D. Reay, Energy Usage for Drying in Selected U.K. Industrial Sectors, *Proceedings of 3rd International Drying Symposium*, 1:201–209 (1982).

factors may change in the dryer's "life." For example, changes in the properties of the raw materials and the final products, changes of solvents, and requirements in both the plant and the general environment concerning devices for dust removal may be the factors. It is further affected, perhaps most importantly, by the availability and cost of energy.

46.2.1 INTENSIFICATION OF HEAT AND MASS TRANSFER

The enhancement of a process affects the energy savings mainly due to:

- The shortening of process time and machine length, resulting in the reduction of energy losses within the shell unit of production
- Supplying of heat precisely to the place at which it is needed and, as a result, achieving the reduction of energy wasted in heating peripheral equipment
- A decrease of drying agent consumption.

In many practical cases, the external and internal heat and mass transfer are the main influences on drying process intensity [11]. Having this in mind, it is important to carry out these transport processes properly, as for the two cases below:

1. In the case of external transport (from the wet material surface to the drying agent), the intensification of heat and mass transfer can be achieved due to:

- An increase in the differences in temperature and concentration of the substance evaporated in the vicinity and at the surface of the material
 - Increase in flow rate and turbulence of the drying medium, leading to rapid changes of flow direction, which in turn result in boundary layer renewal
 - Changes in the properties of the drying medium or its condition by, e.g., replacing air with an inert gas or by carrying out the process under lower pressure
 - The application of electromagnetic wave energy (e.g., infrared and high frequency)
2. In the case of heat and mass transport inside the material being dried, intensification can be achieved by:
 - Changes in the material structure prior to or in the course of drying (if this is technologically feasible)
 - Addition of surface-active substances to the material that can enhance transport processes
 - Control of temperature, moisture content, and pressure distributions in the material by, e.g., varying the rate of energy transfer with time.

In the majority of cases, the most efficient approach to drying intensification is likely to be a combination of several methods. Among the methods for enhancement of heat and mass transfer applied in drying, the following should be considered: (1) mixing of the material during drying; (2) periodic changes in the drying agent flow to the material; (3) application of high temperatures; and (4) changing conditions,

e.g., alternate heating and cooling of the material [12–17]. Arising from the continuous changes of temperature in the region of the material surface, as well as from the flow pulsation [15], heat transfer coefficients can be 1.5 to 3 times greater than under fixed conditions.

It has been shown that a temperature increase from 140 to 180°C in textile industrial dryers resulted in a reduction in energy consumption of 25% [16] and impulse drying of building and ceramic materials resulted in 30% reduction [17].

46.2.2 ENERGY EFFICIENCY OF THE DRYING INSTALLATIONS

The energy performance of a dryer and a drying process is characterized by various indices such as volumetric evaporation rate, surface heat losses, steam consumption, unit heat consumption, and energy (thermal) efficiency. Of all these indices, the energy efficiency η , is most frequently quoted in technical specifications. It commonly relates the energy used for moisture evaporation at the solids feed temperature (E_{ev}) to the total energy supplied to the dryer (E_t):

$$\eta = \frac{E_{ev}}{E_t} \quad (46.1)$$

For low humidity and low temperature convective drying when the specific heat capacities are constant, the energy efficiency can be approximated by thermal efficiency, which is the function of the inlet air temperature (T_1), the outlet air temperature (T_2), and the ambient temperature (T_0):

$$\eta_T = \frac{T_1 - T_2}{T_1 - T_0} \quad (46.2)$$

Equation 46.2 shows that energy efficiency η_T could vary from 0 (for $T_2 = T_1$) to 1 (for $T_2 = T_0$).

Accepting that the outlet air temperature can not be lower than the material temperature at any instant of a drying process, it may be concluded that there is a maximum efficiency determined by the equation:

$$\eta_{Tmax} = \frac{T_1 - T_{wb}}{T_1 - T_0} \quad (46.3)$$

where T_{wb} is the wet bulb temperature corresponding to the outlet air conditions.

A maximum efficiency η_{Tmax} vs. T_1 (at different values of T_0 and absolute humidity of ambient air) for the simplest case of a single pass of air through the dryer is shown in Figure 46.1 [18]. From Equation 46.2 and Equation 46.3 it is clear that the energy efficiency de-

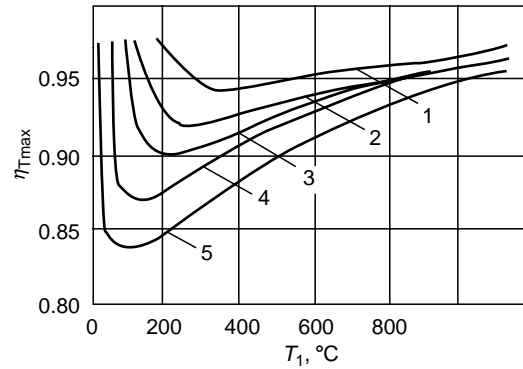


FIGURE 46.1 The effect of T_1 on η_{Tmax} of the single pass of drying agent in the dryer: (1) $T_0 = 40^\circ\text{C}$, $Y_0 = 0.010$ kg/kg; (2) $T_0 = 30^\circ\text{C}$, $Y_0 = 0.000$ kg/kg; (3) $T_0 = 30^\circ\text{C}$, $Y_0 = 0.010$ kg/kg; (4) $T_0 = 30^\circ\text{C}$, $Y_0 = 0.020$ kg/kg; and (5) $T_0 = 20^\circ\text{C}$, $Y_0 = 0.010$ kg/kg where Y_0 denotes ambient air humidity.

pends on the inlet and outlet temperatures of the drying agent, ambient temperature, and air humidity; high energy efficiencies are attained by the use of high inlet temperatures and by arranging the outlet air conditions close to saturation. Hence, the same rotary dryer, e.g., operated at low temperature, will have lower energy efficiency than the one operated at higher temperature. The energy efficiency also depends on heat fluxes supplied and lost, the number of internal heating zones, air and material recirculation, and so on. An extensive review of various expressions for the energy efficiency along with parametric analysis of various drying conditions is presented by Ashworth [21].

The energy efficiency is usually regarded as a cumulative parameter, calculated from the initial–final or inlet–outlet data. For batch drying, the energy efficiency is therefore given as an average value over a drying time, while for continuous drying the energy efficiency is averaged over the range of moisture content or the dryer length. However, the energy efficiency depends on the material moisture content at any instant because thermal energy is used not only for evaporation of free moisture, but also for the removal of bound moisture and eventually breaking the material–moisture bonds [19]. To account for variation of the energy efficiency with material moisture content Kudra [20] proposed the concept of an instantaneous energy efficiency defined as:

$$\varepsilon_E = \frac{\text{energy used for evaporation at time } t}{\text{input energy at time } t} \quad (46.4)$$

Integration of the instantaneous energy efficiency gives the cumulative energy efficiency over a given time interval which is equivalent of the moisture content range:

$$\bar{\varepsilon}_E = \frac{1}{t} \int_0^t \varepsilon_E(t) dt \quad (46.5)$$

The cumulative energy efficiency is identical with the one given by Equation 46.2 if the outlet air temperature is constant, or the time interval is short and the moisture content variation is not significant. Conversely, Equation 46.2 can be used to calculate the instantaneous energy efficiency from temperature data given in terms of temperature if the variation of the outlet air temperature with time or moisture content is known. Furthermore, the instantaneous energy efficiency, suitable for dryer performance analysis, can be determined from Equation 46.1 if values of E_{ev} and E_t are taken for small time increment.

The energy efficiency defined by Equation 46.1 through Equation 46.4 indicates the degree of energy utilization. Departure of the real drying process from the best possible trajectory from the energy point of view can be quantified by the drying efficiency given by the following equations, which account for enthalpy of the exhaust air:

$$\varepsilon_D = \frac{\text{energy used for evaporation at time } t}{(\text{input energy} - \text{output energy with outlet air}) \text{ at time } t} \quad (46.6)$$

$$\text{and } \bar{\varepsilon}_D = \frac{1}{t} \int_0^t \varepsilon_D(t) dt \quad (46.7)$$

From Figure 46.2 it is evident that both energy and drying efficiencies depend on the solid moisture content although their absolute values differ significantly. A distinct maximum for the instantaneous drying efficiency can be attributed to the drying process being controlled by internal moisture diffusion. In the initial drying period, a significant fraction of input energy is utilized for material heating. As drying proceeds, the sensible heat in a drying agent is used for evaporation of nonhygroscopic water near the material surface. When this water is evaporated, the instantaneous drying efficiency falls rapidly because most of the heat is now utilized for overheating the already dry surface layers and removal of the microcapillary water (bound water moisture content for grains is about 22% wb, whereas the hygroscopic moisture content is about 36.5% wb).

The instantaneous and cumulative indices can be used to analyze performance of the dryer and thus identify avenues to reduce the overall energy consumption [20]. As seen from Figure 46.3, both instantaneous drying and energy efficiencies attain maximum values near the dryer inlet, which results from high drying rates due to evaporation of surface

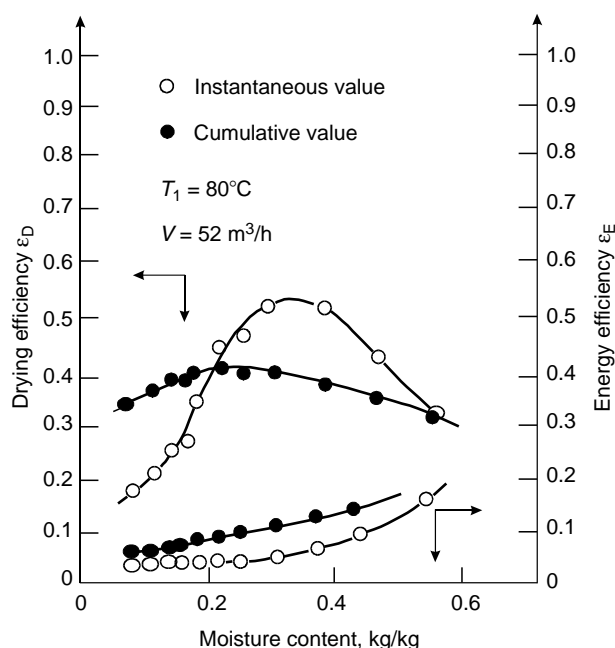


FIGURE 46.2 Instantaneous and cumulative drying and energy indices for fluid-bed drying of corn. (From T. Kudra, Instantaneous Dryer Indices for Energy Performance Analysis, *Inżynieria Chemiczna i Procesowa*, 19(1):163–72 (1998).)

moisture. In this part of a dryer, the outlet air is completely saturated, which practically limits the drying rate. Further along the dryer, the energy and drying efficiencies diminish dramatically as a result of the decreasing drying rate. Although temperature of the outlet air remains practically constant along the dryer, the relative humidity falls sharply because of the reduced evaporation rate. At the dryer outlet, where the material is practically dry, the air temperature starts to increase and the relative humidity drops down to 5%. Considering the pattern of energy and drying efficiency in the dryer, one can conclude that better drying indices can be achieved by reducing the air-flow rate at the dryer end. In practice, starting from 0.5 m from the dryer inlet, the internal mass transfer controls the drying rate, so reducing gas velocity may lower the energy consumption without significantly decreasing the drying rate. Since the outlet air is completely saturated at the dryer inlet, higher air-flow rates are recommended to supply additional heat for water evaporation and to increase the driving force for the mass transfer. The higher inlet air temperature in this part of the dryer can also be taken into account. In conclusion, the improved performance of this particular dryer both from drying rate and energy consumption points of view can be achieved by varying air-flow rate and air temperature

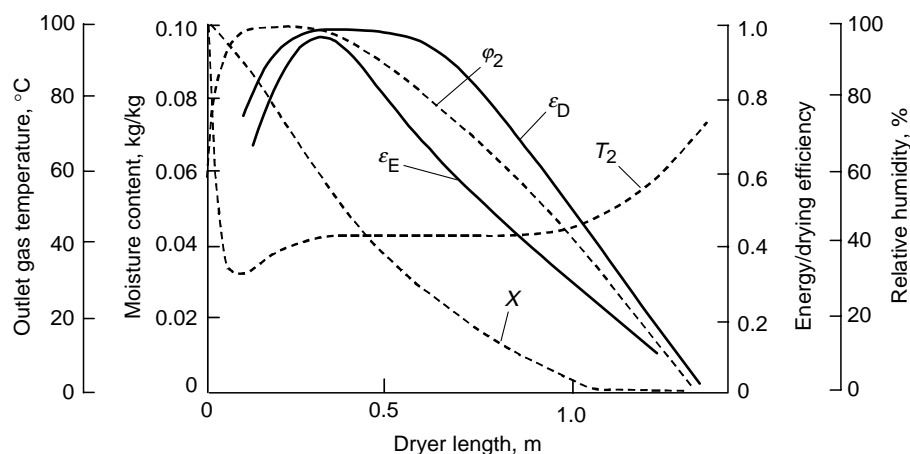


FIGURE 46.3 Performance characteristics of a vibrated fluid-bed dryer. (From T. Kudra, Instantaneous Dryer Indices for Energy Performance Analysis, *Inżynieria Chemiczna i Procesowa.*, 19(1):163–72 (1998).)

along the dryer length. This can be done by dividing the gas plenum into separate compartments fed with individual air streams. Temperatures and superficial air velocities of each air stream can be set upon maximizing instantaneous drying efficiency throughout the dryer.

For most convective dryers, the energy required for evaporation of moisture together with that lost in exit airstreams dominates the energy demand. Statistical data obtained from heat balances of convective dryers show that 20 to 60% of the heat supplied to the dryer is used for moisture evaporation, 5 to 25% for material heating, 15 to 40% for heat losses with the exhaust air, 3 to 10% for heat losses from dryer walls to the atmosphere, and 5 to 20% for other losses [18].

The measure of energy consumption in a drying process is the unit energy consumption for the evaporation of 1 kg of moisture (q). According to some authors [22], in drying chambers used for continuous timber drying with countercurrent circulation of the drying agent the value is 3000 to 4000 kJ/kg, while in batch dryers in general it ranges between 2700 and 6500 kJ/kg. During drying of thin and flexible layers, e.g., fabrics or paper, the value of q varies in the range from 5000 to 8000 kJ/kg. Comparable results are obtained for drying of fine particulate solids. Higher energy consumption occurs in industrial convective dryers for pastelike materials.

Theoretically, the amount of heat required for the evaporation of 1 kg of moisture under standard conditions is 2200 to 2700 kJ/kg. The upper limit of this value refers to the removal of bound moisture. The only drying regime in which such a result could be obtained is the adiabatic equilibrium in which there is no heating of a solid body and accompanying moisture, i.e., they enter at the evaporating temperature.

Furthermore, no additional heat is exchanged to the evaporated vapor and gas carrier and there is no heating of the dryer body and the ambient air.

However, in practice, the process has to be carried out at a realistic rate. So, a compromise between energy consumption, thermal efficiency, and drying rate is inevitable. Most engineers accept that inefficiencies of the drying process are inevitable and consequently they neglect possible energy savings arising from manageable losses.

A comparison of the theoretical consumption of energy q , referred to above, and the industrial practice shows that the studies on the reduction of energy consumption in dryers are needed. Baker and Reay [6] considered various techniques leading to the reduction of energy consumption. The most promising appeared to be heat recovery from the outlet gas stream as well as optimization of dryer design and of the entire drying process.

Pinch analysis as a tool of process integration could be applied here as in most cases drying and all its downstream and upstream processes are necessarily bound to form a hierarchical structure, adequately represented by the “onion” model [23].

However, the first step in any energy-saving program is to balance material and energy flow to and from the dryer. This often makes it possible to reveal areas in which changes of operating conditions could result in a significant improvement in the efficiencies of the dryer operation and can bring about energy savings amounting to 10 to 30%. The heat balance measurements required include airflow rate, temperature, and humidity, as well as material moisture content at various points of the drying system. An example of the installation of measuring points is given in Figure 46.4, showing a fluidized bed dryer.

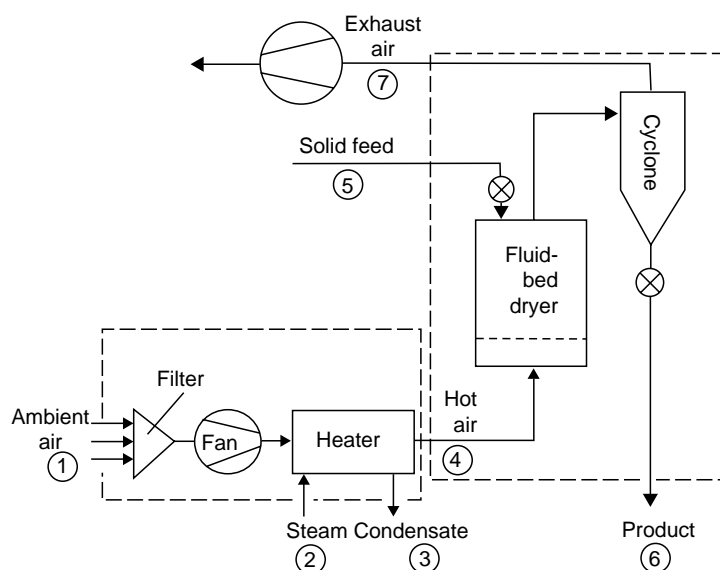


FIGURE 46.4 Measured parameters for testing of the drying system: (1) ambient airflow rate, temperature, humidity; (2) supply steam flow rate, temperature, pressure; (3) condensate temperature; (4) process air temperature; (5) feed solids moisture content, temperature; (6) product solids flow rate, moisture content, temperature; and (7) exhaust airflow rate, temperature, humidity.

Success in preparing the balance of this type depends to a large extent on a proper selection of measuring points in a given drying system and the use of reliable measuring devices. Steady state conditions of the dryer operation should be established before starting the measurements.

The calculations of energy balances arising from such measurements provide a basis for estimating potential energy savings using various options. These options can be divided into two categories: (1) those that require little or no capital and (2) those that need significant investment (Table 46.3 and Table 46.4) [24].

As seen from Table 46.5 [24], there are significant possibilities for reducing energy consumption at a very little capital expenditure.

The approach described above and the data presented by Bahu et al. [24] could be applied to the estimation of energy efficiency of many industrial

dryers. The data obtained allow selection of the most effectual methods of energy savings, provide solutions concerning other aspects of dryer operation (i.e., overcoming restrictions of the dryer output), as well as provide valuable information for new dryer design.

The experience gathered from such an analysis of balances for numerous dryers shows the vast possibilities of design and construction improvements for dryers from the viewpoint of energy savings. In all cases considered, at least 10% savings were possible due to the use of an accurate analysis of drying alone and reached 60 to 70% following modifications to the dryer. It was surprising to find that the data concerning drying kinetics or equilibrium of the material being dried were not available. According to Baker and Reay [6], the most promising options for energy savings can then be selected (Table 46.6).

Results of an extensive study on industrial drying [8] revealed that the drying efficiency of the 17 types of dryers ranged from 20% for the continuous direct

TABLE 46.3
Schemes Involving Little or No Capital Expenditure

Reduction of air leaks
Change of product moisture content
Change of dryer air inlet temperature
Change of dryer mass flow rate
Improvement of dryer insulation
Improvement of heater insulation

TABLE 46.4
Schemes Involving Significant Capital Expenditure

Exhaust gas recycle
Use of exhaust gas to preheat inlet air
Product cooler
Change from indirect to direct heating

TABLE 46.5
Summary of Energy-Saving Schemes

Scheme	Cumulative Energy Saving, kW	Specific Energy Consumption, kg Steam/kg Product	Saving Over Existing Dryer, %
Existing dryer	—	0.53	—
<i>Minor capital expenditure</i>			
1. Rectify internal steam leak on heater	39	0.46	13
2. Scheme 1 + reduce air leaks into dryer	54	0.44	18
3. Scheme 2 + improve insulation of heater ^a	79	0.40	26
4. Scheme 3 + improve insulation of dryer and reduce airflow rate	150	0.26	50
5. Scheme 4 + raise X_0 to 4% and reduce airflow rate	159	0.24	53
<i>Major capital expenditure</i>			
6. Scheme 5 + 60% exhaust air recycle	192	0.20	64
7. Scheme 5 + waste heat exchange with no condensation	178	0.22	59
8. Scheme 5 + waste heat exchange with condensation	214	0.17	71
9. Scheme 5 + switch to direct-fired heater ^b	183	0.21	61

^aBased upon arbitrary chosen reduction in the loss that does not necessarily correspond to the economic value.

^bEnergy saving is given on a primary fuel equivalent basis.

(convective) tower dryer to about 90% for some types of indirect (contact) dryers, e.g., cylinder, drum, rotary (bundle), and agitated pan dryers, where heat for moisture evaporation is supplied by conduction from heated walls, tubes, or vanes (Table 46.7). The drying efficiency in Table 46.7 is defined as “the ratio of evaporation actually obtainable to that theoretically possible from that portion of the energy supplied to the system which is available for evaporation,” and corresponds to the drying efficiency given by Equation 46.6. In case of no explicit definition, care should be taken not to confuse drying efficiency with the more widely used concept of energy efficiency (e.g., Equation 46.1 and Equation 46.2), which generally yields a lower value than the drying efficiency (Figure 46.2). The weighted average annual energy requirement for industrial solid dryers in the United States is shown in Table 46.7 [8]. Similar data based on the French industry are shown in Table 46.8 [7].

The analysis of the above tables shows that few types of dryers account for >90% of the total energy consumption of all the types considered. This conclusion, relatively old and approximate data, and only for U.S. and French industry, can be considered to be a very good pointer as to which type of dryers must be redesigned in order to achieve large overall energy savings.

When new equipment is being purchased, cost and throughput information taken in isolation is inadequate for the selection and proper utilization of

TABLE 46.6
Potential Energy Savings for Selected U.K. Industries

Option	Potential Energy Savings		Penetration Rating ^a
	10 ⁹ MJ/y	% Total	
Heat recovery from dryer exhaust (other than heat pump)	18.9	15	High
Heat pumps (closed cycle)	8.9	6	Medium
Vapor compression	26.2	20	Low
Better instrumentation and control	4.3	3	Medium/high
Optimization of dryer design and operation	11.0	9	High
Improved dewatering of feedstock	5.3	4	High

^aPenetration rate is a guess of the degree of penetration of the potential market for the development that will eventually be achieved.

TABLE 46.7
Weighted^a Average Annual Energy Requirements
and Drying Efficiency for Industrial Solid
Dryers—Based on Surveyed Data

Type of Dryer	Energy Requirement, 10 ⁹ MJ/y	Drying Efficiency, %
Direct continuous		
Tower	137 + 32	20–40
Flash	528 + 211	50–75
Sheeting (stenters)	2.8	50–90
Conveyor	1.9	40–60
Rotary (bundle)	66	40–70
Spray	9.5	50
Tunnel	<1	35–40
Fluidized bed	23	40–80
Batch		
Tray	<1	85
Indirect continuous		
Drum	2.4	85
Rotary	53	75–90
Cylinder	127 ÷ 53	90–92
Batch		
Agitated pan	<1	90
Vacuum rotary	<11	Up to 70
Vacuum tray	<1	—
Infrared	<1	30–60
Dielectric	<1	60
Total	1261	

^aWeighted averages were obtained in selective judgment based on dryer manufacturers' opinions.

Source: Richardson, A.S. and Jenson, W.M.P. *Aerojet-Nuclear Company Report* No. E (10-1)-1375, 1976.

a drying system; it is important that specific energy consumption should also be considered. In performing a drying test for this determination, it is necessary to specify certain test conditions in order to assess the dryer performance adequately. "It is also necessary to specify the type of a drying material and the range of moisture content as energy efficiency of a given dryer may drop considerably when drying materials with high resistance to internal mass transfer. An example is markedly lower (from 0.36 to 0.4) energy efficiency of the fluid bed, vibrated bed, and pulsed fluid-bed dryers when used for drying of osmotically dehydrated cranberries [25]."

There is a wide variety of drying materials and types of equipment, so it is not possible to propose a generalized method for a standardized determination of dryer efficiency. Therefore, standard conditions at which a particular dryer should be tested have to be specified. This ultimately would lead to establishing standards for rating dryer capacity and energy per-

TABLE 46.8
Relative Importance of the Different Equipment

Type of Dryer	Energy Consumption by Equipment 10 ⁹ MJ/y
Drum dryers (e.g., paper-cardboard and puree)	32.0
Rotary dryer (e.g., pulp and alfalfa and aggregates)	27.0
Bin dryer (e.g., corn and malt)	13.2
Spray dryer (e.g., milk and whey)	10.5
Flash dryer (e.g., starch and wood particles)	8.4
Tunnel dryer (e.g., plaster squares, plasterboards, and bricks)	7.6
Air floater dryer (e.g., paper coatings and paper pulp)	7.3
Tube bundle dryer (e.g., meat flour and fish flour)	4.4
Kiln (e.g., wood)	2.6
Fluidized bed dryer (e.g., PVC suspension)	1.97
Conveyor dryer (e.g., plywood panels and vegetables)	1.8
Drying tenter (e.g., fabric)	1.68
Total	118.5

Source: Larreture, A and Laniau, M., *Drying Tech.*, 9(1), 263–275, 1991.

formance. Such standards would allow dryer users to find out whether their dryers operate at or off the optimum conditions.

Some proposals to standardize the procedure for rating dryer capacity and energy efficiency benchmarking can be found in publications concerned with grain drying. Bakker-Arkema et al. [26,27] have proposed a standard index, the Dryer Performance Evaluation Index (DPEI). This is defined as the total energy required by a dryer to remove 1 lb of moisture from the grain during drying under a set of specified conditions. However, this concept has not been accepted by the grain dryer manufacturing industry because of the following reasons:

- Costs involved in accurate field testing are high due to lack of agreement on the standard field-testing conditions
- No procedure exists to account for corn hybrids and nonstandard conditions
- Energy efficiency rating is not (yet) mandatory.

Even within this rather specific range of dryers, to make valid comparisons possible, experimental and simulated data on each dryer need to be available. It

has been proposed that each dryer must be tested experimentally under conditions approximating a set of standard conditions. In the case of corn, these standard conditions are tabulated and the data to be determined for a dryer performance evaluation are listed by Keener and Glenn [28]. Thereafter, the effect of nonstandardized operating conditions or the varying nature of the corn on the capacity and energy efficiency can be assessed with the use of tables generated by computer simulation.

Because exhaustive experimental testing of every dryer model would be extremely expensive, simulation techniques based on a representation of a drying process (or a dryer) by a series of mathematical equations have been developed by Bakker-Arkema et al. [27,29–31].

The predicted or simulated energy efficiency of a dryer is always higher than the experimentally measured value. This is due to losses by radiation and convection to the atmosphere, nonideal fuel combustion, heat losses through leaks in hot air ducts, and other factors in the real situation. Losses vary among dryer types and among models of the same type. To account for losses, the dryer efficiency factor (DEF) can be introduced. This is the ratio of the experimentally measured energy efficiency to the simulated operation energy efficiency. The higher the DEF for a particular dryer type, the better is its construction. A test procedure for rating the capacity and the energy efficiency of dryers has been proposed [26]. The steps of the procedure should be as follows:

1. Conduct an experimental test under approximately standard conditions and determine the capacity, moisture content reduction, and energy efficiency of the dryer.
2. Carry out a simulation using the same standard conditions with respect to capacity and moisture content reduction in order to establish DEF of the dryer introducing a factor that takes into account the characteristic drying behavior of the material used for a particular test (the material drying factor).
3. Determine the effect of varying drying air temperature, ambient conditions, initial material moisture content, and temperature on the dryer capacity and dryer efficiency.

The procedure has been suggested as a means for rating the performance of grain dryers, but it may be applied with equal success to other dryer types or drying methods.

The question who will conduct the experimental and simulated dryer tests on new dryers as well as the existing models on the market is still open. It seems that the establishment of an agreement for a standard procedure of experimental tests is a task for drying

working groups in particular countries or possibly, in the next stage, for a body such as the Drying Working Party of EFCE or the International Energy Agency. Dryer manufacturers and users must also be involved. The simulation part of the rating scheme can be perhaps best obtained with the aid of universities and research centers involved in drying processes.

46.3 HEAT RECOVERY METHODS

Data presented in Table 46.6 show that heat recovery is an important factor for increasing energy efficiency of the drying process. Heat recovery methods commonly applied in the industry are discussed below.

46.3.1 HEAT RECOVERY FROM THE DRYER OUTLET

Traditional heat recovery included the recovery of heat from a dryer outlet, from the hot dried product and/or the outlet gas. Typical devices for heat recovery are presented in Figure 46.5. Conventional heat exchangers have the following efficiency indices: thermal wheel, 75 to 90%; plate exchanger, $\approx 70\%$; heat pipe, $\approx 60\%$; scrubber, $\approx 60\%$; and two-section exchanger, $\approx 50\%$.

Heat may also be recovered from a hot product if this does not cause its rewetting. Heat recovery from a hot stream of outlet gases is not difficult from the technical point of view, provided the outlet gases do not contain dust or condensable volatiles. Such conditions are encountered in drying of timber, paper, and so on, where dust filters and heat exchangers that are easy to disassemble and clean should be used. As an alternative, scrubbers can be used.

46.3.2 HEAT PUMPS

It is worth remembering that if the dryer is working efficiently, the outlet gas should have a temperature close to the wet bulb temperature and also be at high humidity. Therefore, the majority of the gas enthalpy is the latent heat in moisture vapor and heat recovery should, when possible, include moisture condensation from the drying gas. This method is applied in heat pump dehumidifiers.

The principle of the heat pump, which is the same as that involved in the refrigeration operation, has been known for over 100 y. In the last three decades, heat pump applications have been limited only by economics. Among many types of heat pumps (e.g., vapor compression, absorption, ejector, Brayton cycle, and thermoelectric), only the first found wide application because of its high efficiency and relatively simple construction.

A schematic diagram of a heat pump dryer-dehumidifier is shown in Figure 46.6a. The following processes take place in the system: humid gas A leaving the dryer is cooled by an evaporator [5] to a

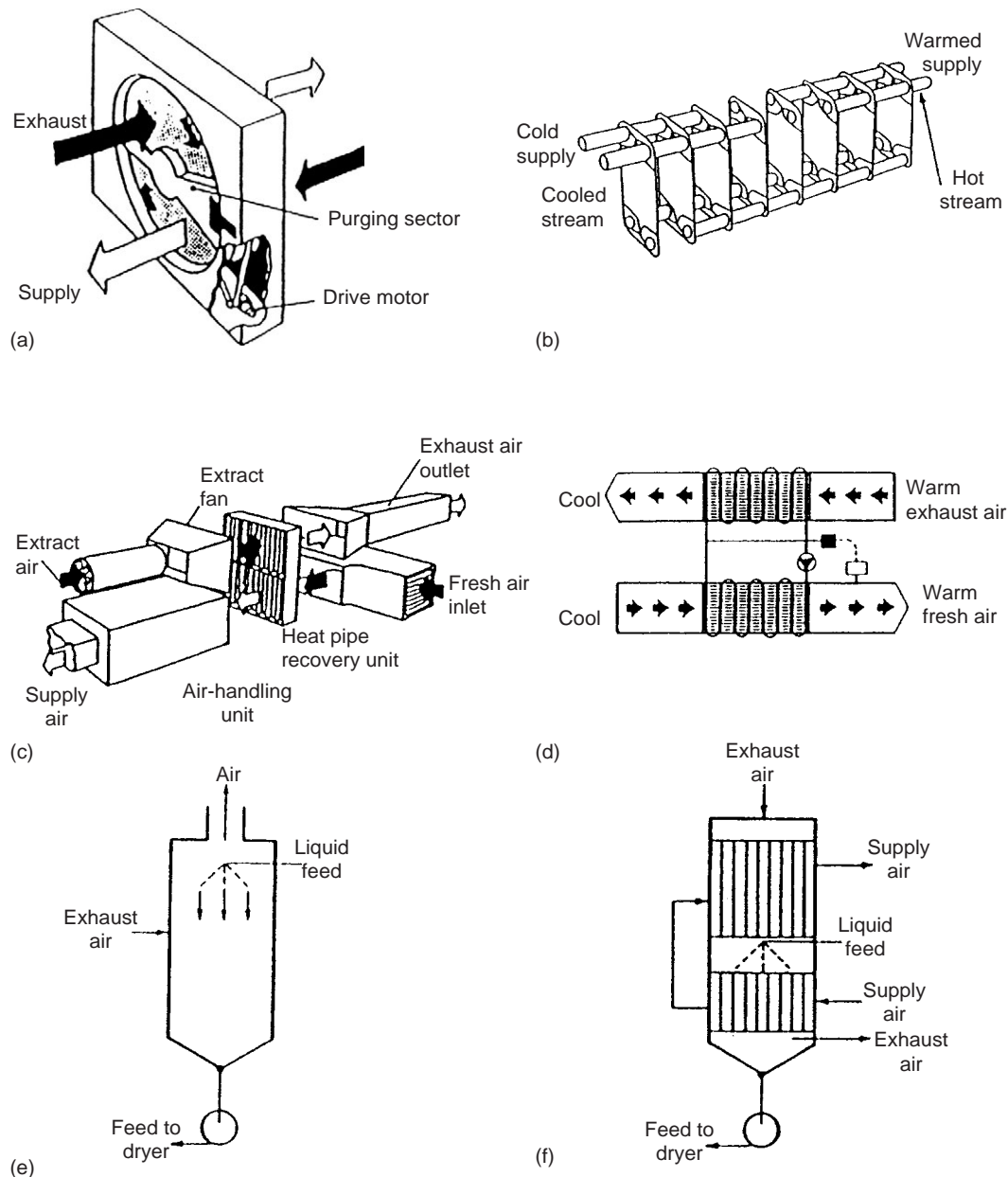


FIGURE 46.5 Heat-recovery systems in dryers: (a) thermal wheel; (b) plate heat exchanger; (c) heat pipe installation; (d) runaround coil; (e) scrubber; and (f) two-section heat exchanger.

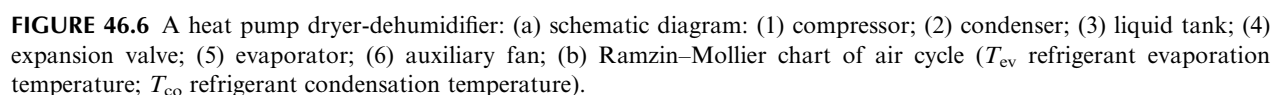
temperature close to T_{ev} below the dew point (B) and (2) consequently, part of the moisture from the air is condensed (F) and heat recovered in this way causes the working fluid (freon) to boil in the evaporator [5]. With the addition of the external work provided by the compressor (1), the freon increases its pressure and temperature; it is then condensed in the condenser (2), thus exchanging heat to the air (C) and raising it to the appropriate temperature (D). The heated and moisture-reduced air (E) can be used in the dryer. The Ramzin–Mollier chart of the air cycle is shown in [Figure 46.6b](#).

The efficiency of a heat pump is measured by the coefficient of performance (COP), and is defined as:

$$\text{COP} = \frac{\text{useful heat output}}{\text{power input}} \quad (46.8)$$

the limiting case being Carnot efficiency, defined as:

$$\text{COP}_{\text{Carnot}} = \frac{T_{\text{condenser}}}{T_{\text{condenser}} - T_{\text{evaporator}}} \quad (46.9)$$



For drying, a better efficiency parameter is the specific moisture evaporation rate (SMER), defined as

The theoretical maximum SMER for conventional thermal drying is 1.55 kg/kWh (i.e., the latent heat of water evaporation at 100°C). A typical value of SMER achieved by a heat pump is 3 kg/kWh, which compares very favorably with conventional convective drying, for which values ranging from 0.5 to 1 kg/kWh are the norm.

cations, e.g., in drying of timber and ceramic products that are easily damaged if subjected to high drying rates and for which the more gentle regime of heat pump drying has economic advantages over and above those associated with energy.

The scope of this chapter does not allow the authors to present a complete discussion on the optimization of heat pumps in dryers. However, the most important conclusions from [33] and [34] are given below:

1. Compressors of heat pump dryers should be able to operate continuously for several days
2. Heat capacity and working temperature of the heat pump should be designed to suit the drying process
3. The heat pump should not necessarily be the only heating device in the dryer and should be treated as a moisture-extracting device
4. For optimum performance, the heat pump should operate at the same heat load (i.e., with the same amount of moisture condensed and at a constant temperature), and the compensation for any changes in dryer operating conditions should be by other methods
5. The temperature difference between an evaporator and condenser for a single-stage vapor-compression heat pump should not exceed 30 to 40°C. This gives practical COP in the range of 3 to 4.

The combination of the heat pump system with other heat sources improves the economic case. Figure 46.7 shows a coupled heat pump–gas engine used for heating of a malt house and for drying of malt [38]. The application of this system reduced energy consumption by 40%, which resulted in £121,600 savings per year [39]. However, a number of problems were encountered with the gas engine drives, e.g., compressor seal damage due to vibration, as well as maintenance costs and corrosion of the heat exchanger. Consequently, no further systems have been installed [32].

Electrically driven heat pumps are potentially very attractive as most of the main components were directly transferred from refrigeration technology. This, however, limited the early applications to the temperature range 60 to 65°C. By choosing working fluids other than R-12 and R-22 it was possible to extend the range of heat pumps to temperatures up to 120°C, thereby reducing energy consumption and allowing the use of this drying method for a wider range of materials. Tests on batch timber drying comparing modern conventional kilns and a heat pump system using R-114 showed that, with the high-temperature heat pump, energy costs savings of 40% for drying of hardwoods and 65% for softwoods could be achieved with U.K. energy price ratios [32].

The wider application of the vapor-compression heat pump in the future may be jeopardized by concern about the ozone-depleting potential of the working fluids (freons) that were developed for high-temperature heat pumps (e.g., R-114) [40] (Figure 46.8).

One of the working fluids that is extensively researched is carbon dioxide used in transcritical conditions [35–37]. Technically advanced heat pump was used for drying of several materials giving promising results. Application of this working fluid is limited to 110°C by high pressure at elevated temperatures that in turn leads to special construction of compressors and heat exchangers.

Extension of heat pump technology to large-scale continuous operations and temperatures above 120°C points to water vapor as a working fluid, provided steam compression can be done efficiently. In such a system, water evaporated from material being dried is compressed and used for heating incoming material. The principle of operation is similar to the well-known mechanical vapor recompression (MVR) systems. Efficiency of the process will decrease if noncondensable gases such as air are present with the system.

46.3.3 MECHANICAL VAPOR RECOMPRESSION

MVR technology can be used for concentrating dilute solutions prior to drying. This technique is characterized by high efficiency and is used as a mechanical evaporator in many branches of the industry, e.g., the Carver–Greenfield process for wastewater utilization [41].

Figure 46.9 shows an installation in which wastewater is concentrated from 2 to 50% of solid content. By using MVR, the process has a high efficiency, giving an energy usage of about 134 kJ/kg water, which is 20 times better than the typical adiabatic spray drying.

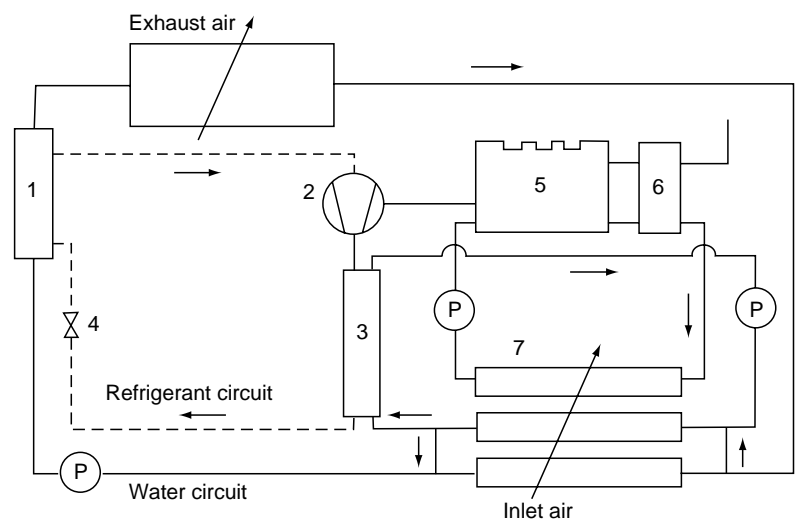


FIGURE 46.7 Coupled heat pump–gas engine: (1) evaporator; (2) compressor; (3) condenser; (4) expansion valve; (5) gas engine; (6) engine exhaust gas heat exchanger; and (7) radiator and heat exchanger (P), pump.

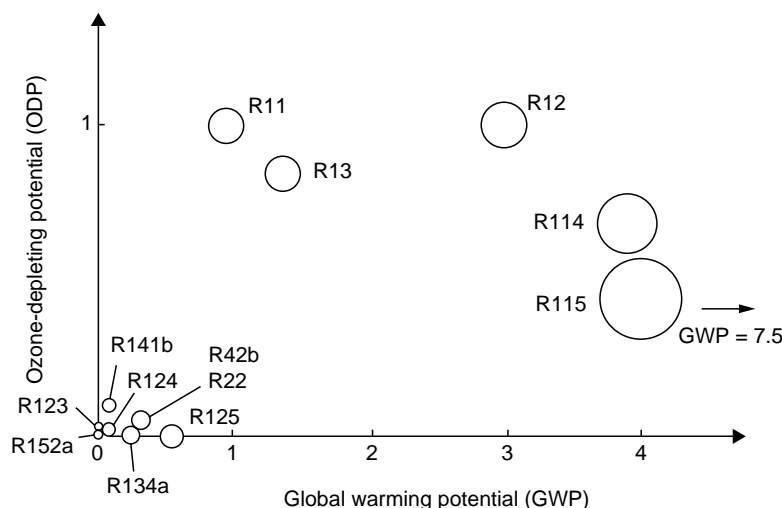


FIGURE 46.8 Ozone-depleting potential and global warming potential of chlorofluorocarbons (CFC) and hydrochlorofluorocarbons (HCFC) relative to R11 = 1.

46.4 NONCONVENTIONAL ENERGY SOURCES IN THE DRYING PROCESS

Alternative energy sources for drying include waste heat from other processes carried out in the same plant as the drying. For efficiency, these should be continuously available and have a sufficiently high temperature to avoid the need to use some form of heat accumulator and heat pump. Otherwise, what

was apparently a “free” heat source becomes uneconomic. A typical successful example is the drying of bricks prior to firing, in which the flue gases from the kiln are utilized.

46.4.1 SOLAR ENERGY DRYING

Solar energy can be treated as a free energy, but its application for drying is limited geographically and

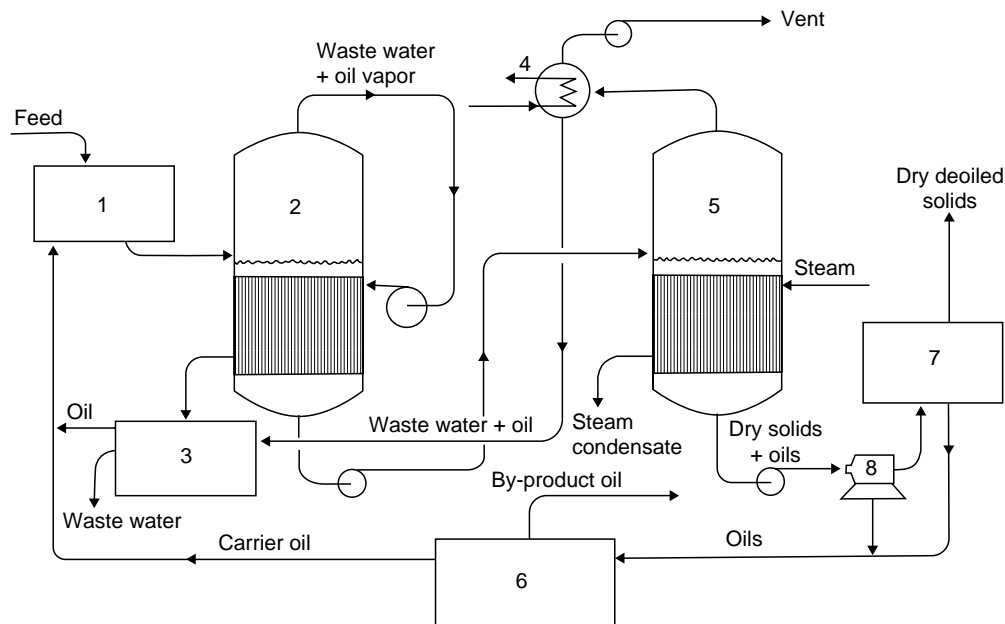


FIGURE 46.9 Mechanical vapor recompression +1 (MVR + 1) Carver–Greenfield process configuration: (1) fluidizing tank; (2) MVR stage; (3) separator; (4) vacuum system; (5) final evaporator stage; (6) oil separation system; (7) deoiler; and (8) centrifuge.

in general to agricultural products, e.g., ground nuts, coffee beans, cocoa, maize, cassava, grapes, rice, barley, wheat, salt, and pepper. The cost of solar drying varies from 0.1% of product value for coffee beans up to 15% for cassava [42]. In a temperate climate, solar energy of countries is limited to the drying of hay and straw. Various solar dryer system configurations were studied by Reddy [43]; further information on their application in agriculture (e.g., grain drying, design of forage dryers, fruit dryers, lumber kilns, and free convection dryers) and related topics can be found in Ref. [45].

46.4.2 BIOMASS ENERGY SOURCE

Agricultural crop residues, as by-products in normal crop production systems, are among the cheapest, most readily available energy sources on the farm. These renewable fuels provide an option for meeting some part of the future energy needs. Biomass from agriculture was predicted to make up to 5% of the U.S. energy needs by the year 2000 [44]. Direct combustion of crop residues and biomass is perhaps the best way to use its stored energy; this could heat air for grain drying. Burning is presently the most widely used technique for converting biomass materials to energy. According to an analysis [46], the residues recovered from one hectare of corn should provide enough heat energy to dry corn from 10 ha or more.

46.4.3 INDUCTION DRYING

The dryer shown in Figure 46.10 consists of a slowly rotating inclined drum, the shell of which is heated by large electric currents induced in it by induction. The drum, incorporating a rotating transformer, revolves around a fixed copper tube carrying a very large current at low voltage; this large current is transferred to the stainless steel drum, thus heating the shell wall and the internal lighting flights.

Because the heating is very efficient and the heat is produced directly in contact with the product, substantial savings can be made in some applications. For example, in the largest copper powder factory in Europe, the substitution of the existing infrared dryers by ROTEK gave savings amounting to £38,000/y and a payback time of 3 y [47].

46.4.4 PULSE COMBUSTION DRYING

Pulse combustion drying combines high economic of direct fire heating system with high intensity of drying of disperse material. Typical power consumption is 3,000 to 3,500 kJ/kg H₂O evaporated (4,500 to –11,500 for spray dryer and 3,200 to 6,500 for drum dryer) [48,49]. The main advantages of pulse combustion drying are:

- Improved heat transfer, high difference between the material and flue gas temperature and high drying rate

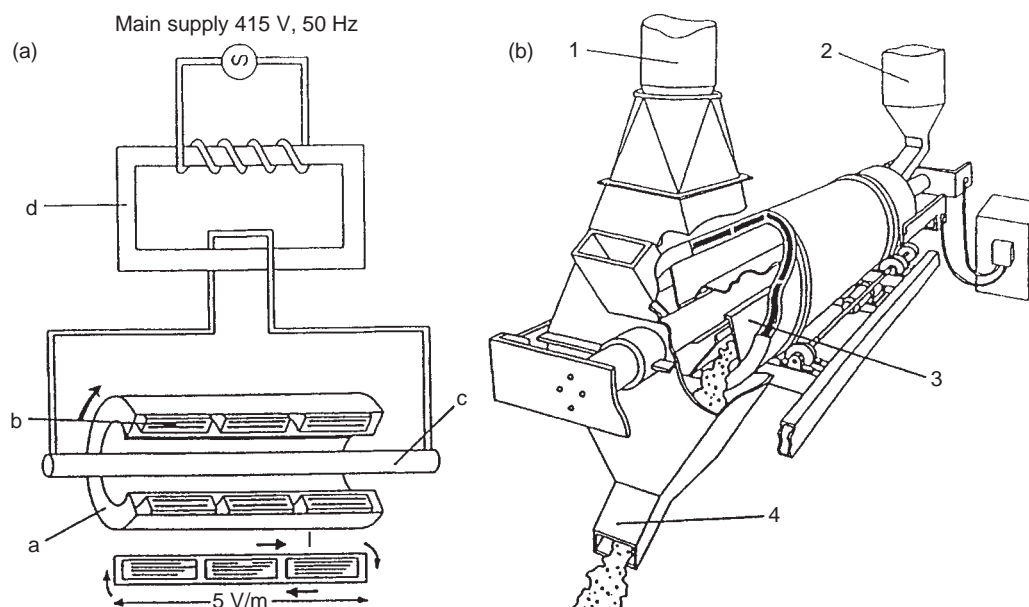


FIGURE 46.10 The ROTEK dryer: principle of induction heating: (a) rotating drum at 15 to 20 rpm; (b) single-turn tertiary load 8000 A at 5 V/m; (c) secondary bus bar; and (d) silicon iron transformer core; (b) cross section of ROTEK dryer: (1) air extraction duct; (2) feed hopper; (3) flights of fins; and (4) delivery chute.

- Efficient combustion with low emission of toxic substances and low amount of air discharged to atmosphere
- Wide variety of feedstock handled including sticky materials and heat sensitive products
- Better atomization and thus no need for atomizer or HP nozzle
- Saving on auxiliary equipment as combustor delivers energy to run the dryer and displaces fan—typically requires less electrical energy.

The main disadvantage is noise produced by the pulse combustor and difficult control of the combustor.

46.5 SOME NEW DRYING TECHNIQUES

Special technological methods for the reduction of energy use during drying include, among others, the application of sorbents that interact with the material being dried [50,51], where the energy-consuming process of moisture evaporation is replaced by the mass transfer process.

46.5.1 MULTISTAGE DRYING

The thermal efficiency of any dryer depends on drying temperatures; therefore, drying of temperature-sensitive materials such as organic or biomaterials is usually a more energy-consuming process. After the energy crisis, manufacturers of drying equipment modified designs to improve overall utilization of energy. Designers considered that constant-rate drying that is governed by moisture diffusion in air, could be carried out in a separate unit with high air temperature. The falling rate drying period that is controlled by internal moisture diffusion, is a much slower process and thus moisture requires time to diffuse out of the material. This second phase of the process can therefore take place in a separate section with less demanding conditions. This led to the concept of the two-stage dryers.

An example is a skimmed drying system that consists of a spray and vibrofluidized bed dryer. This lowered the specific energy consumption to 4300 kJ/kg from 5550 kJ/kg water evaporated for the spray dryer alone [52]. The next substantial advance in spray dryer design was the compact spray dryer. In this system, a stationary fluid bed is positioned on the lower part of the spray chamber, forming the bottom and discharge part of the dryer. Combining two units into one dryer and with a proper utilization of the drying air gives a much improved economy (3620 kJ/kg) over the conventional two-stage dryer. This design was further developed by applying

different air temperatures to the spray nozzle zone, main chamber, and fluid bed part. For the same material, it was possible to decrease energy consumption to 3550 kJ/kg water evaporated.

46.5.2 SUPERHEATED-STEAM DRYING

As described above, under some circumstances it is possible to use superheated vapor instead of traditional dry air for drying. This would be supplied to the dryer at temperatures well above 100°C. By using superheated steam, higher heat transfer rates can be achieved. Clearly, however, there is no effect on the other limitation of the drying rate, the moisture diffusion within the material.

It has been shown that, above a certain temperature, i.e., inversion temperature, the evaporation rate in a superheated-steam dryer exceeds that in an adiabatic air dryer [53,54]. The theory indicates that the inversion temperature can vary from 160 to 300°C.

An economic analysis of a superheated-steam drying of aluminum oxide from 40 to 2% moisture content and an output of 2000 kg/h showed that the optimum inlet vapor temperature into such a dryer is 374°C [53]. The dryer cost was lower than that of a comparable air dryer and the operating costs were lower by about 15%. [Figure 46.11](#) shows a schematic diagram of a superheated drying system in a sugar factory [55].

In the literature, a number of advantages are claimed for the method, which include: (1) elimination of product oxidation; (2) possibility of almost complete heat recovery from the outlet vapor by its recompression; (3) smaller dryer size; (4) reduced fire risk; and (5) improved process control. There are, however, some disadvantages; e.g., all materials cannot withstand the elevated temperatures and in some cases low moisture contents cannot be attained. The product feed and discharge ports are more difficult to build because of the need to contain the steam and prevent the ingress of air.

In general, it is considered that the advantages of superheated-steam drying are such that over the next decade the method will be widely adopted [54].

46.5.3 RADIOFREQUENCY AND MICROWAVE

RF and microwave techniques offer the unique feature of generating heat in the water held in a wet substrate. Because of this and the fact that no part of the product is subjected to temperatures in excess of the wet bulb temperature, efficiencies can be very high. However, because of the relatively high cost of the equipment, it is not practical to consider these techniques for large throughput materials. It is more usual to associate them with high value, modest throughput items, e.g.,

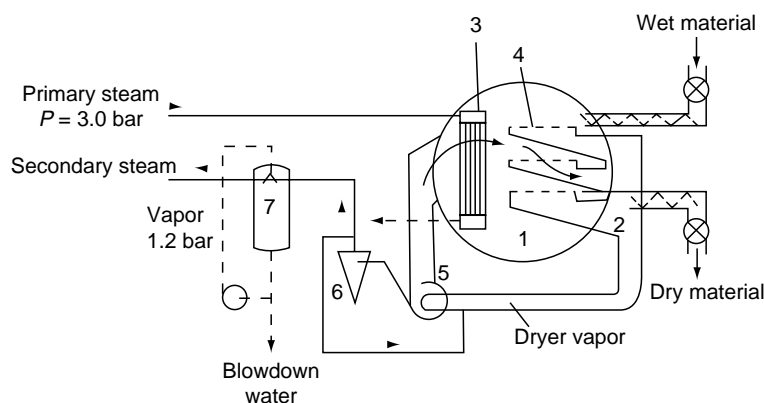


FIGURE 46.11 Schematic diagram of superheated-steam dryer: (1) drying chamber; (2) vapor outlet; (3) heat exchanger; (4) 3 screen levels; (5) circulating fan; (6) centrifuge separator; and (7) scrubber.

wool textiles and certain food products, e.g., biscuits and breakfast cereals. The economic case is often strengthened by the fact that very uniform moisture distributions are obtained within a product, particularly important in paper making [56].

A more recent advance has been the introduction of the combined RF convective dryer in which RF heats the volume of the product and, by reducing the viscosity and surface tension of the water, causes liquid phase flow to the surface, at which greater part of the required energy is provided by the convective, normally gas-fired heat. These equipments are used in a range of industrial applications on particulates, three-dimensional solids, and works for which pure RF would not be cost effective. The overall energy efficiency is very high [57].

view of both energy consumption and product quality [58].

A microcomputer control makes it possible to adapt drying conditions to the real material parameters and usually leads to a decrease in drying time, which prevents product overdrying and limits energy usage. Such control of drying requires a knowledge of the process itself, the measurement parameters in real time, as well as programming skills. The strategy presented in Figure 46.12 is used to develop a microcomputer control of the process.

The most successful applications of computer-aided control are based on the experience and knowledge of people operating a given dryer type since the dryer start-up, the constant- and falling drying-rate periods, as well as completion of the drying process

46.6 PROCESS CONTROL

In practical applications, from the point of view of both the product quality and energy savings, control of the drying process is of significant importance. For optimization of heat balance, it follows that the drying should be finished at the maximum material moisture content that is acceptable.

The quality control of the products is a significant element in energy savings. For example, if the required parameters are not achieved, the products may require additional treatment at extra cost, which may cause a reduction in the throughput of the dryer or involve a further process. Moreover, the risk of environmental pollution and explosion hazards increases when drying powders to moisture contents below the optimum condition. The energy consumption may also be reduced considerably by the use of automatic control of the dryers. These ensure optimum operating conditions from the point of

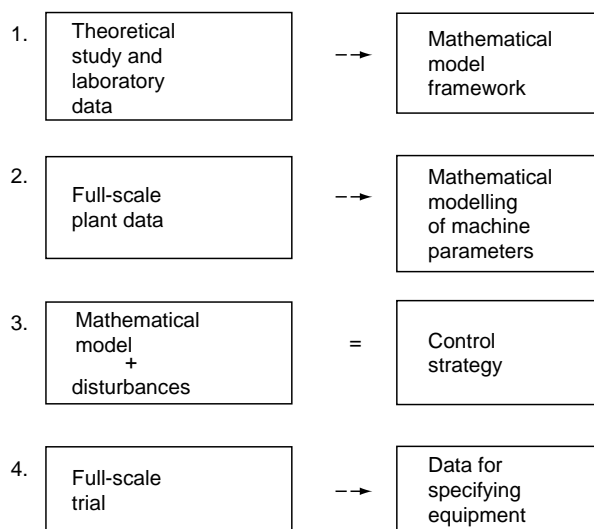


FIGURE 46.12 Strategy for finding a suitable dryer control.

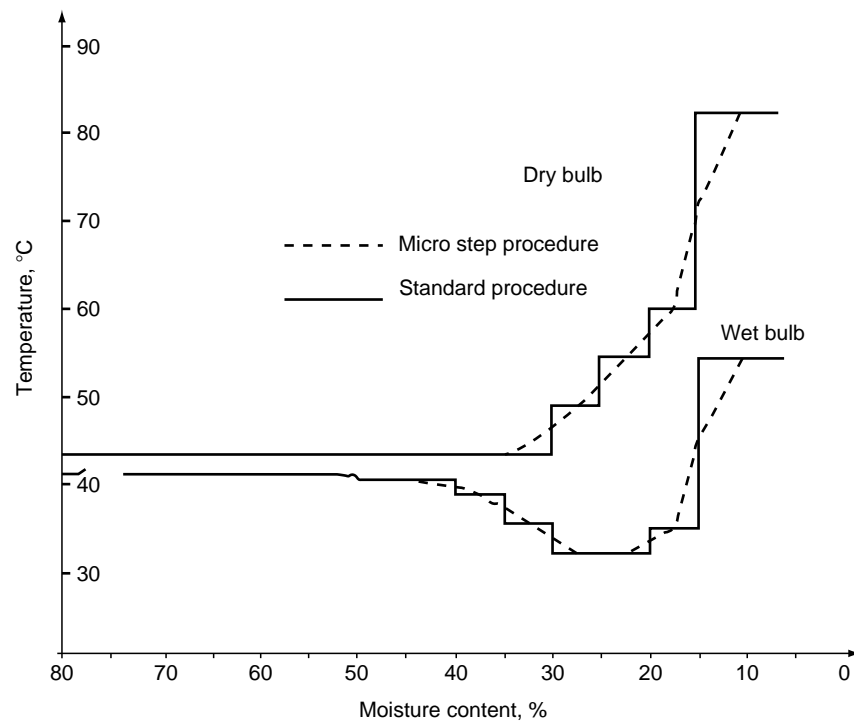


FIGURE 46.13 Examples of standard vs. ministep drying schedule drying.

require different procedures. For instance, in timber drying, which lasts for several days, the optimal drying conditions on particular stages were determined [59] (see solid line in Figure 46.13) [60]. In the early stages, timber requires drying with warm, humid air, while in the final stage, hot and dry air could be used. The goal of computer-aided control was to pass smoothly from one range of conditions to another (see dotted line in Figure 46.13).

An algorithm of a computer program controlling the process start-up, both drying periods, and process completion was presented by Little and Toennisson [60]. The application of such a controller made it possible to reduce the time of drying from 74 to 7% moisture content from 43 to 34 days (Figure 46.14) [60]. The average estimated energy savings were 6.5%. Even greater energy savings can be obtained, as reported by Richard et al. [61]. In the case of a rotary talc dryer, the application of a graduate control loop involved energy savings of about 20% as well as a 40% increase of the dryer throughput.

In the case of continuous dryers, the monitoring and control of the process is usually limited to maintain the constant moisture content of the final product under variable conditions, e.g., with changing material inlet moisture content and drying gas humidity.

The major problem faced by the designer of the control system is the slow response of the equipment

that measures moisture content of a solid when compared with the process time. This makes direct control of the process nearly impossible and, therefore, indirect measurements are made, e.g., by measuring the dielectric constant or electrical resistance of material being dried. In addition, there is an approach based on the knowledge of interrelations of the final moisture content and temperature of the product and the condition of the outlet gas. These parameters are more easily measured in real time and can, therefore, be used in control systems.

The most common and simple applied control system (Figure 46.15) varies with the amount of heat supplied to the dryer in such a way as to maintain constant outlet gas temperature [62]. If, however, an initial moisture content of the material varies, this technique does not give a constant final moisture content. In the modification of control systems, the experimentally tested dependence of the outlet gas temperature T_o on the inlet temperature T_i is taken into account. For a given material at constant final moisture content, a linear relation $T_o = kT_i + b$ occurs for higher temperatures (Figure 46.16) [62]. The value of k depends on material properties: a higher k corresponds to a material that is more difficult to dry. The linear relation makes it possible to introduce a simple automatic system without the use of a microcomputer. Figure 46.17 shows

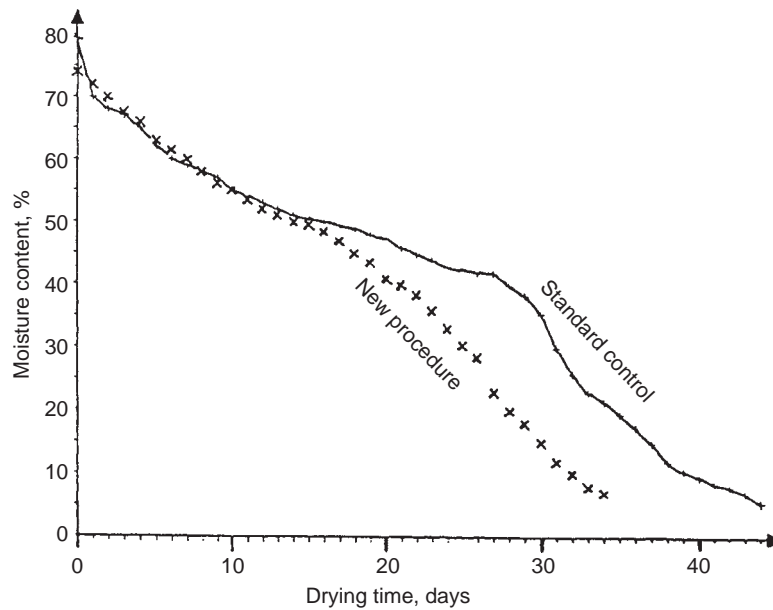


FIGURE 46.14 Comparison of drying process for standard control and new microcomputer algorithm.

a dryer with modified automatic control system [62] in which, unlike the previous control system, proportioning and time delay elements are also used.

Similar considerations can be made for steam-heated batch and continuous drum dryers. The automatic system presented in [62] also takes into account

and compensates for any fluctuations of heating steam pressure.

Despite the fact that, for a given material and dryer layout, the automatic system settings may be constant, it is desirable to install the microcomputer monitoring system. This may lead to increased knowledge of the

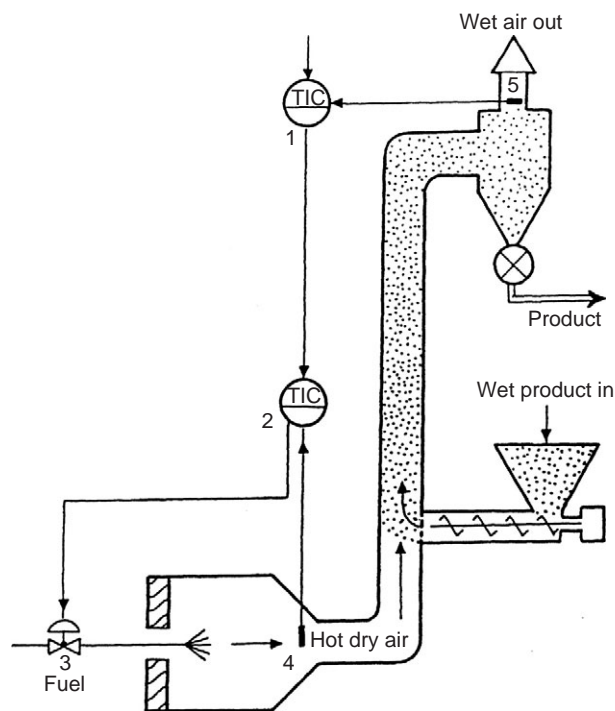


FIGURE 46.15 Flash dryer temperature control: (1) outlet temperature controller; (2) inlet temperature controller; (3) fuel valve; and (4, 5) temperature sensor.

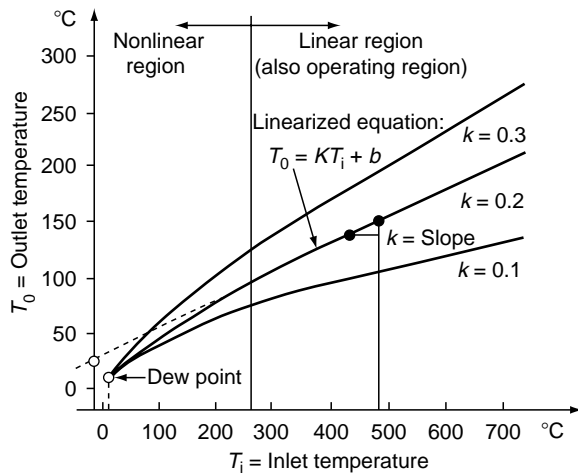


FIGURE 46.16 Temperature characteristic of a flash dryer.

process, instantaneous calculation of its efficiency, and, subsequently, the ability to relate the process parameters to product quality.

The recent availability of low-cost, powerful PCs has made it possible to introduce learning of drying processes based on fuzzy logic theory and

neural network concepts [63]. The main control objective in drying is to minimize fuel consumption and to keep the moisture content of dried product less than or equal to a certain value, despite the external perturbations acting on the process. The computer program implemented in the controller uses a reinforcement scheme to update its action probabilities according to the response of the environment, in this case drying, and thus improves its performance with time. This control approach does not need a development of a model for the process to be controlled. Simulation results demonstrate the feasibility of using such a control for a phosphate rotary dryer [64].

46.7 ENVIRONMENTAL PROBLEMS

In general, the recovery of heat and the elimination of other waste should be considered in terms of both cost and ecology. Drying may have several impacts on the environment. First, as a significant user of fuels, it contributes to atmospheric pollution and global warming. However, because of a wide variety of energy sources, dryer constructions, process parameters,

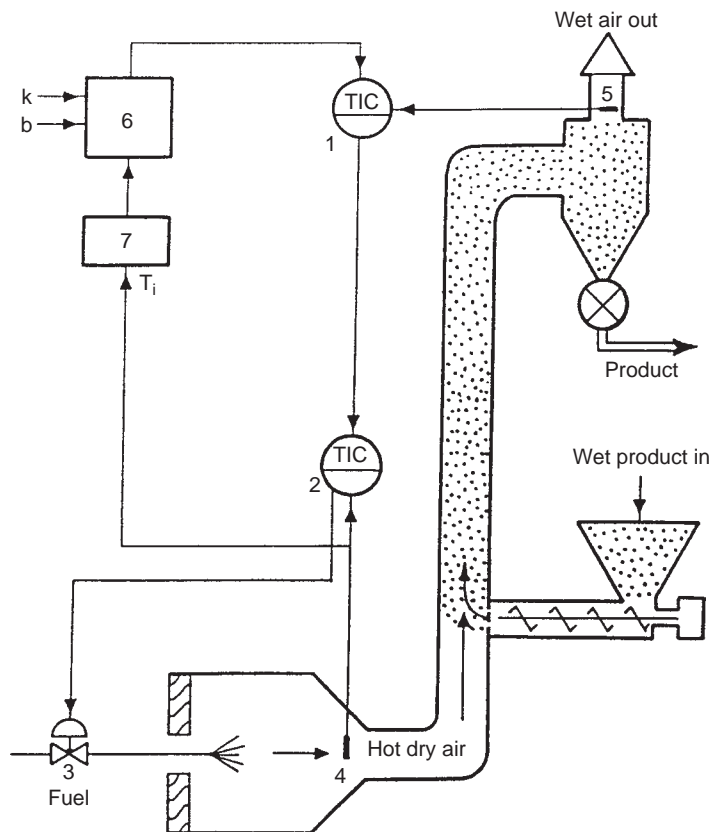


FIGURE 46.17 New compensated dryer control system: (1) outlet temperature controller; (2) inlet temperature controller; (3) fuel valve; (4, 5) temperature sensor; (6) linear compensation module; and (7) time lag module.

and product properties, it is not possible to offer a general solution for the ecological problems that arise. The dryer should not necessarily be regarded in isolation but considered as a part of a larger system. For example, although electric power may be rightly considered as a clean and environmentally acceptable source of energy at point of use, its production may have implications in terms of pollution. The corollary is that low-efficiency, fuel-fired systems can contribute a greater burden in terms of atmospheric pollution. In any case, the whole of the energy input needs to be considered since most dryers have some electrical load, albeit mainly motive power.

The methods of energy supply to the drying installation include combustion of fossil fuel, with the heat from the products of combustion being directly transferred to the product or indirectly with a heat exchanger. Intermediate heat transfer fluids such as steam or hot water are sometimes used either from a heat exchanger or the dryer air from a central plant. In very large-scale operations such as paper making, the plant may have its own power station that allows both heat and power to be used, resulting in very high efficiencies.

The second effect of drying on the environment arises from waste streams other than products of combustion and heat, i.e., dust and noncondensable vapors.

According to some authors [65], in certain countries the increased use of nuclear energy and nontraditional renewable sources in place of oil and gas can contribute to a reduced energy cost for drying.

Having in mind the variety of energy supply methods, dryer constructions, operating parameters, and properties of materials it is not possible to propose a universal solution from the point of view of ecology. The problems should be, therefore, solved individually. There are two approaches [66]:

1. The gathering and possible treatment of waste as it is currently produced. This, however, does not reduce the quantity to be dealt with unless it can be put back into the process.
2. A second, more radical approach is the development of low by-product, energy-saving technologies by radically changing techniques (e.g., drying). The purpose of this is to reduce the quantity of waste produced or to reuse it as a product in its own right.

The efficiency of drying installations based on direct firing depends on minimizing the solid particle emission with combustion products. For example, the replacement of chamber furnaces by cyclone furnaces leads to a reduction in fuel consumption, improves

reliability of the equipment, and provides a significant decrease in the quantity of noxious gases emitted to the atmosphere. Automatic combustion control in burners used in dryers can result in energy savings as well as a decrease in pollution of both the product and environment. The exhaust air from textile dryers with temperatures above 100°C often contains dust, fibers, toluene, formaldehyde, amines, and other pollutants. In order not to pollute the heat transfer area, these gases are often dumped when they are judged by the operator to be close to saturation, with the setting of bypass and exhaust dumpers often being rather arbitrary. These gases can be used in heating up of steam boilers instead of directing them to heat exchangers [67].

In addition to certain increase in energy cost, the efficiency of drying needs to be considered in the context of global warming and environmental protection. Energy consumption, whether by the burning of fossil fuels in power stations or in the dryer itself, leads to CO₂ emission. These indirect atmospheric effects of drying systems will have to be reduced by more efficient drying processes. At the same time, environmental considerations require that the emission of solvents be minimized and, as far as is practical, eliminated.

46.8 EXAMPLES OF APPLICATIONS OF SOME ENERGY-SAVING METHODS

Several typical examples of possible energy savings in drying processes applied in various industries are given below. These examples and [Figure 46.18](#) through [Figure 46.23](#) are presented with permission of the Energy Efficiency Office of the U.K. Department of Energy.

46.8.1 APPLICABILITY OF HEAT EXCHANGERS

[Figure 46.18](#) presents a schematic diagram of a heat-recovery system in the drying section of a paper machine [68]. Air, which previously was dumped to the atmosphere, now passes through a highly efficient compact heat exchanger in which it heats up inlet air before it enters the gas-heated air heater. The energy saving in one paper mill was shown to amount to 25,000 GJ (i.e., about £66,000).

46.8.2 PROPER INSULATION

Another very simple and inexpensive energy-saving solution used in a paper machine was the fitting of thermal insulation to the drying cylinder end caps ([Figure 46.19](#)) [69]. Measurements revealed that after insulation energy utilization decreased between 1.7 and 3%. This very

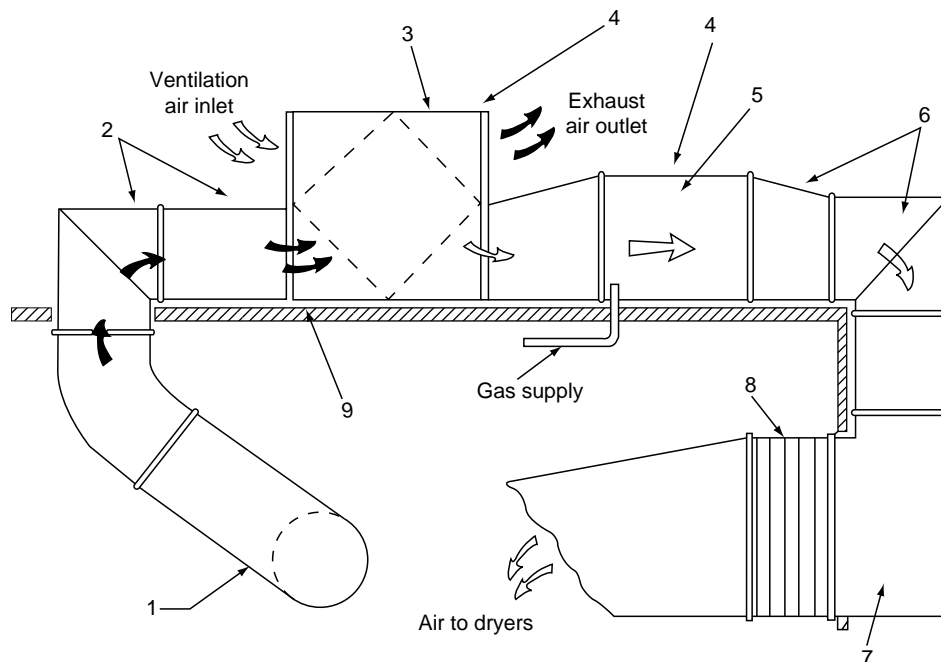


FIGURE 46.18 Installation of coupled heat recovery and direct firing: (1) exhaust duct; (2) new ducting; (3) air/air heat exchanger; (4) weather and thermal insulation; (5) direct gas-firing unit; (6, 7) new ducting; (8) steam coil unit; and (9) roof structure. (Courtesy of Energy Efficiency Office of the U.K. Department of Energy.)

simple improvement gave savings of about 1325 GJ/y (i.e., about £3500 for an investment of £4700).

46.8.3 APPLICATION OF A RUNAROUND COIL HEAT EXCHANGER

As mentioned above, heat recovery from outlet vapors is a well-documented method for energy saving. In Figure 46.20, the application of two run-around coils in a spray dryer for dyes is shown schematically [70]. An important feature of this heat exchanger is the use of glass tubes from which it is easy to remove water-soluble dyes. The use of this

system leads to a reduction in heat consumption by about 16% and annual savings of 4,500 GJ (i.e., £12,000).

46.8.4 APPLICATION OF GLASS-TUBE HEAT EXCHANGERS

The advantages of glass tubes used in a heat exchanger include corrosion resistance, easy cleaning, easy visual inspection, and good resistance to thermal shock. The heat-recovery systems containing glass tubes (Figure 46.21) were used in a 3-m diameter spray dryer used for drying chemicals [71].

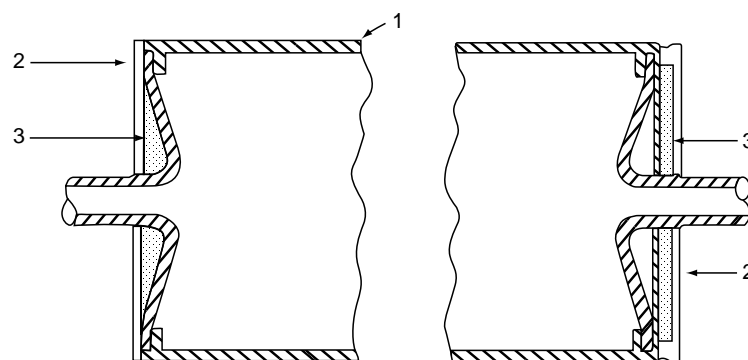


FIGURE 46.19 Drying cylinder insulation: (1) drying cylinder; (2) aluminum cover; and (3) thermal insulation material. (Courtesy of Energy Efficiency Office of the U.K. Department of Energy.)

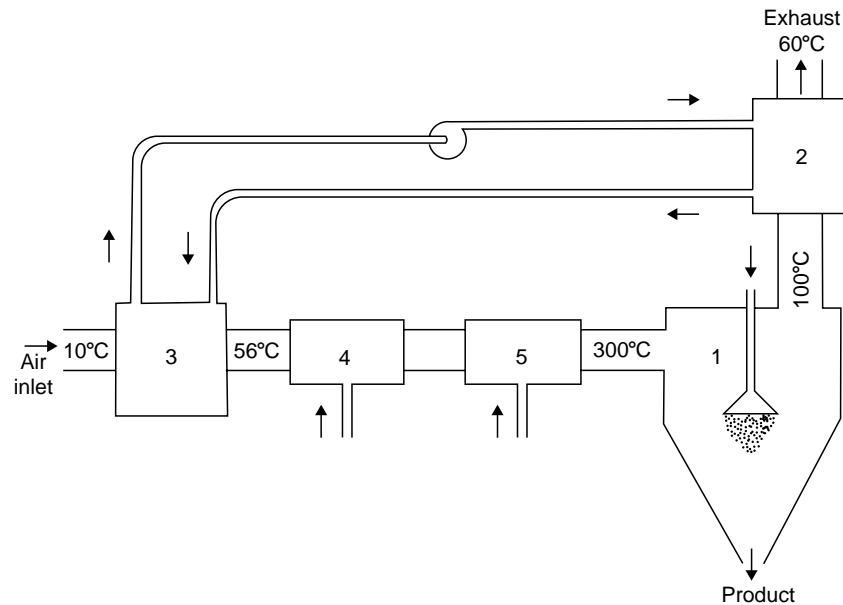


FIGURE 46.20 Schematic of two runaround coils in a spray dryer for dyes: (1) spray dryer; (2) exhaust heat exchanger; (3) inlet heat exchanger; (4) steam heater; and (5) gas burner. (Courtesy of Energy Efficiency Office of the U. K. Department of Energy.)

46.8.5 DRYING OF LARGE OBJECTS AND STRUCTURES

An unusual example of a dryer is the one used for drying of tundishes that are used for continuous casting of ingots [72]. In this particular plant, 14 tundishes, 6.4 m long, 1.3 m wide, and 0.75 m high are coated and then dried thoroughly. The solution shown in Figure 46.22 consists of setting gas burners in the plates covering the tundishes during drying.

Following an improvement in the construction of burners and the addition of a PLC control, the energy

consumption was reduced by 97% and the average life of tundish coating was increased. Figure 46.23 shows the energy consumption for tundish drying (in GJ/week) a year before modernization and a year after. The capital expenditure was £33,000 and annual savings was £65,000.

46.9 CONCLUSIONS

At present, the users of drying equipment see their profit in high-quality products, while excessive energy consumption is a problem of minor significance.

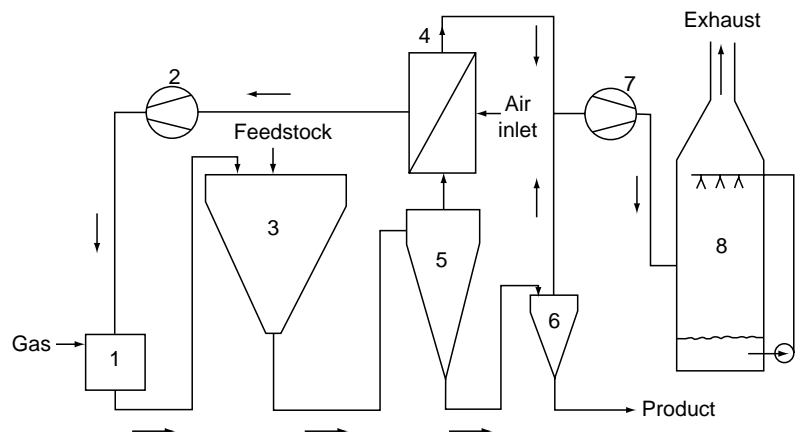


FIGURE 46.21 Schematic diagram of plant layout including heat exchanger: (1) air heater; (2) auxiliary fan; (3) spray dryer; (4) glass-tube heat exchanger; (5) main cyclone; (6) auxiliary cyclone; (7) main fan; and (8) scrubber. (Courtesy of Energy Efficiency Office of the U.K. Department of Energy.)

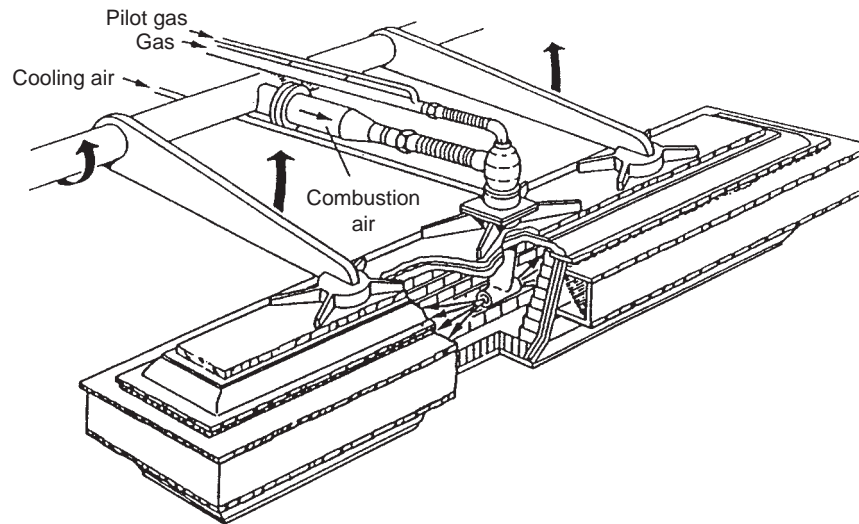


FIGURE 46.22 Tundish dryer in the foundry. (Courtesy of Energy Efficiency Office of the U.K. Department of Energy.)

Thus, energy-saving techniques should be recommended not only for economic reasons, but also to prevent energy spilling to the environment. The authors realize that in 1994 energy is not as valuable as it was during the oil crisis of the 1980s. However, the sole saving of money is not the most important objective. Even more crucial is the recognition of the effect of excessive energy consumption on the natural environment, emission of the greenhouse effect gases, and, as a consequence, global warming.

Recommendations in this area are:

- Analyze the existing dryers more thoroughly both from the point of view of the dryer itself and from the point of view of material characteristics
- Use advanced models to improve dryer design
- Take advantage of the users' experience and knowledge
- Respect energy as a valuable source
- Treat exhaust heat as a factor polluting the environment and consider energy problems jointly with ecology.

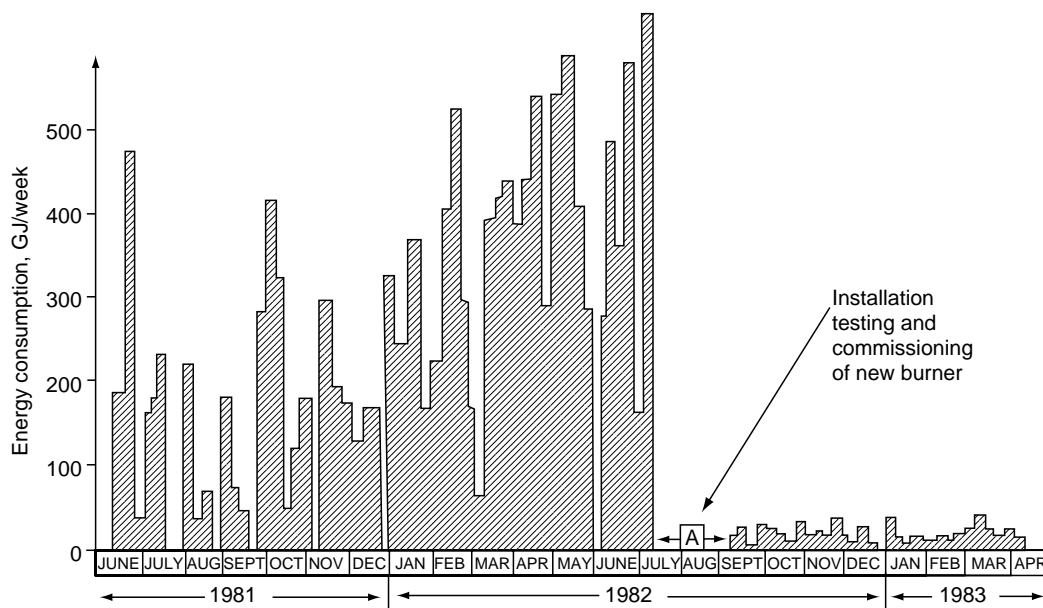


FIGURE 46.23 Energy consumption in a foundry drying system before and after installation of new burner. (Courtesy of Energy Efficiency Office of the U.K. Department of Energy.)

SYMBOLS

E	energy, J
t	time, s
T	temperature, K (°C)
T_2	outlet air temperature, °C
X	material moisture content (db), kg/kg
X'	material moisture content (wb), kg/kg
y	air moisture content (db), kg/kg
$\bar{\epsilon}$	cumulative efficiency
$\bar{\epsilon}$	instantaneous efficiency
η	energy efficiency (lumped parameter)
φ	relative gas humidity, %
wb	wet basis
db	dry bases

SUBSCRIPTS

Co	condensation
D	drying
e	evaporation
E	energy
t	total
T	temperature
m	material
max	maximum
wb	wet bulb
0	ambient
1	inlet
2	outlet

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47 Heat Pump Drying Systems

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47.1 INTRODUCTION

Heat pump dryers have been known to be energy efficient when used in conjunction with drying operations. The principal advantages of heat pump dryers emerge from the ability of the heat pumps to recover energy from the exhaust gas as well as their ability to control the drying gas temperature and humidity. Many researchers have demonstrated the importance of producing a range of precise drying conditions to dry a wide range of products and improve their quality. At the same time, MacArthur (1984) has mentioned the need to optimize component and system design to increase energy efficiency in heat pump systems.

Any dryer that uses convection as the primary mode of heat input to the dryer (with or without supplementary heat input by other modes of heat transfer) can be fitted with a suitably designed heat pump (HP). Although batch shelf, tray dryers, or kilns (for wood) are the most commonly reported dryers used in conjunction with heat pumps, other types may also be used, e.g., fluid beds (Alves-Filho and Strømme, 1996; Strømme and Jonassen, 1996) and rotary dryers. However, dryers that require large amounts of drying air, e.g., flash or spray dryers, are not suited for HP operation. Figure 47.1 displays a generalized classification scheme for heat pump dryers based on the processing mode, number of drying stages, number stages of heat pump, types of

auxiliary heat input, and heat pump dryer operation. Many of these classes of heat pump dryer have been proposed and reported over the last two decades. However, this flowchart includes some classes that are proposed here for the first time.

47.2 THE BASICS OF A HEAT PUMP

The basic components of the heat pump system comprise an expansion valve, two heat exchangers (evaporator and condenser), and a compressor. A schematic diagram depicting the operation of a heat pump dryer is shown in Figure 47.2.

Figure 47.3 and Figure 47.4 show the temperature–entropy and pressure–enthalpy diagrams of the heat pump cycle, respectively. The heat pump cycle works as follows:

- The cooling and dehumidification of the air occurs at the evaporator. The refrigerant, moving from point 1 to 2, absorbs heat from the air and undergoes a two-phase change from vapor–liquid mixture to vapor. Evaporation of the refrigerant is achieved by the gaseous escape of the molecules from the surface of a liquid while maintaining its temperature and pressure.
- The refrigerant vapor enters the suction line of the compressor at point 3. The electrical energy

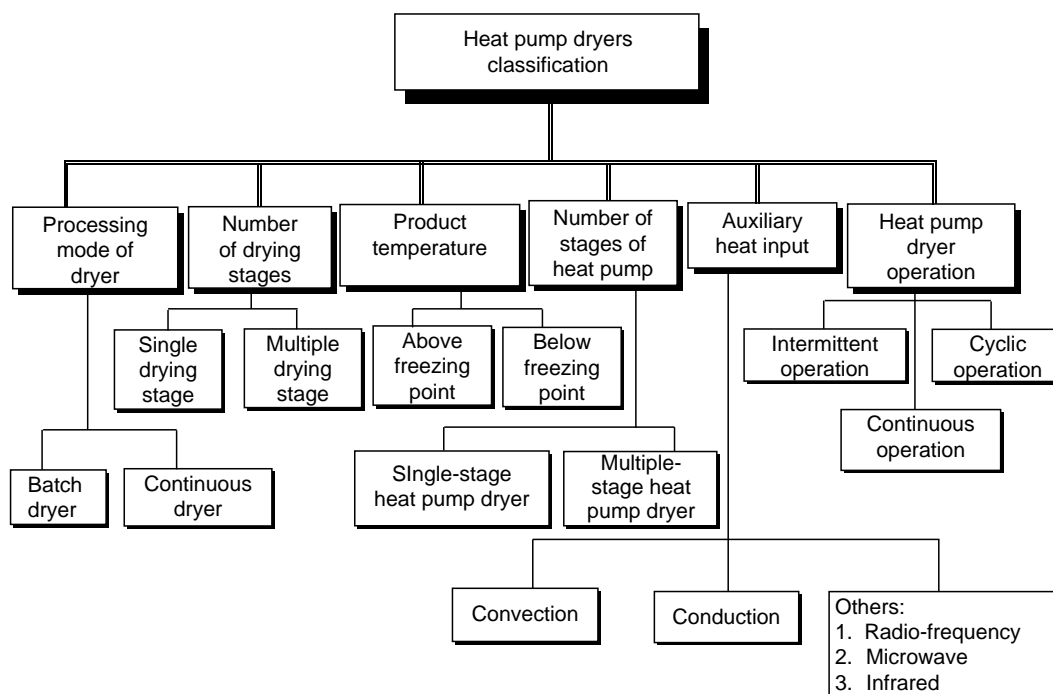


FIGURE 47.1 A generalized classification scheme for heat pump dryers.

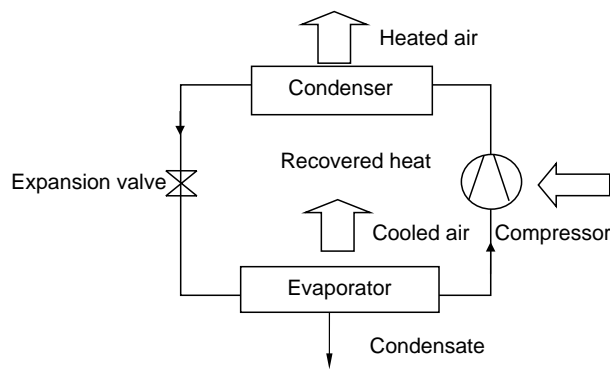


FIGURE 47.2 Schematic diagram of an air-to-air heat pump cycle.

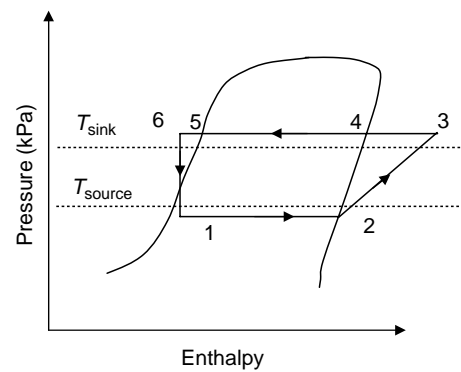


FIGURE 47.4 Pressure–enthalpy diagram of the heat pump cycle.

- input to the compressor is converted to shaft work to raise the pressure of the refrigerant vapor to that of the condenser at point 4. By increasing the vapor pressure, the boiling and condensing temperatures of the refrigerant is raised to a level higher than that of the heat sink temperature (the surrounding temperature). At this stage, the vapor is in a superheated state.
- After compression, the refrigerant vapor is directed to the condenser that is basically a heat exchanger to carry out the condensing process. The refrigerant first undergoes quick desuperheating change from superheated vapor to saturated vapor and then undergoes condensing process in the condenser. At the condenser, the refrigerant undergoes two-phase condensation, changing from vapor to liquid phase. During this process, heat is rejected by the condenser to heat the surrounding air.
 - Heat recovery occurs when the heat energy absorbed in the evaporator and the work energy

from the compressor is “pumped” to the condenser side for sensible heating of the air.

- Once the vapor refrigerant exits the condenser, it undergoes an additional stage of subcooling (point 5 to 6) in another heat exchanger. There are two advantages of subcooling. Firstly, additional heat can be recovered for sensible heating of the air. Secondly, it reduces flashing when the refrigerant pressure is reduced in the throttling device.
- After the condensing process, a throttling device such as a valve, orifice plate, or capillary tube is used to expand the liquid refrigerant in order to reduce the pressure of the refrigerant liquid level to a boiling temperature below that of the heat source. After the expansion process, the refrigerant enters the evaporator in a two-phase state.
- The entire cycle repeats itself.

47.3 PRINCIPLE OF HEAT PUMP DRYER

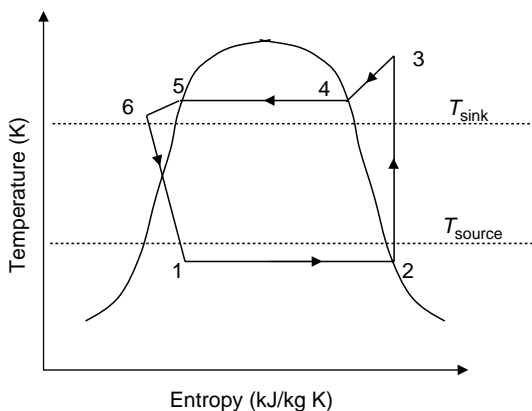


FIGURE 47.3 Temperature–entropy diagram of the heat pump cycle.

Figure 47.5 represents a schematic layout of various refrigeration components integrated with the drying chamber. The inlet drying air passes through the drying chamber at point 1 and picks up moisture from the product. The moisture-laden air at point 2 is then directed to the evaporator coil. Two types of evaporator systems exist. One is a direct expansion coil whereby the refrigerant undergoes a two-phase change from liquid to vapor to cool and dehumidify the air. The other is a chilled water system wherein the flow of chilled water to the coil is controlled for cooling and dehumidification. During the dehumidification process from point 2 to 3, the air is first cooled sensibly to its dew point. Further cooling results in condensation of water from the air. Latent heat of vaporization is then absorbed by the evaporator for

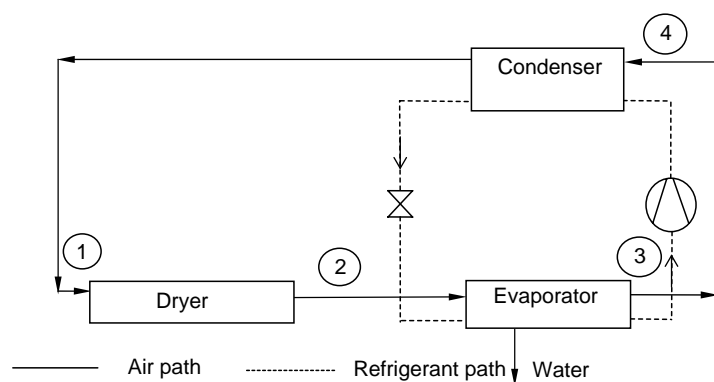


FIGURE 47.5 Schematic representation of heat pump drying system.

boiling the refrigerant. The recovered heat is “pumped” to the condenser. The cooled and dehumidified air then absorbs the heat at the condenser moving from point 4 to 1 for sensible heating to the desired temperature.

The energy efficiency of a heat pump is defined by the coefficient of performance (COP). COP is given by

$$\text{COP} = \frac{\text{useful heat output}}{\text{power input}} \quad (47.1)$$

The maximum theoretical heat pump efficiency is given by the Carnot efficiency as

$$\text{COP}_{\text{carnot}} = \frac{T_{\text{condenser}}}{T_{\text{condenser}} - T_{\text{evaporator}}} \quad (47.2)$$

The $\text{COP}_{\text{carnot}}$ cannot be realized physically but is used as a gauge to determine how far a refrigeration system is from the ideal system. In practice, the actual efficiency of the heat pump is usually 40 to 50% of the theoretical Carnot efficiency (Geeraert, 1976; Strumillo and Zylla, 1985).

A performance indicator that is commonly used to define the performance of the dryer is the specific moisture extraction rate (SMER).

SMER is defined as

$$\text{SMER} = \frac{\text{amount of water evaporated}}{\text{energy input to the dryer}}, \text{kg/kWh} \quad (47.3)$$

Alternatively, another parameter known as specific energy consumption (SEC) that is the reciprocal of SMER, can be used to compare energy efficiencies of different types of dryer. There is a relation between SEC and COP (Strumillo and Zylla, 1985) that is given by

$$\begin{aligned} \text{SEC} &= \frac{W}{G_M} = \frac{\frac{Q_{\text{ev}}}{(\text{COP})_A - 1}}{M_a \cdot (\omega_1 - \omega_2)} \\ &= \frac{h_1 - h_2}{((\text{COP})_A - 1)(\omega_1 - \omega_2)} \end{aligned} \quad (47.4)$$

47.4 ADVANTAGES AND LIMITATIONS

The key advantages and limitations of the heat pump dryer are:

Advantages:

1. Heat pump drying (HPD) offers one of the highest specific moisture extraction ratio (SMER), often in the range of 1.0 to 4.0, since heat can be recovered from the moisture-laden air.
2. Heat pump dryers can significantly improve product quality by drying at low temperatures. At low temperatures, the drying potential of the air can be maintained by further reduction of the air humidity.
3. A wide range of drying conditions typically -20°C to 100°C (with auxiliary heating) and relative humidity 15 to 80% (with humidification system) can be generated.
4. Excellent control of environment for high-value products and reduced electrical energy consumption for low-value products.

Limitations:

1. Chlorofluorocarbons (CFCs) are used in refrigerant cycles, some of which are not environment-friendly.
2. Requires regular maintenance of components (e.g., compressor and refrigerant filters) and charging of the refrigerant.
3. May incur higher capital costs.

47.5 ENERGY EFFICIENCY

The ability of heat pump dryers to convert latent heat of condensation into sensible heat at the hot condenser makes them unique heat recovering devices for drying applications. The energy efficiency of HPD can be reflected by the higher SMER values and drying efficiency when compared to other drying systems as shown in Table 47.1. Consequently, higher SMER would then be translated to lower operating cost, making the payback period for initial capital considerably shorter.

47.6 DRYING MODE AND DRYER CONFIGURATION

Heat pump dryer can operate in different modes, e.g., the batch (parallel and cross-flow) and continuous modes. Figure 47.6 shows a batch heat pump dryer. The products are placed on trays that are positioned in the drying chamber and removed once the desired product moisture content is reached. The drying air can flow parallel or perpendicular to the product surface. A cross-flow dryer configuration is shown in Figure 47.7. Batch drying is generally suitable for smaller production rates but entails higher labor cost.

Figure 47.8 shows a drying chamber designed to operate in continuous mode. The products are placed on trays positioned on the conveyor belt system. The speed of the conveyor can be varied with a controller-gears system. Continuous systems involve faster loading and unloading of the drying products and are less labor-intensive. The selection of the drying mode

TABLE 47.1
Comparing Heat Pump Drying with Other Drying Systems

Parameter	Hot Air Drying	Vacuum Drying	Heat Pump Drying
SMER (kg water/KWh)	0.12–1.28	0.72–1.2	1.0–4.0
Drying efficiency (%)	35–40	≤70	95
Operating temperature range (°C)	40–90	30–60	10–65
Operating % RH range	Variable	Low	10–65
Capital cost	Low	High	Moderate
Running cost	High	Very high	Low

Source: Adapted from Perera, C.O. and Rahman, M.S., *Trends Food Science Technology*, 8(3), 75, 1990. With permission.

largely depends on the drying characteristics of the product and the required product loading capacity.

47.7 MULTISTAGE HEAT PUMPS

Many of the commercial heat pump dryers consist of a single-stage vapor compression cycle. In such systems, only one evaporator is used for cooling and dehumidifying, and recovering the latent heat of vaporization from the drying air. A mechanical constraint is then imposed on the amount of heat recovered because of the physical area available for heat transfer. Further, the single-stage heat pump cycle is not able to produce several streams of drying air with different drying conditions, both in terms of

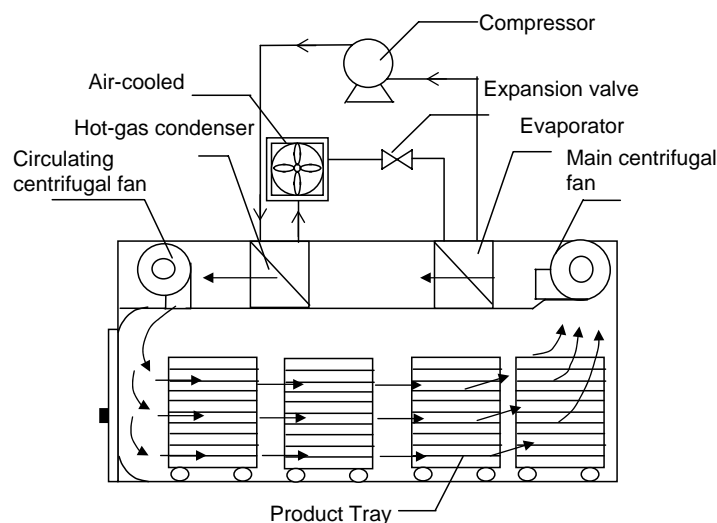


FIGURE 47.6 A parallel flow batch mode heat pump dryer.

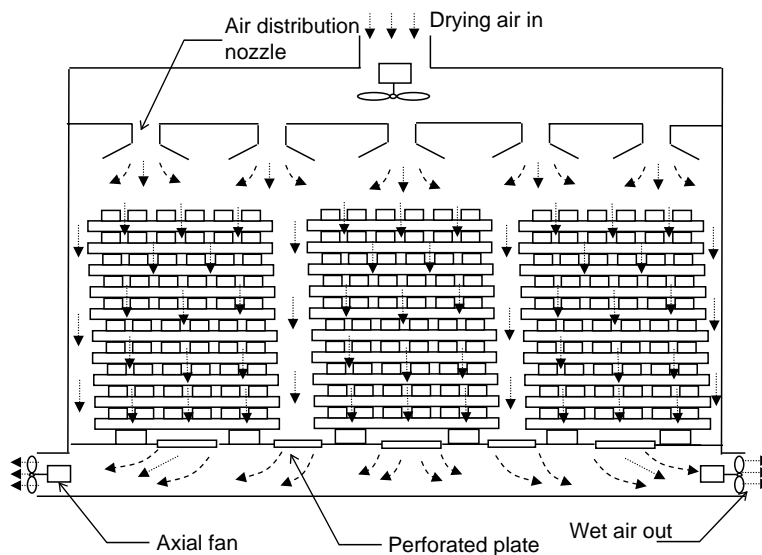


FIGURE 47.7 A perpendicular flow batch mode dryer.

temperature and humidity, to several independent drying chambers. [Figure 47.9](#) shows the improvement in energy efficiency for different heat pump configurations. It can be observed that there is a significant improvement in the heat pump energy efficiency from a single to two-stage dryer.

Multistage vapor-compression systems can be designed to integrate with the drying chambers as shown in [Figure 47.10](#). One additional advantage of the incorporation of a multi-stage heat pump cycle is that it allows a control mechanism for regulating the humidity of the air to be implemented.

In a two-stage heat pump dryer, the refrigerant vapor is split into two streams at the exit of the condenser. One stream enters an expansion valve at a higher discharge capacity to be regulated to the “low” evaporator temperature while the other enters another expansion valve to be expanded to a higher

temperature as shown in [Figure 47.11](#). At both high- and low-pressure evaporators, the evaporation processes take place. The pressure of the refrigerant vapor at the exit of the high-pressure evaporator is regulated by a backpressure regulator to that of the low-pressure evaporator before mixing takes place at a vapor chamber. The pressure of the mixed vapor is then raised by the compressor to that of the condenser. This two-stage cycle then repeats itself.

47.8 REAL-TIME CONTROL OF DRYING ENVIRONMENT

The complex chemical reactions involved in the destruction of heat-sensitive materials during drying are well documented. Optimization based on the reduction of quality degradation of such processes is difficult. The

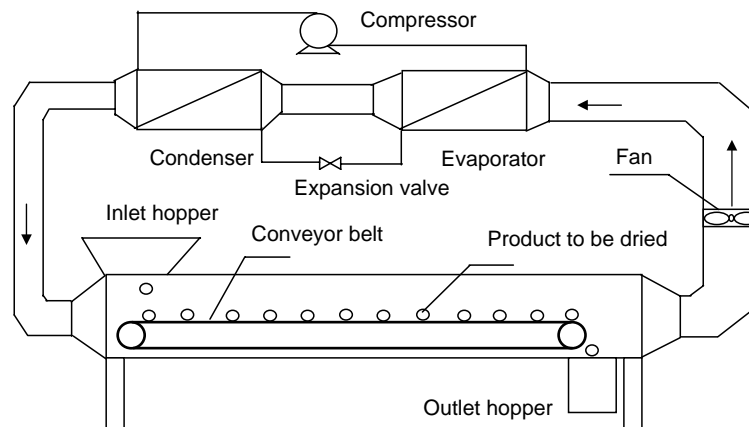


FIGURE 47.8 A continuous mode heat pump dryer.

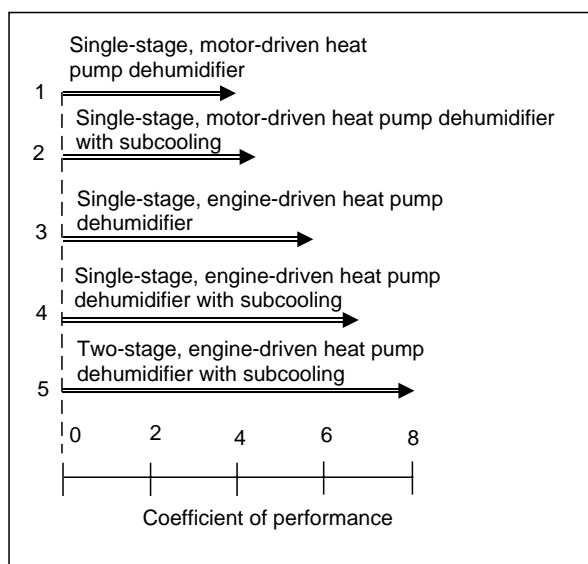


FIGURE 47.9 Coefficient of performance of various heat pump configurations. (Adapted from Perry, E.J., *Inst. Refrig. Mtg.*, 1, 1981.)

traditional approach in food technology is based on employing well-known technologists who conduct trial and error tests. Quite often, the task is time consuming and arduous. In most competitive industries, this method is no longer appropriate. Yet modern food technology makes it imperative that solutions be found that will allow the optimization of complex processes with respect to complex quality factors (Karel, 1988).

On the basis of the state-of-the-art technology, the direction toward solution to this problem lies in the combinations of line-sensors and expert systems with feedback response to allow immediate quality-related decision to be made. The sensors are placed in strategic locations to measure real-time quality parameters. The signals are then fed to expert systems, usually comprising a software system that has the ability to receive and transmit decision signals to controllers. It is well known that the quality degradation of food products, such as browning effects and ascorbic acid (AA) degradation, is mainly due to the thermal effect of the drying air. It is thus possible to reduce these quality effects through a proper feedback system to regulate the air or product temperature.

Figure 47.12 shows an example of a real-time process control strategy for a heat pump dryer to improve the product color and minimize surface cracking through time variation of the drying air temperature. A thermovision camera is used to capture the surface temperature profiles. On the basis of predefined constraints on the surface temperature, a signal from the computer is sent to the PID controller to tune the temperature of the drying air. In this way, the quality degradation of the product can be minimized without compromising the drying rate excessively to achieve the desired final moisture content.

Figure 47.13 shows another example of a real-time process control for the heat pump dryer to reduce nutrient degradation. Experiments have been carried out with hypodermic thermocouple needles to measure

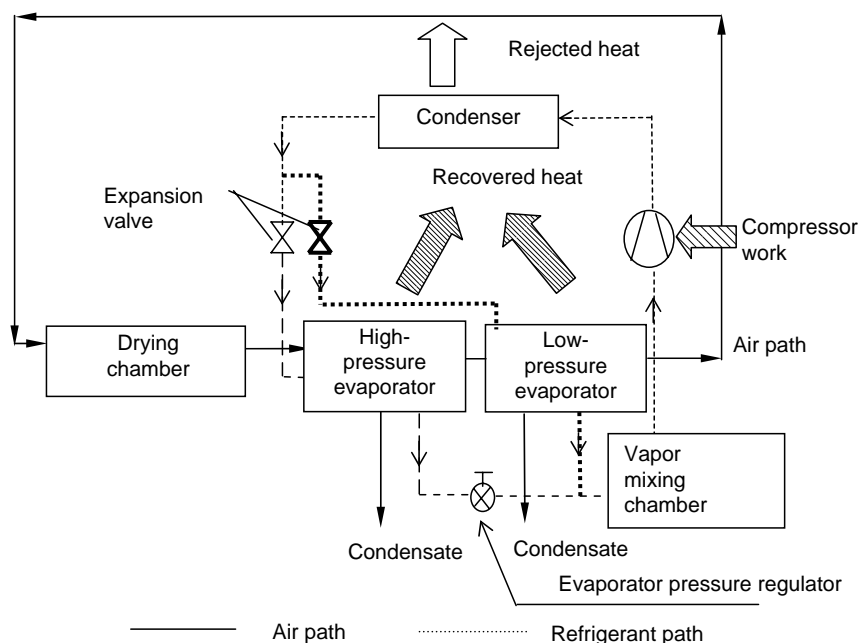


FIGURE 47.10 Two-stage heat pump system coupled with a drying chamber.

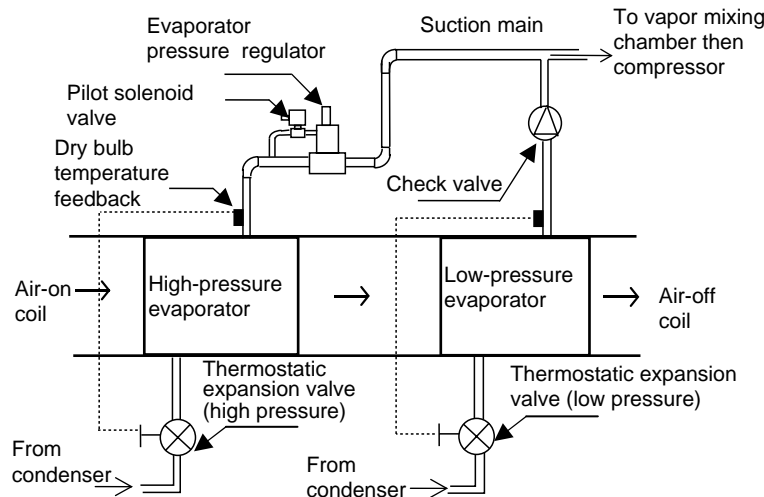


FIGURE 47.11 The refrigeration flow of a two-stage heat pump cycle.

the transient temperature profiles of food products (Chou et al., 1997). These measured values make it possible to tune the drying air temperature to prevent the internal product temperature from reaching a threshold value, hence reducing thermally induced nutrient degradation.

47.9 FIXED AND TIME-VARIABLE OPERATING SCHEMES

In order to reduce the energy consumption per unit of product moisture, it is necessary to examine different

methodologies to improve the energy efficiency of the drying equipment. One possible method is to apply time-dependent drying schemes to reduce the drying time to obtain the desired product moisture content. Time-dependent drying schemes which imply time-varying supply of thermal energy for drying in the batch mode can be classified into the following categories:

1. Intermittent drying whereby heat is supplied intermittently rather than continuously. This can be achieved by interrupting the airflow to the product or by intermittently heating the drying air.

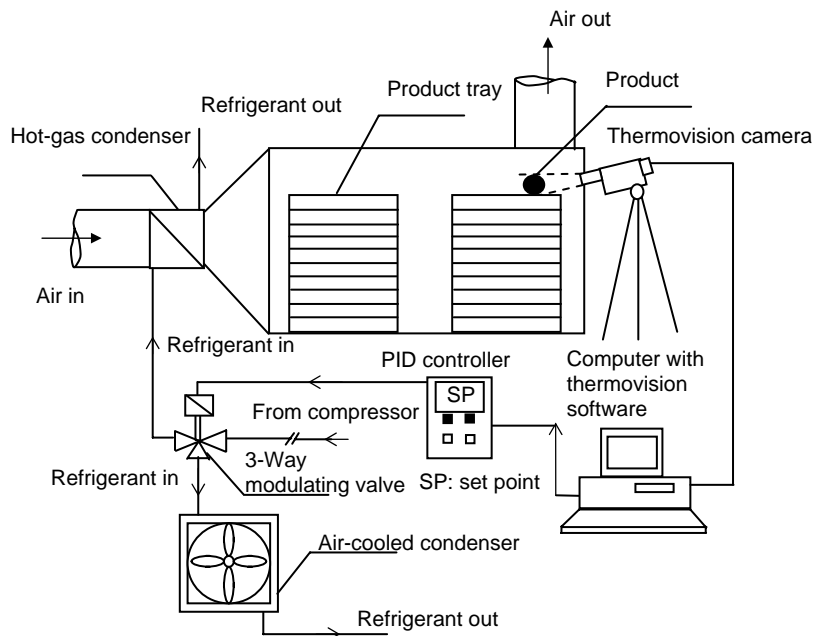


FIGURE 47.12 An online feedback system for a heat pump dryer to control the color of the product.

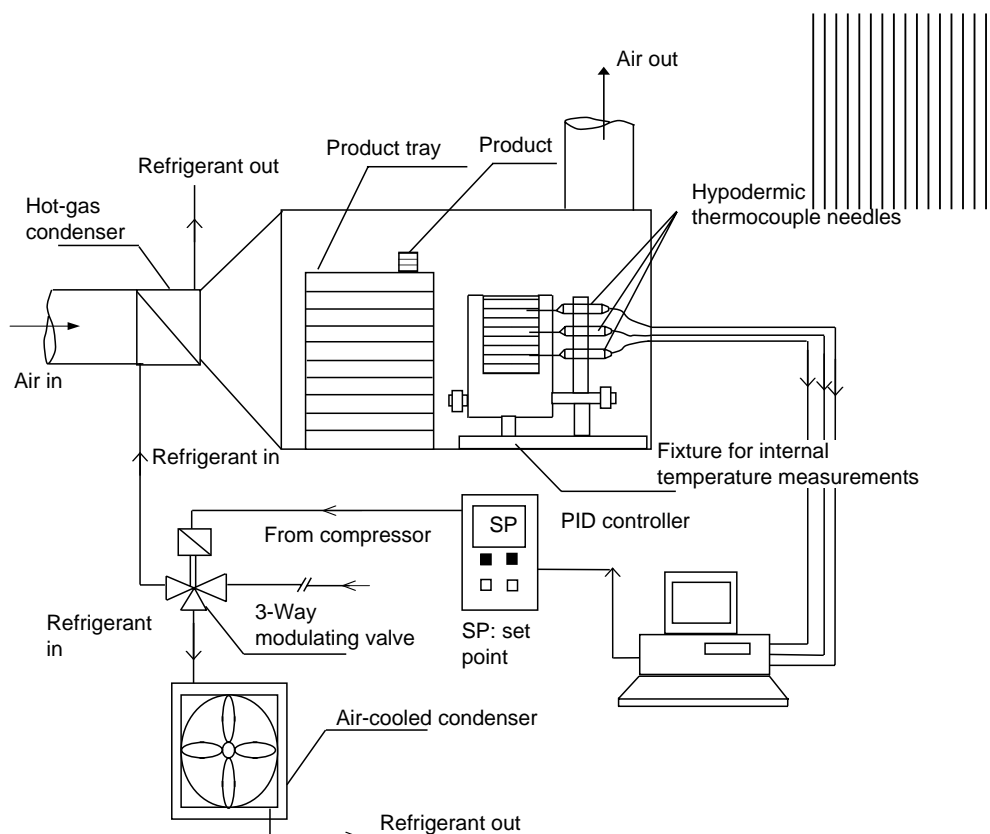


FIGURE 47.13 An on-line feedback system for a heat pump dryer to reduce nutrients degradation.

2. Dryaeration which is a drying process involving a combination of high temperature, short drying period, tempering, and slow cooling concluded by final drying.
3. Air reversal drying which is reversing the direction of the airflow for a period of time and then revert it back to its original direction. This is applied to deep bed drying of particulates.
4. Cyclic drying which is a drying process whereby the air temperature, humidity, or even velocity undergoes a specified cyclic pattern variation such as sinusoidal, square-wave, or sawtooth patterns.

Several experimental studies have been carried out to investigate various time-dependent drying schemes and their impacts on dryer energy consumption and product quality. These studies, summarized in [Table 47.2](#), have found several interesting features of time-dependent drying. These features are:

- Thermal energy savings
- Shorter effective drying time
- Higher moisture removal rates
- Lower product surface temperature

- Higher product quality; these include reduced shrinkage, cracking and brittleness, improved color, and nutrient retention

Studying the effect of regulating the airflow on the drying process, Ratti and Mujumdar (1993) presented a simulation study on batch drying of shrinking hygroscopic materials in a fixed bed using time-varying flow rate. Their work has shown that the total air consumption for drying is reduced with minor or no increase in drying time. Later, Ratti and Mujumdar (1995) extended their study to include the case of air-flow reversal. The results showed that both moisture and temperature profiles in the bed were flatter when air reversal was applied to the drying process while the mean drying curves remained practically unchanged.

Based on the above literature reviewed, it can be concluded that intermittent and time-varying drying have significant advantages in terms of reducing the required drying energy and enhancing the product quality of heat sensitive products.

In the next few sections, the recent results on HPD will be presented to demonstrate the advantages of employing time-varying drying schemes over fixed operating drying conditions.

TABLE 47.2
Summary of Different Time-Dependent Drying Studies

Study	Material and Dryer Type	Drying Scheme
Sabbah et al. (1972)	Corn (thin layer)	Dryaeration: tempering periods: 0–4 h
Troger and Butler (1980)	Peanuts	Intermittent drying: airflow interrupted at 1 in 4 h
Harnoy and Radajewski (1982)	Maize (bin dryer)	Intermittent drying: aeration periods: 1–6 min; rest periods: 3–90 min
Giowacka and Malczewski (1986)	Wheat (fluidized bed)	Sinusoidal heating
Hällstrom (1986)	Compound fertilizer (fluidized bed)	Intermittent drying: drying periods: 2.5–6 s; rest periods: 4.5–6 s
Zhang and Litchfield (1991)	Corn (thin layer)	Intermittent drying: drying period: 20 min; rest periods: 0–120 min
Hemati et al. (1992)	Corn (flotation fluid bed)	Intermittent drying: drying period: 20 min rest; periods: 0–60 min

47.9.1 DRYING WITH FIXED OPERATING CONDITIONS

Most of the industrial drying operations are carried out under fixed drying conditions, i.e., the temperature, humidity, and velocity of the airflow are kept relatively constant throughout the entire drying process until the desired product moisture content has been attained. Many of these fixed drying conditions have been obtained from several trials until the product dryness and desired quality have been obtained. Prescribing a fixed drying condition would indeed simplify the drying process. Often the optimal set of fixed operating parameters cannot be obtained and since the intrinsic properties of the material change as moisture is removed and the material gets heated up, what may be optimum drying conditions for initial drying stages may not be optimum for later stages. With the development of advanced controllers and process control techniques that can incorporate different product quality constraints in the feedback control strategy, drying conditions used in HPD can now be “tuned” to produce a few “constant” drying conditions at different stages of the drying. Therefore, several fixed operations can be incorporated resulting in a series of fixed drying operations with favorable results in drying kinetics as well as product quality.

47.9.2 DRYING WITH INTERMITTENT OPERATION

Three intermittent HPD patterns were used in the present study and their prescriptions are shown in Figure 47.14. These intermittent profiles are prescribed by raising the inlet air temperature for a defined period (τ) and dropping the air temperature back to its original level until the periodic interval. The intermittency, α , is defined as the fraction of time during which the inlet air temperature is raised to the defined cycle time, i.e.,

$$\alpha = \frac{\tau_{\text{on}}}{\tau} = \frac{\tau_{\text{on}}}{\tau_{\text{on}} + \tau_{\text{off}}} \quad (47.5)$$

47.9.3 DRYING WITH CYCLIC OR ARBITRARY TIME-VARYING OPERATION

Devahastin and Mujumdar (1999) have demonstrated by a mathematical model the feasibility and advantages of operating a dryer by varying the temperature of the inlet drying air in terms of reducing drying time by up to 30%. As technology advances, more options are available to improve the quality. One potential avenue in improving quality degradation in food products during drying is to employ time-varying temperature profiles that minimize quality degradation and dry the products to the desired moisture content within an allowable production time. Several researchers have studied the degradation of quality of dried products under sine or square wave temperature fluctuations (Wu et al., 1974; Kamman et al., 1981) during storage. However, little work has been reported to study the effect of temperature profiles on quality during convective drying process. In the following section, we examine several time-varying drying operations and their impacts on drying kinetics and product quality.

47.9.4 DRYING KINETICS

47.9.4.1 Intermittent Operation

The limited study on intermittent drying of materials in a batch dryer have confirmed the potential advantages of time-dependent supply of energy for drying, e.g., reduced energy consumption, reduced air consumption, and enhanced quality of heat-sensitive products.

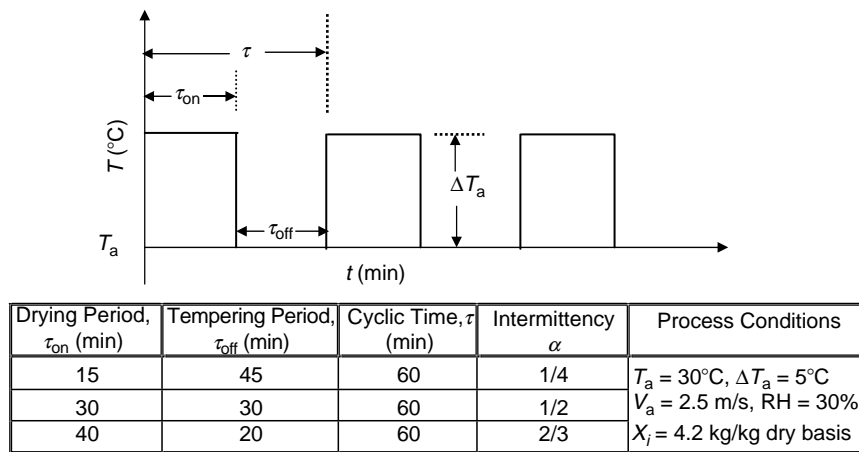


FIGURE 47.14 Intermittent varying of inlet air temperature profiles. (Adapted from Chou, S.K., Chua, K.J., Hawlader, M.N.A, Mujumdar, A.S., and Ho, J.C., *Transactions of the Institution of Chemical Engineers*, 78(C), 193, 2000. With permission.)

For HPD systems, intermittent drying can be classified into three categories:

1. Intermittent regulation of the air temperature
2. Intermittent supply of airflow
3. Intermittent regulation of the air humidity

Among the three, the intermittent regulation of air temperature is considered to have the most significant influence on the product drying kinetics and various quality parameters and is the focus of the following sections.

47.9.4.2 Mean Moisture Content

Figure 47.15 shows the mean moisture content of the potato samples conducted under each intermittent

drying profiles and continuous drying process of temperature 35°C . It is evidenced that employing intermittent temperature drying profile can result in saving drying time by about 25, 48 and 61% for $\alpha = 1/4$, $\alpha = 1/2$, and $\alpha = 2/3$, respectively. Longer product tempering period occurs for $\alpha = 1/4$, resulting in a flat region in the drying curve. It can also be observed from Figure 47.15 that, during each tempering process, i.e., during τ_{off} period, some form of partial drying occurs, particularly during the earlier stage of drying. However, toward latter stage, little or no drying occurs as the mean moisture content curve tapered following an asymptotic value. Therefore, it can be inferred from these curves that as substantiate amount of moisture is removed during the last phase of drying, product tempering enables to bring about

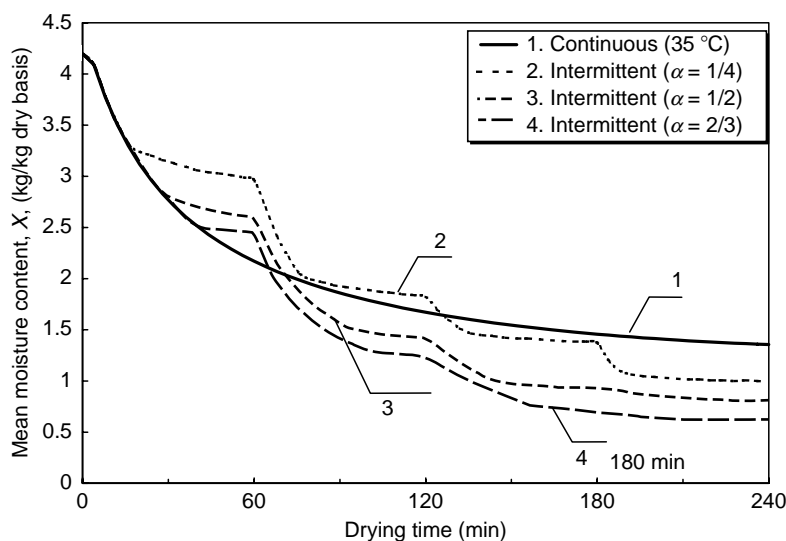


FIGURE 47.15 Time-evolution of the average moisture content for various drying profiles. (Adapted from Chou, S.K., Chua, K.J., Hawlader, M.N.A, Mujumdar, A.S., and Ho, J.C., *Transactions of the Institution of Chemical Engineers*, 78(C), 193, 2000. With permission.)

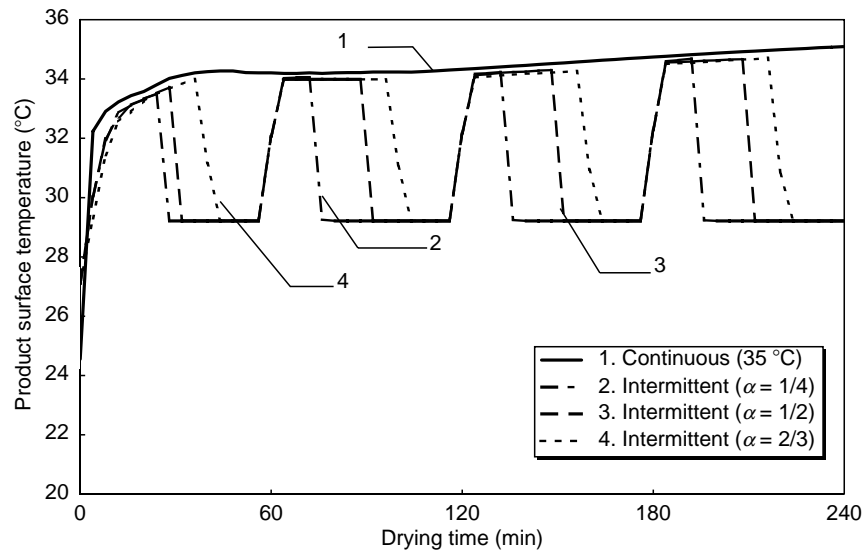


FIGURE 47.16 Time-evolution of product surface temperature for various drying profiles. (Adapted from Chou, S.K., Chua, K.J., Hawlader, M.N.A, Mujumdar, A.S., and Ho, J.C., *Transactions of the Institution of Chemical Engineers*, 78(C), 193, 2000. With permission.)

uniform moisture distribution within the product rather than reducing moisture content.

47.9.4.3 Surface Temperature

Figure 47.16 shows that intermittent heating about the mean temperature, $T_m = 30^\circ\text{C}$, can produce lower surface temperature than the continuous one. This might be of interest in the drying of sensitive materials, particularly those of biological nature (e.g., marine and agricultural products). Gentle time-varying heating using these heating schemes offer greater flexibility in the control of the surface temperature and minimize many product quality problems such as nonenzymatic browning (NEB) and surface cracks. It is interesting to note that surface temperature due to continuous temperature drying envelops those surface temperature profiles due to time-temperature varying schemes especially during the initial stage of drying. It is also expected that a longer surface tempering period at cooler temperature is observed for $\alpha = 1/4$. A combination of lower surface temperature and longer tempering period might be the method to enhance the product quality related to product surface conditions.

47.9.4.4 Surface Moisture Content

Figure 47.17 shows the evolution of the surface moisture content under different intermittent temperature drying. On the onset of each tempering period, the surface moisture content increases. The increment is particularly prominent for the second tempering period. For the first tempering period, the product surface is still moisture saturated and vapor flux continues to be transferred from the surface to the

air. Once the surface becomes partially dried, internal moisture moves to the surface and an immediate increment in surface moisture is observed. There may be an increased moisture flux from the internal to the surface as drying progresses. When the tempering period is longer, the allowance time for the new surface moisture to form is also longer. Therefore, a gentle raise in surface moisture content can be observed when compared with the samples with surfaces subjected to a longer heating period ($\alpha = 1/2$ and $\alpha = 2/3$). Once tempering is over, the surface moisture resumes its drying process and moisture is rapidly removed again. According to Jumah et al. (1996), the effect of introducing cooler air, in addition to partial drying, is able to repeat the initial steep drying curve. Consequently, high drying rate follows after each heating period. This favorable supply of more surface moisture during the tempering period enhances the removal of moisture from the product and, eventually, reduces the time to obtain the desired moisture content.

47.9.5 CYCLIC OR ARBITRARY TIME-VARYING OPERATION

The evolution of the mean moisture content with time for drying of banana samples in a heat pump dryer is shown in Figure 47.18. It can be observed that step-down temperature profile is able to reduce the drying time to reach desired moisture content. Taking the moisture content of each product at 240 min as the basis for comparison, it can be observed that the step-down temperature profile was able to reduce the drying time for banana samples by 180 min. A noteworthy point is that step-down temperature profile was more effective in reducing in the drying time for banana samples.

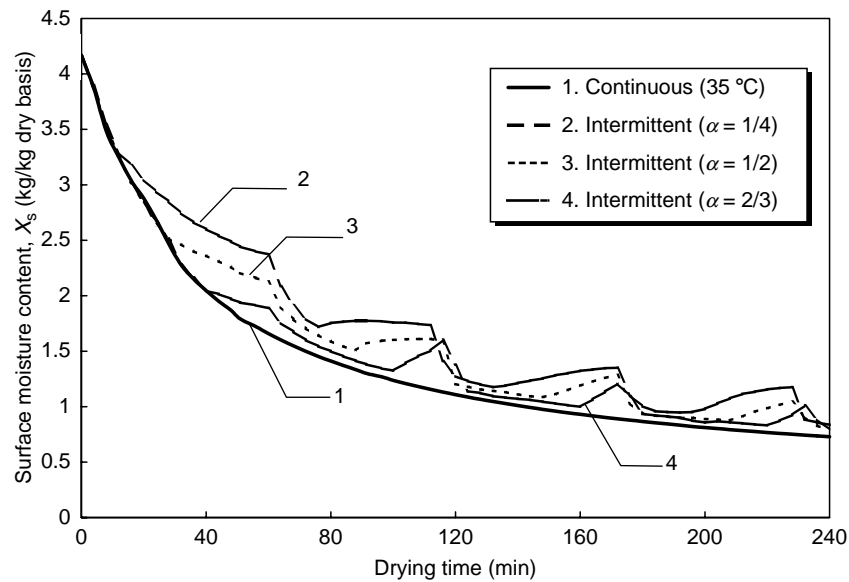


FIGURE 47.17 Time-evolution of product surface moisture for various drying profiles. (Adapted from Chou, S.K., Chua, K.J., Hawlader, M.N.A, Mujumdar, A.S., and Ho, J.C., *Transactions of the Institution of Chemical Engineers*, 78(C), 193, 2000. With permission.)

The drying rate curves for the banana samples are portrayed in Figure 47.19. It can be observed from this figure that stepwise variation of the air temperature produces unusual drying rates. For the step-down temperature profile, two conventional, i.e., first and second, falling drying rate curves exist in tandem. Such findings maybe attributed to rapid removal of moisture from the saturated surface during the initial drying stage, followed by the high temperature gradient between product surface and internal once a drop in air temperature was initiated. Two drying rate peaks can be detected for the step-up

temperature profile. Due to the slow initial drying resulting rate from reduced sensible heat transferred, the samples with high moisture content were still able to experience increased drying rate when temperature step increment of 5°C was implemented.

47.9.6 PRODUCT QUALITY ASPECTS

The impact of constant temperature drying on product quality is well known. Most of the product quality parameters such as NEB and AA content are often manifested by a progressive loss with increasing

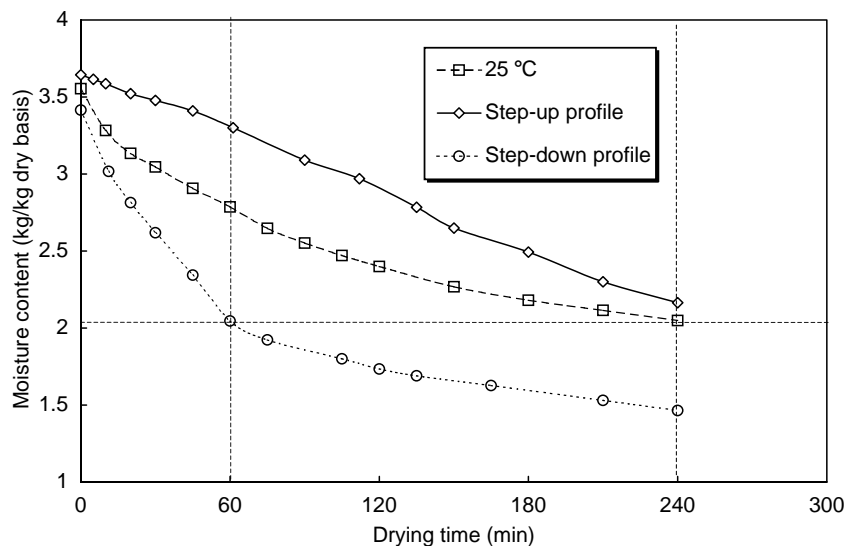


FIGURE 47.18 Moisture content of banana samples versus drying time. (Adapted from Chua, K.J., Mujumdar, A.S., Hawlader, M.N.A, Chou, S.K., and Ho, J.C., 2001, *Food Research International*, 34, 721, 2001. With permission.)

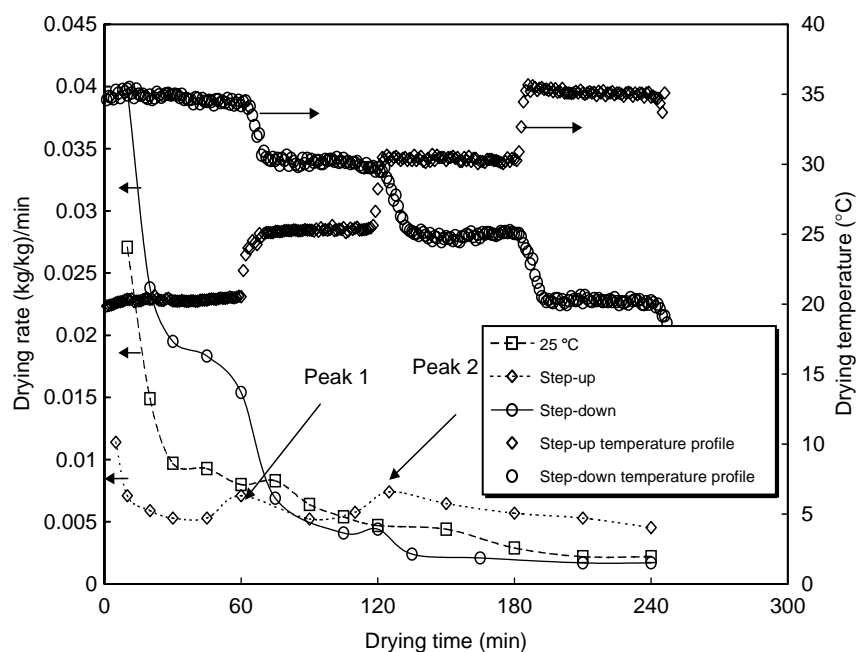


FIGURE 47.19 Drying rate of banana samples undergoing stepwise varying of drying air temperature. (Adapted from Chua, K.J., Mujumdar, A.S., Hawlader, M.N.A, Chou, S.K., and Ho, J.C., 2001, *Food Research International*, 34, 721, 2001. With permission.)

temperature. The following sections will discuss the impact of employing the different time-varying drying profiles on some product quality parameters during HPD.

47.9.6.1 Color Change due to Nonenzymatic Browning

Chua et al. (2000a) have demonstrated that a two-stage heat pump dryer can be controlled to produce prescribed time-varying air temperature profiles to study the effect of nonuniform temperature drying on color change of food products. They have also shown that by subjecting food products to different temperature profiles in a heat pump dryer, it is possible to reduce the change in individual color parameters as well as in the overall color change in the food products. High-sugar content product such as banana favors a time-varying profile with a cold starting temperature of 30°C while high-moisture product like potato with low sugar content enabled the use of higher temperature profiles to yield higher drying rates without any pronounced change in the overall color change. Prescribing the appropriate cyclic temperature variation schemes, Chua et al. (2000a) have shown that the percentage reductions in overall color change for potato, guava, and banana were 87, 75, and 67%, respectively.

47.9.6.2 Ascorbic Acid Content

On the basis of an extensive experimental study of the kinetics of batch drying and AA degradation of guava

pieces under isothermal as well as time-varying drying air temperature, Chua et al. (2000b) have shown that with proper selection of the temperature schedule, the AA content of the guava pieces can be up to 20% higher than that in the isothermal drying without significant enhancement in drying time. Results from Chua et al. (2000b) indicate that employing reduced air temperatures at the onset of drying followed by temperature elevation as drying proceeds yield a better quality product. Recently, Pan et al. (1999) have clearly demonstrated the advantage of intermittent drying as far as product quality is concerned. They have shown that in vibrated bed batch drying of carrot pieces the retention of beta-carotene in the dried product is higher in intermittent drying while at the same time the net energy consumption is reduced and even the actual drying time is shortened somewhat.

47.9.6.3 Other Quality Parameters

Ginger dried in a heat pump dryer was found to retain over 26% of gingerol, the principal volatile flavor component responsible for its pungency, compared with the rotary dried commercial samples that have only about 20% (Mason et al., 1994). The higher volatile retention in heat pump-dried samples is probably due to reduced degradation of gingerol when lower drying temperatures are employed compared with higher commercial dryer temperatures. Since HPD is conducted in a closed chamber, any compound that volatilizes will remain within it, and the partial

pressure for that compound will gradually build up within the chamber, retarding further volatilization from the product (Perera and Rahman, 1990).

The color and aroma herbs (e.g., parsley, rosemary, and sweet fennel) can be improved when compared with the commercial products. The sensory values were nearly doubled in case of heat pump-dried herbs compared with commercially dried products. There was no significant difference in the quality of herbs dried below moisture content of 0.04 for experimental drying temperatures (40 and 50°C) and relative humidity (0.30 and 0.40).

The use of modified atmospheres for drying sensitive materials such as food products is another important potential aspect of the HPD technology. Drying with oxygen-sensitive materials such as flavor compounds and fatty acids can undergo oxidation, giving rise to poor flavor, color, and rehydration properties. Use of modified atmospheres to replace air would allow new dry products to be developed without oxidative reactions occurring (Perera and Rahman, 1990).

47.10 SELECTED INDUSTRIAL APPLICATIONS

Heat pump dryers are gaining recognition as energy efficient drying devices for production of quality products. The following sections are devoted to selected examples of industrial applications of the heat pump dryer.

47.10.1 HEAT PUMP DRYERS FOR TIMBER DRYING

HPD can provide efficient and cost-effective drying of timber, particularly where quality is a key issue. In the last 10 y, the industry emphasis has been on increasing throughput through higher temperature processing. With increasing concerns about the ability of the international market to absorb lower grade timber, the cycle is returning to examine the need to produce a wider range of product types of acceptable quality. In this market, HPD can be made competitive, as it has distinct cost advantages over conventional heat-and-vent kilns at lower operating temperatures and where higher humidity levels need to be maintained (Bannister et al., 1999). Furthermore, heat pump dryers operate at higher energy efficiency when the amount of water removed increases. Table 47.3 shows the performance of a heat pump dryer for timber drying conducted at the Forest Education Center in Rotorua (Bannister et al., 1999). It can be observed that the dryer performed with better energy efficiency (measured by SMER) when more water was removed. It is expected because of the higher amount of latent

TABLE 47.3
Energy Performance of a Heat Pump Dryer for Timber Drying

SMER (kg/kWh)	Water Removed (kg)
2.2	1300
2.6	1750
2.9	5200
3.2	7200

Source: Bannister, P., Carrington, G., Chen, G., and Sun, Z., *Energy Group's Heat Pump Dehumidifier Research Programme Report*, EGL-RR-02, ed. 1.1, 1999.

heat recovered when higher amount of water is removed. This would translate to a shorter payback period for higher volume of timber dried.

The advantages of HPD for timber include efficient utilization of recovered heat, and slow and controlled drying rates resulting in reduction of physical defects. It is also possible to accomplish simultaneous drying of different wood species in the same kiln in low-temperature drying. Figure 47.20 shows the schematic of a heat pump dryer designed for timber drying.

According to Bannister et al. (1999), timbers that are physically slow to dry, tend to warp, split, or collapse during drying. These timbers are classified under “hard-to-dry” woods. Many of these timbers have high commercial value and so there is a need to establish a drying schedule that is able to produce consistent high quality results. Examples of hard-to-dry timbers in Australia include Red Beech (*Notofagus fusca*), Hard Beech (*Notofagus truncata*), and many *Eucalyptus* species (Bannister et al., 1999).

Bannister et al. (1999) proposed the following schedules to improve timber quality by HPD:

- Initial drying at low temperature (less than 30°C) and high humidity (85% and higher)
- A ramping period during which air temperature is increased and its humidity is lowered
- A finishing period at a reasonable temperature (~50°C) and lower humidity (40 to 50%)

Potential benefits of HPD of timber that can be reaped include (Bannister et al., 1999):

1. Improved quality—this is important since the present timber market is now driven by quality
2. Reduced drying time resulting in enhanced productivity.
3. Low energy costs—with energy consumption of around 2 to 2.5 kg/kWh. This is comparable

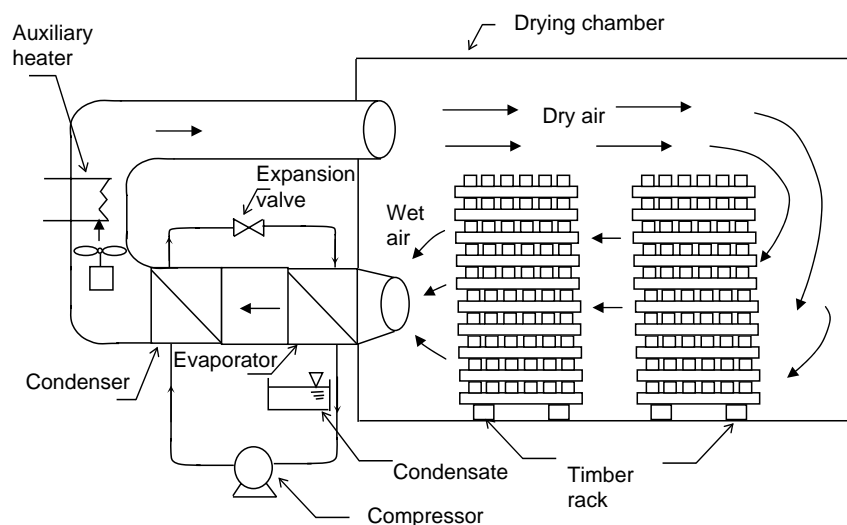


FIGURE 47.20 Schematic diagram of a heat pump dryer for timber drying.

with efficiencies for dehumidifiers drying easy-to-dry timbers.

4. Increased throughput by using air with lower humidity to enhance drying rate with little impact on drying-induced stresses.

47.10.2 HEAT PUMP DRYERS FOR FOOD PRODUCTS

There has been a growing interest in recent years in applying HPD technology to foods and biomaterials where low-temperature drying and well-controlled drying conditions are required to enhance the quality of food products. High-value products, which are extremely heat-sensitive, are often freeze-dried. This is an extremely expensive drying process (Baker, 1997). Recently, there has been great interest to look at the HPD system as a substitute system for freeze-dried products. Table 47.4 presents a summary of recent work on HPD of selected food products.

In most of the research studies presented in Table 47.4, the common conclusion was that the HPD offers products of better quality with reduced energy consumption. This is particularly true of food products that require precisely controlled drying atmosphere (temperature and humidity). Heat-sensitive food products, requiring low-temperature drying, can take the advantage of HPD technology since the drying temperature of HPD system can be adjusted from -20 to 60°C . With proper control, it is also possible for HPD to produce freeze-drying conditions at atmospheric pressure (Prasertsan and Saen-saby, 1998b). As far as food drying is concerned, HPD offers an alternative to improve product quality through proper regulation of the drying conditions. Chua et al. (2000a) have demonstrated that HPD can produce preselected cyclic temperature schedules to improve the

quality of various agricultural products dried in the two-stage HPD. They have shown that with appropriate choice of temperature–time variation, it is possible to reduce the overall color change and AA degradation by up to 87 and 20%, respectively.

The ability of HPD to regulate drying conditions quickly is another advantage that it offers for food drying. In countries where the level of the air humidity is high, high spoilage rates occur during the rainy season when the drying air is very moist. Clearly, HPD can reduce product spoilage by maintaining the humidity of the drying air through the regulation of latent heat removal at the evaporator.

Besides yielding better food quality, Rossi et al. (1992) has reported that onion slices dried by HPD used less energy in comparison to a conventional hot air system. Food products with high water content can be dried efficiently with HPD. As the drying air absorbs more of this available energy, this latent heat energy can be transferred at the evaporators for higher heat recovery. Lower energy input is then required at the compressor to enable sensible heating of the air when it passes through the condenser.

To summarize, when the quality of dried food products is paramount, HPD offers an attractive option to enhance product quality and reduces spoilage through better regulation of the drying conditions.

47.11 HEAT PUMP DRYER INCORPORATING OTHER TRANSFER MECHANISMS

Thus far, the research work has been primarily aimed at the performance of heat pump dryers for different products. Little work has been reported on the performance

TABLE 47.4
Recent Work Conducted on Heat Pump Drying of Selected Food Products

Researchers	Application(s)	Conclusions
Chou et al. (1997, 1998) Chua et al. (2000a)(Singapore)	Agricultural and marine products (Mushrooms, fruits, sea-cucumber, and oysters)	The quality of the agricultural and marine products can be improved with scheduled drying conditions
Prasertsan and Saen-saby (1998a) and Prasertsan et al. (1997) (Thailand)	Agricultural food drying (Bananas)	HPD is suitable for drying high moisture materials and the running cost of HPD is cheap making them economically feasible
Theerakulpisut (1990) (Australia)	Grain	An open cycle HPD performed better during the initial stage when the product drying rate is high
Meyer and Greyvenstein (1992) (South Africa)	Grain	There is a minimum operating period that makes the HPD more economical than other dryers
Rossi et al. (1992) (Brazil)	Vegetable (Onion)	Drying of sliced onions confirmed energy saving of the order of 30% and better product quality due to shorter processing time
Strømme and Kramer (1994) (Norway)	Marine products (Fish)	The high quality of the dried products was highlighted as the major advantage of HPD and introducing a temperature-controllable program to HPD makes it possible to regulate the product properties such as porosity, rehydration rates, strength, texture, and color

of heat pump dryers using external energy sources to compliment it. HPD is primarily a convective drying process. Use of infrared (IR) or radio frequency (RF) sources can be incorporated along with the usual HPD to enhance the drying rates while reducing the thermal load on the heat pump itself.

In the following sections, we look at several potential areas of supplementary heat sources to further enhance the performance of the conventional HPD in terms of lower energy consumption, better product quality, and enable the removal of bound water in thick materials.

47.11.1 FLUIDIZED BED HEAT PUMP DRYER

Fluidized bed drying (FBD) has found many applications for drying of granular solids in the food, ceramic, pharmaceutical, and agriculture industries. For drying of powders in the 50 to 2000 μm range, FBD competes successfully with other more traditional dryer types, e.g., rotary, tunnel, conveyor, and continuous tray. FBD has the following advantages (Mujumdar and Devahastin, 1999):

1. High drying rates due to excellent gas-particle contact leading to high heat and mass transfer rates
2. Smaller flow area
3. Higher thermal efficiency
4. Lower capital and maintenance costs compared with rotary dryers
5. Ease of control

However, FBD suffers from certain limitations such as:

1. High power consumption due to the need to suspend the entire bed in gas phase leading to high pressure drop
2. High potential of attrition, in some cases granulation or agglomeration
3. Low flexibility and potential of defluidization if the feed is too wet

A schematic layout of a fluidized bed heat pump dryer developed at the Norwegian Institute of Technology (NTNU) is shown in [Figure 47.21](#) (Strømme and Jonassen, 1996; Alves-Filho and Strømme 1996). The drying chamber receives wet material and discharges dried product through the product inlet and outlet ducts. The desired operating temperature is obtained by adjusting the condenser capacity, while the required air humidity is maintained by regulating the compressor capacity by frequency control of the motor speed. According to Alves-Filho and Strømme (1996), this setup can produce drying temperatures from -20 to 60°C and air humidities from 20 to 90%. With these features, heat-sensitive materials can be dried under convective air or freeze-drying conditions. It is also possible to sequence these two operations (convective and freeze drying). It will be advantageous for drying of food and bioproducts since freeze drying causes minimal shrinkage but produces low drying rates while convective air drying can be applied to

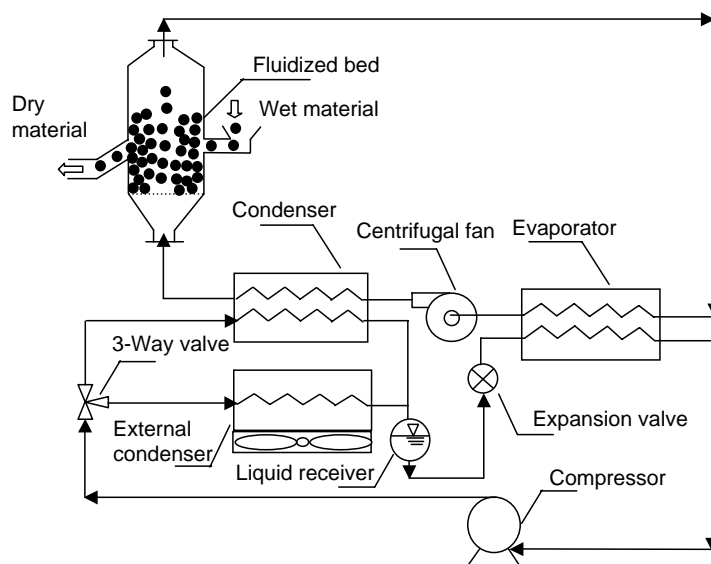


FIGURE 47.21 A schematic layout of a fluidized bed heat pump dryer. (Adapted from Alves-Filho, O. and Strømme, I., *Drying '96*, Strumillo, C. and Pakowski, Z. (Ed.), Krakow, Poland, 1996.)

enhance drying rates. Therefore, a combination of drying processes, e.g., freeze drying at -5°C followed by convective drying at 20 to -30°C , enables the control of quality parameters such as porosity, rehydration rates, strength, texture, color, and taste (Alves-Filho and Strømme, 1996). Experiments performed at NTNU on various heat-sensitive materials such as pharmaceutical products, fruits, and vegetables have shown that heat pump FBD offers a better product quality, but at higher cost. Since this technique produces a premium quality product, the incremental increase in drying cost may be offset by the higher market value of the product.

47.11.2 INFRARED-ASSISTED HEAT PUMP DRYING

Infrared drying helps to reduce the drying time by providing additional sensible heating to expedite the drying process. IR energy is transferred from the heating element to the product surface without heating the surrounding air (Jones, 1992). Several researchers have demonstrated the significant advantages of IR drying. These advantages (Navarri et al., 1992) include:

1. High heat transfer rates (up to 100 kW m^{-2} in paper industry) can be obtained with compact heaters.
2. Easy to direct the heat source to drying surface.
3. Quick response times, allowing easy and rapid process control (if needed).
4. Incorporating IR into an existing heat pump dryer is simple and capital cost is low.

IR drying has been the subject of investigations of recent researchers. Works by Paakkonen et al. (1999) has shown that IR drying improves the quality of herbs and Dontigny et al. (1992) have demonstrated that IR drying of graphite slurry significantly increases drying rate. Zbicinski et al. (1992) investigating convective air drying and IR drying have suggested intermittent irradiation drying mode coupled with convective air drying for heat-sensitive materials. Figure 47.22 shows an IR-assisted HPD system.

To dry heat-sensitive materials, a combined radiant-convective drying method or an intermittent drying mode may be applied. An IR-augmented heat pump dryer could be used for fast removal of surface moisture during the initial stages of drying, followed by intermittent drying over the rest of the drying process. This mode of operation ensures a faster initial drying rate. Therefore, an IR-assisted HPD would offer the advantages of compactness, simplicity, ease of control, and low equipment costs (Mujumdar, 2000a). Also, there are the possibilities of significant energy savings and enhanced product quality due to reduced residence time in the drying chamber. On the flip side, the high heat flux may scorch the product and cause fire and explosion hazards (Mujumdar, 2000a). Clearly, good control of IR operation is essential to achieve the desired results in terms of drying kinetics and product quality, as well as to ensure safe operation. So, a good feedback control is the one that enables the IR power source to cut off if excessively high temperatures are measured in the chamber, which may lead to overheating of the product. No work has been reported on this concept for HPD to date.

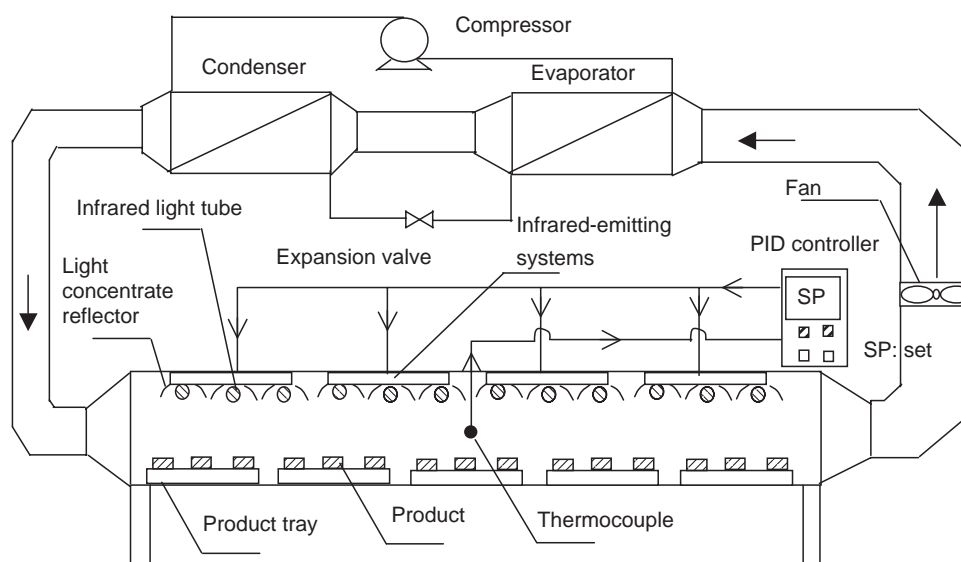


FIGURE 47.22 Schematic diagram of IR-assisted heat pump dryer.

47.11.3 RADIO FREQUENCY-ASSISTED HEAT PUMP DRYING

A limitation of heat transfer in conventional drying with hot air alone, particularly in the falling rate period, can be overcome by combining RF heating with conventional HPD (Marshall and Metaxas, 1998). RF generates heat volumetrically within the wet material by the combined mechanisms of dipole rotation and conduction effects that speedup the drying process (Metaxas and Meredith, 1983). A typical RF-assisted heat pump dryer comprises a vapor compression heat pump system retrofitted with a RF generating system capable of imparting RF energy to the drying material at various stages of the drying process. Figure 47.23 shows a schematic diagram of the RF-assisted heat pump dryer.

Materials that are difficult to dry with convection heating alone are good candidates for RF-assisted drying. Materials with poor heat transfer characteristics, e.g., ceramics and glass fibers, are traditionally the problem materials when it comes to heating and drying. RF heats all parts of the product mass simultaneously and evaporates the water *in situ* at relatively low temperatures usually not exceeding 180°F or 82°C (Thomas, 1996). Since water moves through the product in the form of a gas rather than by capillary action, migration of solids is avoided. Warping, surface discoloration, and cracking associated with conventional drying methods are also avoided (Thomas, 1996).

The following are some of the characteristics of RF-assisted HPD:

1. RF-assisted drying improves the color of the products especially those that are highly sus-

ceptible to surface color change since RF drying starts from the internal to the product surface, minimizing any surface effect.

2. Cracking, caused by the stresses of uneven shrinkage in drying, can be eliminated by RF-assisted drying. This is achieved in the dryer by even heating throughout the product maintaining moisture uniformity from the center to the surface during the drying process.

The potential for direct application of the RF-assisted HPD in the industries is appreciable for the following reasons:

1. Simultaneous external and internal drying significantly reduces the drying time to reach the desired moisture content. The potential for improving the throughput of product is good. For example, in the bakery industry, the throughput for crackers and cookies can be improved by as much as 30 and 40%, respectively (Clark, 1997).
2. By greatly reducing the moisture variation throughout the thickness of the product, differential shrinkage can be minimized. This promotes RF-assisted heat pump dryer for drying materials with high shrinkage properties.
3. Closer tolerance of the dielectric heating frequency, (1) 13.56 MHz \pm 0.05%, (2) 27.12 MHz \pm 0.60%, and (3) 40.68 MHz \pm 0.05% (Clark, 1997), significantly improves the level of control for internal drying and thus has potential in industry that produces products that require precision moisture removal.
4. The moisture-leveling phenomenon of RF drying ensures a uniform level of dryness

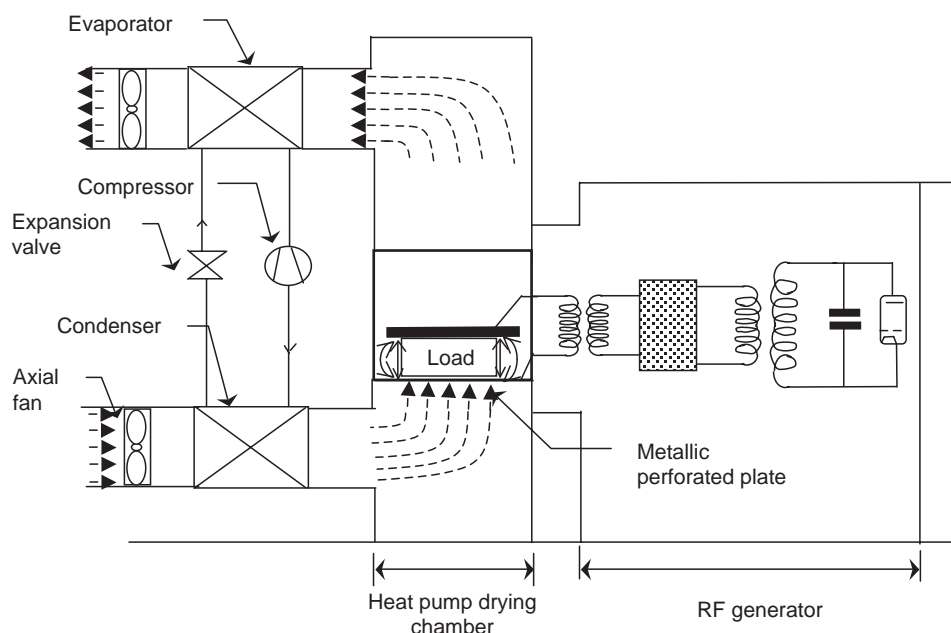


FIGURE 47.23 Schematic diagram of RF-assisted heat pump dryer.

throughout the product. Industries that have products requiring uniform drying, such as the ceramics, can consider RF drying as a good alternative.

47.11.4 SOLAR-ASSISTED HEAT PUMP DRYING WITH ENERGY STORAGE SYSTEM

In places where very rich sources of solar energy are available, the incorporation of a solar heating system to HPD may further improve the energy efficiency of the overall drying system. Such a system may also be appropriate for higher drying temperature. Instead of using the conventional heating system to provide for auxiliary heating, the storing of solar energy in a phase-change material such as paraffin wax for discharging sensible energy to the drying air leads to a cheaper means of employing higher drying temperature. Further, such a system offers the flexibility of operating with the heat pump, solar system, or with both systems complementing each other. Troger and Butler (1980) have experimentally evaluated a solar collector-cum-rockbed storage system for peanut drying. Chauhan et al. (1996) have studied the drying characteristics of coriander in a stationary 0.5 t/batch capacity deep-bed dryer coupled to a solar air heater and a rockbed storage unit to receive hot air during off sunshine hours. They found that to reduce the average moisture of coriander grains from 28.2 (dry basis) to 11.4% (dry basis) requires 27 cumulative sunshine hours. Using the stored heat from the rockbed energy storage system, the removal of

the same moisture can be accomplished with just 18 cumulative sunshine hours.

The solar energy supply system proposed in this section consists of solar collectors, blowers, phase-change storage tank, air-valves, and pipes as shown in Figure 47.24. Depending on the type of drying material which determines the air temperature, the air may be flown with open full partial discharge circulation or full discharge circulation mode.

The advantages of a solar-assisted HPD can be summarized as follows:

1. Easy conversion of natural energy for storage resulting in significant saving of energy
2. Environment-friendly process
3. Easy to implement control strategy
4. Higher operating drying temperature

However, the disadvantages of such system can also be summarized as follows:

1. Higher capital costs are incurred for additional solar panels, blowers, storage tank, and valves.
2. Implementation of the system is practical and cost effective provided the average annual sunshine time (approximately greater than 2600 h [Zhang et al., 2000]) is high and the annual total quantity of radiation is sufficient (more than 6×10^6 KJ/m²).
3. The amount of stored solar energy is greatly subjected to the weather conditions.

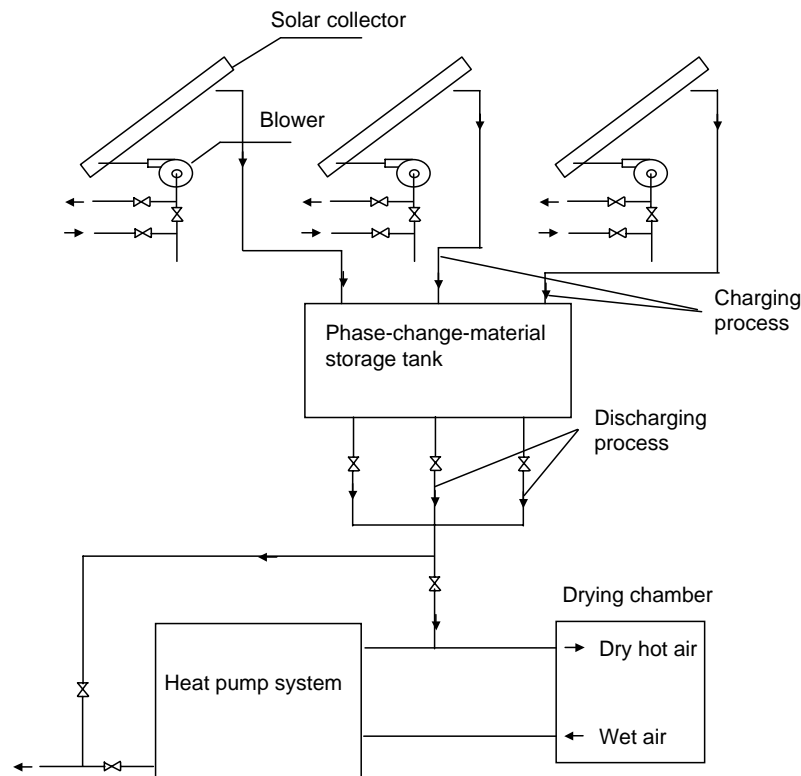


FIGURE 47.24 A solar-assisted heat pump system incorporating energy storage system.

Table 47.5 provides the energy cost figures of three methods for drying 1 m³ lumber. It is apparent that steam drying consumes more energy and has the highest cost. Even though the solar-assisted HPD consumes more energy than the pure solar system the energy cost is only slightly higher.

47.11.5 MASS TRANSFER MODE—VACUUM/ATMOSPHERIC PRESSURE

A very useful way to enhance the quality of heat-sensitive products and yet achieve a fairly dried

product is through the use of a pressure-regulatory system. The operating pressure range is usually from vacuum to below atmosphere. A total vacuum system may be very costly to build because of the need for stronger materials and better leakage-preventing facility. Therefore, the system that is proposed here is recommended to operate above vacuum condition. The period of operating at lower pressure may be continuous at a fixed level, intermittent, or a cyclic prescribed pattern. The suitability of employing the appropriate type of pressure-swing pattern depends chiefly on the drying kinetics of the product and its thermal properties.

TABLE 47.5
The Consumed Energy and Cost of Three Methods for Drying 1 m³ Lumber

Drying Method	Wood Species	Plate Thickness of Lumber (cm)	Initial Moisture Content (%)	Final Moisture Content (%)	Energy Consumption (W/m ⁻³)	Energy Cost (\$/m ⁻³)
Solar	<i>P. Koraensis</i>	5.0	31	14	10.5	31.5
Solar-heat pump dehumidifier	<i>P. Strobus</i>	5.0	38	15	34	59.4
Steam	<i>P. Koraensis</i>	5.0	40	12	65.5	85

Source: Adapted from Zhang et al., 2000.

Most heat-sensitive materials often undergo a freeze-drying process to minimize any quality degradation that may arise due to temperature effects. Generally, freeze drying yields the highest quality product of any dehydration technique (Mujumdar, 2000b). However, the cost of freeze drying has been found to be at least one order-of-magnitude higher than other drying system such as spray dryer.

According to Nijhuis et al. (1996), freeze drying (known as a suitable dehydration process for pharmaceutical and food products) is not suitable for the production of homogeneous films because the films obtained are generally very spongy. Also, a freeze-dried product tends to be porous and the problem of rapid rehydration may arise once the product is exposed to a more humid environment. Moreover freeze drying is very energy intensive. The equipment is also more expensive than atmospheric pressure dryers. It is best suited for heat-sensitive materials, or when solvent recovery is required, or if there are risks of fire and/or explosion.

Maache-Rezzoug et al. (2001) have recommended a pressure-swing drying mechanism to produce homogeneous thin sheets. The experiments they conducted recently to dry a collagen gel in order to obtain a homogeneous film were carried out using a new process: dehydration by successive decompression. Their process involves a series of cycles during which the collagen gel is placed in desiccated air at a given pressure and then subjected to an instantaneous (200 ms) pressure drop to a vacuum (7 to 90 kPa). This procedure is repeated until the desired moisture is obtained. A comparative study between this new drying process and conventional methods indicated that the drying time was reduced from 480 and 700 for vacuum and hot air drying, respectively to 270 min

for the process ($P_h = 600$ kPa; $t_h = 18$ s; $p_v = 5$ kPa, and $t_v = 5$ s).

Integrating such a pressure-swing system to a heat pump dryer would significantly improve product quality, by the use of lower drying temperature, and at the same time reduce the drying time that would result in a smaller drying chamber to obtain similar product throughput.

47.12 AIRFLOW DISTRIBUTION IN DRYING CHAMBER

The distribution of the drying air within the drying chamber influences the heat and mass transfer process between the air and the drying product. It affects the drying kinetics of the products placed at different locations in the drying chamber. When the drying air absorbs moisture from the product, it becomes gradually saturated. Its potential to remove moisture from the product placed downstream in the direction of the airflow diminishes. If the distribution of the drying air in the drying chamber is not properly considered then nonuniform drying occurs within the chamber. Sections of the product that first contact the conditioned air would be drier than other sections that interact with a more saturated air. Therefore, it is essential that the drying air be uniformly distributed within the drying chamber to minimize the problem of “uneven” drying within the drying chamber.

Floor racks, sidewall flues, and special ceiling ducts are common items used to facilitate air circulation. Figure 47.25 shows the example of a cross-flow air circulation in the drying chamber. The air is first distributed to a ceiling duct, which discharges the conditioned air perpendicularly to the product trays.

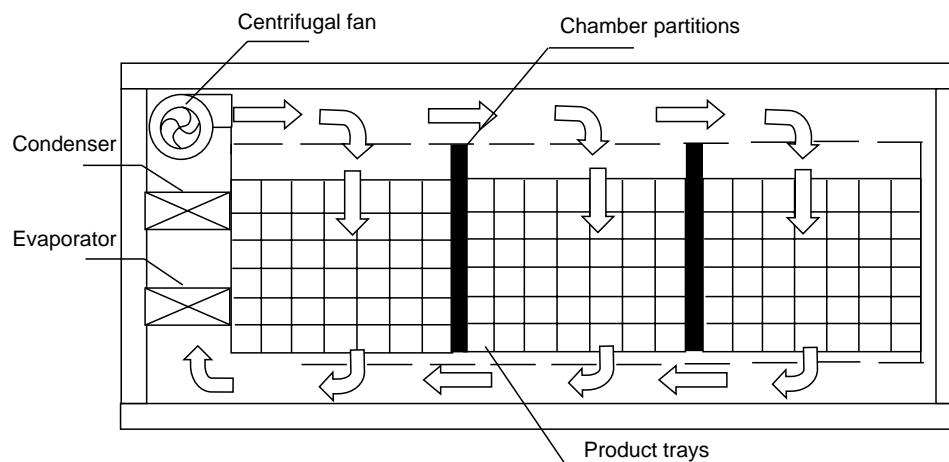


FIGURE 47.25 Air distribution within the drying chamber.

The ceiling duct, wall ducts, and a special floor duct together form a completely closed envelope around the drying chamber. The fan draws air through the evaporator and condenser and discharges into a ceiling duct above the trays. From here, it is distributed to the product on either sides of the wall. It then goes into the flow duct, which carries the air back to the heat exchangers. When the wall dividers are in place, dividing the chamber into three sections, the air circulation pattern is essentially the same because airflow is from ceiling to floor and the wall dividers do not interfere with the circulation pattern. Further, heat and mass transfer is enhanced for a cross-flow pattern when compared with parallel flow. Therefore, less airflow may be required to achieve the amount of moisture removal resulting in the saving of operating fan cost.

47.13 REFRIGERANTS

The heat pump dryer depends on the properties of CFCs and hydrochlorofluorocarbons (HCFCs) to enable: (1) heat recovery; (2) sensible heating and cooling; and (3) dehumidification. CFCs and HCFCs are widely used throughout the food industry as described in Table 47.6. The principal CFC refrigerants in current use are R502 and R12, while the principal HCFC in use is R22. It is estimated that the total usage of these refrigerants in the food sector in Canada is about 1800 t/year (TCEA, 1995).

Issues of the impact of emissions on the Earth's stratosphere and its protective ozone layer has led to the imminent schedule of CFC and HCFC phase-out. The phaseout of CFCs (and HCFCs in the long-term)

will force the food industry to evaluate their refrigeration systems for refrigerant replacement. The most important challenge is the need to develop refrigerant management plans to convert CFCs to new ozone-friendly refrigerant alternatives. On the flip side, this challenge presents an opportunity for the food industry to examine the energy efficiency of their refrigeration equipment and to introduce new energy savings methods at the time of CFC (and HCFC) containment, conversion, or replacement. There are a large number of alternative refrigerants on the market. Table 47.7 lists some of these alternatives.

The present direction in refrigerant management is to look at naturally occurring fluids as potential replacements. Hydrocarbons and naturally existing fluids, e.g., propane and ammonia, are among future environment-friendly refrigerants with zero global warming and ozone depletion potentials. The heat pump dryer with hydrocarbon and natural working fluids can save significant amounts of energy while playing the role of an environment-friendly refrigerant (Strømme et al., 1999). Strømme et al. (1999) have also compared the performance of several refrigerants, e.g., R717, R290, R22, and R134a, in their heat pump dryers. They found that for lower air temperature, propane has about 3% higher SMER than ammonia. However, above 10 to 30°C ammonia has higher SMER than R290, R22, and R134a. Ammonia was found to be the most favorable refrigerant for heat pump dryers in the temperature range of 30 to 80°C.

Other natural working fluids, e.g., steam and air, have been proposed for heat pumps. Steam and air are readily available, cheap and environment-friendly. Steam was proposed as a heat pump working fluid for paper impingement drying (Nassikas

TABLE 47.6
Application of CFC/HCFC in the Food Industry

Application in Food Industry	CFC/HCFC
Drying	R12 and R22
Transport refrigeration	R502
Retail unitary refrigeration—display/storage	R12, R502, and R22
Retail central refrigeration—display/storage	R12, R502, and R22
Cold storage	R502
Refrigerated storage	R12 and R22
Refrigerated vending machines	R12 and R22

Source: TCEA, Capitalising on the Energy Saving Opportunities Presented by CFC and HCFC Phaseout in Nondomestic Refrigeration, E.A. Technology for the Canadian Electrical Association, Montreal, 1995.

TABLE 47.7
Some Alternative Refrigerants

Alternative Refrigerants	Molecular Weight	Critical Pressure (bar)	Critical Temperature (°C)
R12 Alternatives			
R134A	102.0	40.7	101.1
R409A	97.5	46.0	107.0
R22	86.48	49.9	96.1
R502 Alternatives			
R404A	97.6	36.9	72.4
R507	98.9	37.9	76.1
Natural refrigerants			
Ammonia (R717)	17.03	111.5	132.4
Propane (R290)	44.10	42.4	96.8

et al., 1992). An air cycle heat pump was proposed by McKay (1993). He showed that the condenser temperature was as high as 206°C and the evaporator temperature was as low as -40°C. This wide range of operating temperature makes the air cycle heat pump versatile for freeze drying, hot air drying, and process heating (Prasertsan and Saen-saby, 1998b).

47.14 VERSATILITY OF HEAT PUMP DRYING SYSTEM

Commercially available heat pump dryers are designed with both the refrigerating components and the drying chamber integrated as one complete drying unit. For such dryers, the required refrigeration capacity of the heat pump to cover the heat duties, both latent and sensible, is sized according to the size of the drying chamber and the maximum product loading. These dryers are inflexible for further scale-up of the refrigerating equipment to deal with increase in product drying capacity. This results in the partial or complete replacement of the refrigerating components. Furthermore, in the event of a component breakdown, the drying operation has to be terminated, resulting in downtime of the dryer with significant loss in the production capacity.

The design of a versatile heat pump system is proposed here whereby the refrigerating equipments are assembled exclusively from the air-handling chamber before they are interfaced through the use of industrial air couplers. These couplers allow the air to be received for conditioning and discharge the

conditioned air to the chamber for drying. The conditioning of the air refers to the air temperature, velocity, and absolute humidity.

The principal advantage of such a versatile heat pump system is the flexibility to allow the heat pump components, assembled as a modular unit with portable or transportable features, to be coupled and decoupled to any portable or transportable chamber for drying and general air-conditioning applications. Also, this heat pump design enables easy scale-up of dryers with varying chamber size and configuration to meet higher production demands. Lastly, easy maintenance and repair work can be carried out on the heat pump components with little or no production downtime in the drying process, which minimizes product spoilage in the storage chambers.

The present design comprises:

1. A heat pump system assembled in a modular unit comprising two internal evaporators, one external evaporator, one internal condenser, one air-cooled condenser, two expansion valves, one backpressure-regulating valve, one external evaporator, one compressor, and one circulating fan
2. A chamber incorporating two or more ports to enable air communication between heat pump and chamber
3. Two or more industrial air couplers to link the heat pump system to the chamber

Figure 47.26 shows the proposed design of the heat pump field dryer. This design is unique because it allows further scale-up of the drying chamber

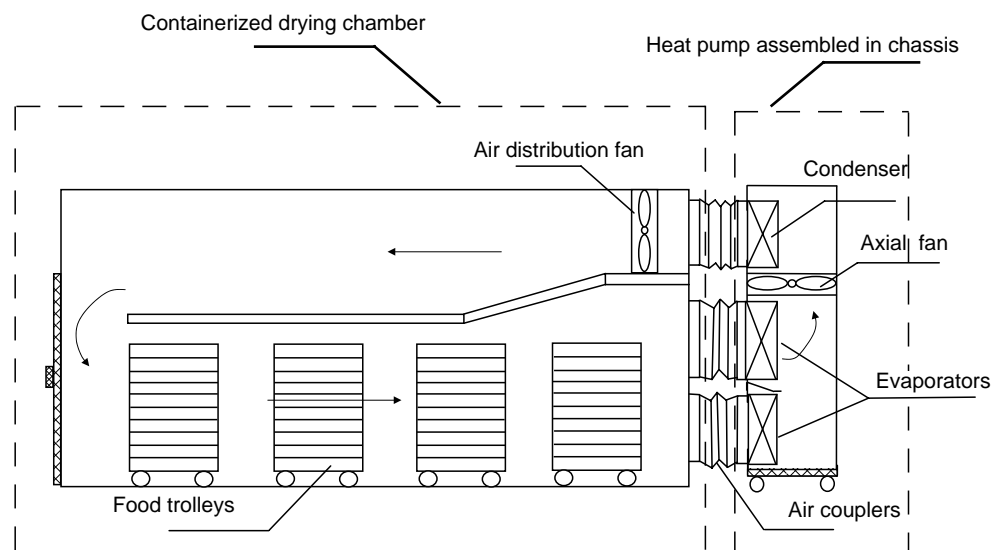


FIGURE 47.26 The proposed design of a versatile heat pump field dryer.

without changing the position of the coils and other refrigerating components. For reasons pertaining to simplicity and cost, a conventional transport container can be used as the chamber. The drying chamber is partitioned into two segments by three partitions. These partitions create a U-channel duct for the drying air to flow from the ports to the loading trolleys with product trays before the drying air is returned to the heat pump by the return ports. A recirculation fan is installed to provide additional airflow to the products. The products are first laid in the product trays with fixed wheels before they are pushed through the loading door. The product, after being dried to the desired moisture content, leaves the drying chamber by the unloading section.

It should be mentioned that the application of this modular design is not exhaustive, it extends beyond drying and cold storage applications. The concept of a coupled and decoupled process, for easy interface between equipment and air-handling unit, can be used in many air-conditioning applications through appropriate selection of expansion valve, compressor, condensers, evaporators, and refrigerant.

47.15 ECONOMICS OF HEAT PUMP-ASSISTED DRYING SYSTEM

Potential investors in drying technology need to have information pertaining to the capital investment, total operating cost, and breakeven period. Equally important is the knowledge of the various drying parameters that can affect the drying process. Such information can help them to reduce the operating cost required to remove per unit of moisture from the product. The following section examines the techno-economics of the heat pump dryer.

The total cost of a HPD system is made up of two kinds of basic costs:

1. Fixed cost elements. These are items unrelated to the amount of moisture removed from the product over the years. The dryer cost is the primary element because once the dryer is purchased, interest has to be repaid whether or not the dryer is actually in use. This item also includes the maximum demand charges for the electricity supply.
2. Variable costs. These are costs that progressively increase as the dryer operates. Its main element is the energy used but it could also include the dryer maintenance element. The actual cost is dependent on the product of the energy cost and the effectiveness of the dryer.

Therefore, the equation for total cost is

$$\text{Total cost of removing one liter of water} = \frac{\text{total variable cost} + \text{total fixed cost}}{\text{total water removed}}$$

where

total variable cost = dryer power (kW) × running time (hours) × energy cost (\$/kWh)

and total fixed cost = interest or amortization of the capital cost.

The total cost of removing 1 L of water from the product can be portrayed in graphical form as shown in Figure 47.27. It can be readily observed that if the dryer is operating for only a short period in a year, say below 2000 h, then the total cost of extracting a liter of water from the product is critical. However, if the operating hours are longer at 8000 h/y, the total cost of removing a liter of water from the product is significantly lower. This trend is typical of most heat recovery systems because greater amount of heat is recovered for longer operating periods. In the case of the dryer, the direct translation is then an offset of the operating cost resulting in lower cost for each unit of water removed.

Several factors are expected to influence the overall economic viability of a heat pump dryer. Some of these are process or design parameters while others are economic parameters. Zylla et al. (1982) in evaluating the potential for heat pumps in drying

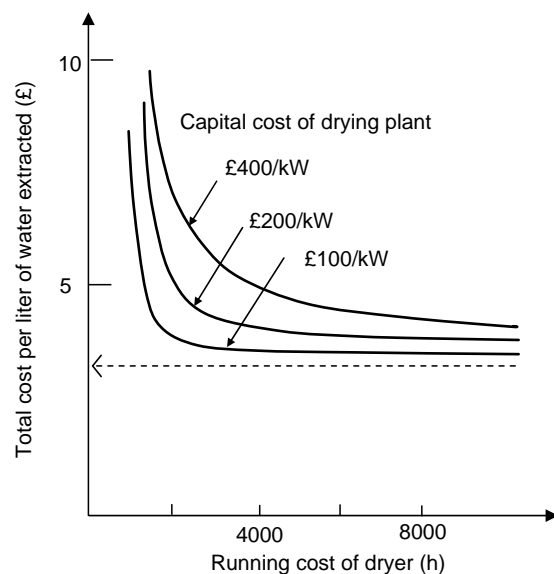


FIGURE 47.27 Total cost of drying. Assumption: effectiveness = 1 l water/kWh. (Adapted from Brundrett, G. W., *Handbook of Dehumidification Technology*, Butterworths, London, 1987.)

systems found that heat pump dryers can reduce the operating cost as much as 70% in comparison to a recuperative heating dryer. Pendyala et al. (1986) have found that for typical drying conditions with air temperature and relative humidity in the respective range of 25 to 65°C and 40 to 100%, a heat pump dryer designed for a drying capacity of 200 kg/h has a payback period of 2 to 3 y. Judging from the humidity of the exhaust air, it was further observed that payback period can be shortened from 3.2 to 2.0 y if relative humidity of exhaust air increases from 0.4 to 0.7. It can then be deduced that the payback period for initial investment is generally reduced if more product moisture is available for heat recovery. Also, the payback period is sensitive to the operating pressure of the evaporator and condenser. The minimum effectiveness factor should be more than 0.55 if the payback period is less than 3 y.

47.16 FUTURE TRENDS IN HEAT PUMP DRYING—MULTIPLE DRYERS

Although not reported heretofore, it is possible to design a HPD system that uses a single low capacity HP to supply drying air to several different chambers according to a preprogrammed schedule. This is feasible because many food products have long falling rate periods. There are several advantages of operating the heat pump with multiple drying chambers. They are:

- Improved quality of products such as surface color and reduced case hardening
- Improved energy efficiency with more latent load for heat recovery
- Reduced capital cost and floor space requirement
- Easy temperature schedule control for different products in different drying chambers

When only a marginal amount of convection air is needed to evaporate moisture, the drying chambers are operated in sequence. The air from the heat pump can be directed sequentially to two or more chambers or can be divided according to a preset schedule to two or more drying chambers, which may dry the same or different products. Thus, the heat pump can be operated at near optimal level at all times. Even if the drying times for each chamber may increase due to the intermittent heat input, the overall economics should improve considerably. A smaller HP can double or triple the drying capacity, especially with the help of supplementary heating by IR, MW, or RF.

Figure 47.28 shows a schematic diagram of a multiple-chamber HPD process. When the drying rate of the product approaches the second falling rate, the drying air is channeled to the second chamber to dry the freshly changed product of higher moisture content. Auxiliary heating may then be used to provide the thermal requirement for drying. Figure 47.29 shows the scheduling sequence. Such a scheduling enables reduction in product temperature, resulting

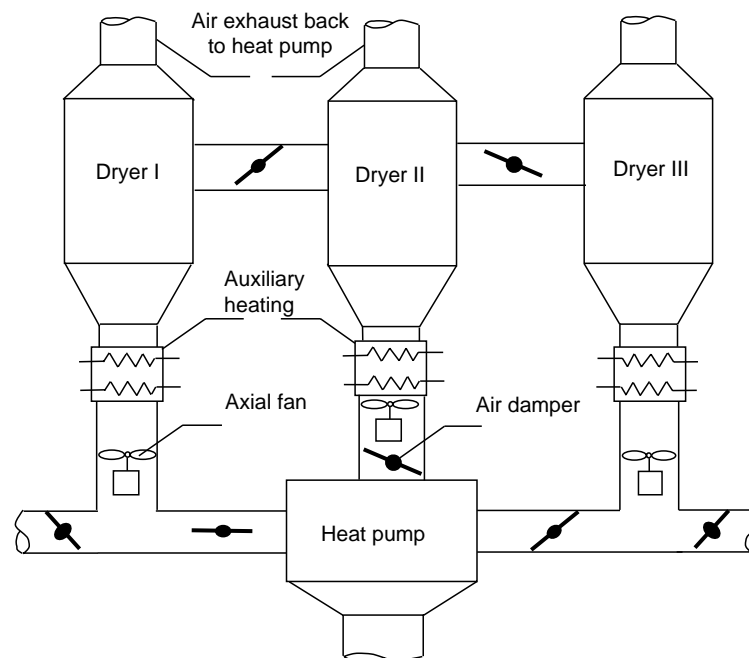


FIGURE 47.28 Schematic layout of multiple-chambers heat pump drying process.

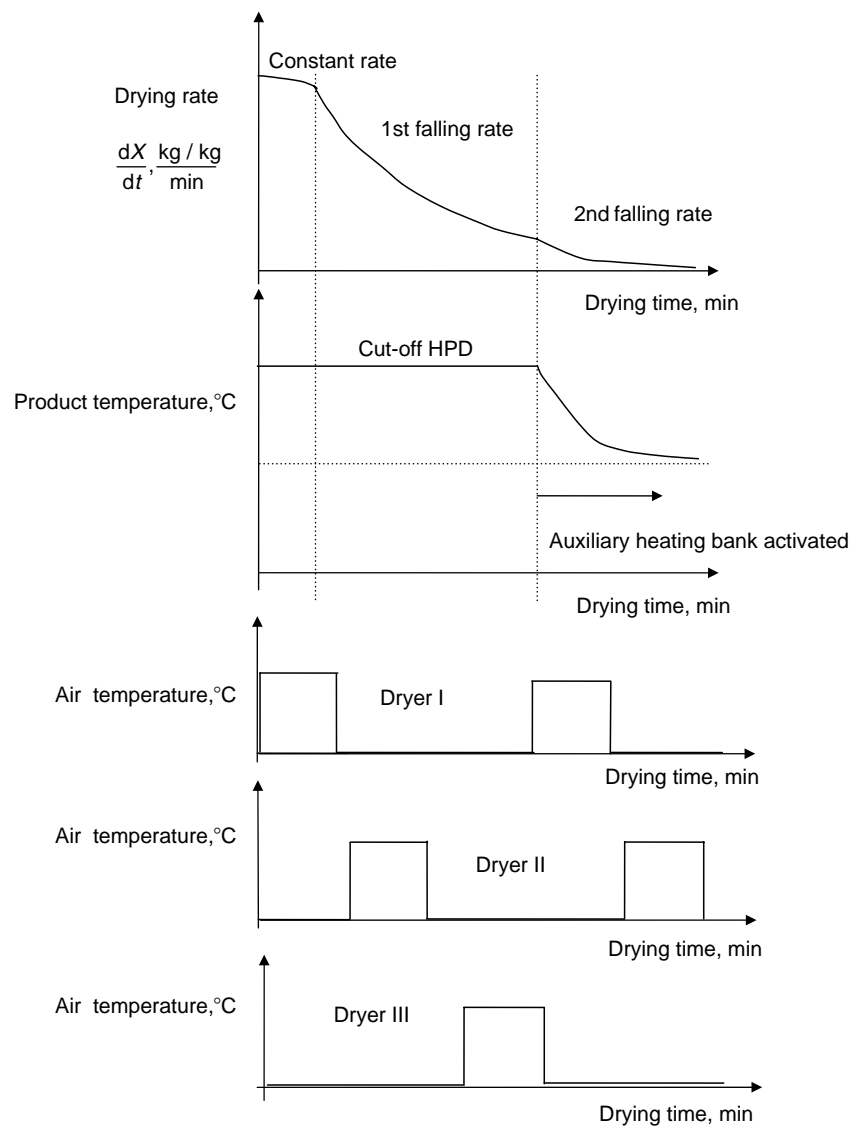


FIGURE 47.29 Sequence for multiple heat pump batch drying.

in improved quality and reduces case hardening. Improved energy efficiency of the heat pump system is also expected because higher latent loading in the second chamber in the form of product moisture is available for heat recovery.

From an economic perspective, the most attractive aspect of multiple-chambers HPD is the reduction in capital cost, because one heat pump system is capable of accomplishing the drying task of two or more separate HPD units. Further, a control strategy can be easily implemented through control of air dampers.

47.17 CONCLUSION

HPD has evolved to become a mature technology over the last two decades. However, it is not applied

as widely as it should or could be. Initial costs as well as operating costs remain a problem, since few major vendors of drying systems offer HPD systems. Efficient use of energy in such energy-intensive operations as drying is crucial to the reduction of net energy consumption and hence emissions of greenhouse gas. With the eventual acceptance of a carbon/energy tax around the world energy, conservation will become a key concern in many industrial operations. Heat pumps appear to have attracted interest as a means of energy recovery since this idea was first proposed by Lord Kelvin in 1852. Commercial heat pumps based on the vapor compression cycle or the absorption cycle are operational in numerous applications. New HP technologies such as the adsorption cycle or the chemical reaction cycle are emerging rapidly, although they have yet to find major

industrial applications. In future, the need to develop cost-effective heat pumps using environment-friendly refrigerants (e.g., CO₂ and NH₃) to replace CFCs will also impact HPD systems. With more versatile designs of heat pump dryers in the pipeline, their applications can be extended to include other air-conditioning applications, resulting in effective investment of capital costs. While enough is known about heat pumps and various dryers, optimal integration of the technologies remains a challenging R&D task. It is hoped that more attention will be paid to this development in the coming decade.

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NOMENCLATURE

(COP) _A	actual coefficient of performance
h	enthalpy of drying air, J/kg dry air
h_{fg}	heat of evaporation, J/kg K
HGC	hot gas condenser
HP	heat pump
HPD	heat pump drying
HPE	high pressure evaporator
LP	low pressure evaporator
M_a	mass flow rate of drying air, kg/s
PID	Proportional–Integral–Differential
Q	power, KW
Q_e	energy, kJ
SC1	subcooler 1
SC2	subcooler 2
SEC	specific energy consumption, kWh/kg moisture
SMER	specific moisture extraction rate, kg moisture/kWh
T	temperature, °C

GREEK SYMBOL

α	intermittency cycle ratio
ω	humidity ratio of air, kg water/kg dry air
τ	profile periodic time, min

SUBSCRIPTS

100%	saturation level
a	actual

c	condenser
ev	evaporator
off	nonheating period
on	active heating period

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48 Safety Aspects of Industrial Dryers

Adam S. Markowski and Arun S. Mujumdar

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48.1 INTRODUCTION

The statistics of industrial accidents show that drying should be regarded as a potentially hazardous operation that has brought a number of reported incidents with serious results for personnel and equipment [1,2]. The data indicate that the accident rate per 10^5 work-

ers at risk is considerably greater in the food industry than, e.g., in the chemical industry. Approximately 8 to 9% of all dust explosions in the food industry is related to the drying operation (Figure 48.1). The other data for the period 1967–1983 in the German sugar industry indicate that drying contributes to 37% of all accidents [3], whereas in the French milk

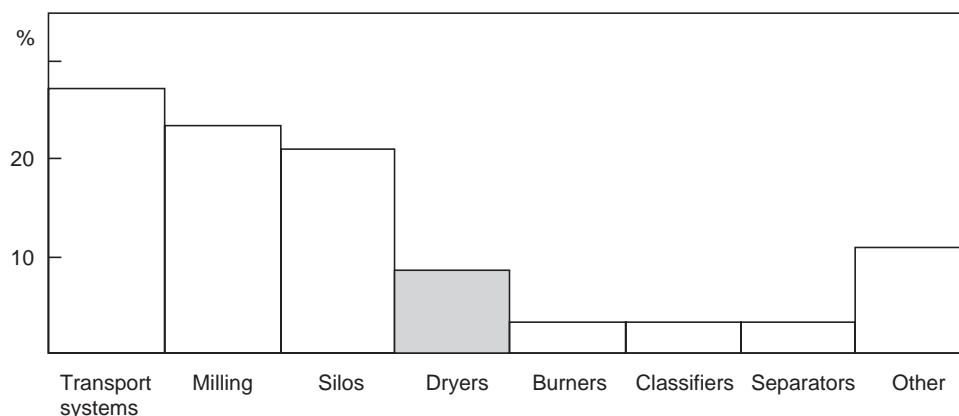


FIGURE 48.1 Distribution of explosion location for dust explosion in the food industry. (From W. Hamm, *The Chemical Engineer*, 8-9: 22-25 (1984). With permission.)

industry an average of four major accidents in spray dryers were reported annually [4]. Based on 89 accidents that happened in 1965–2000, 415 people were injured and 16 fatalities were reported in *The Accident Database* [5]. It is worth to note that in most cases of spray dryer accident in the food industry fire was observed whereas an explosion experienced in <10% [6]. The above reports underline the importance of safety from fire and explosion hazards in dryers and in the ancillary equipment.

Drying is particularly prone to fire and explosion hazards because it is a process in which external heat sources are applied.

Most of the materials, whether natural products or synthetic organic products, are combustible or sensitive to heat as well as are exposed to elevated temperature in the air.

All possible ignition sources may be present in dryers.

There are three factors that influence the potential for fire and explosion: (1) process; (2) engineering; and (3) management.

The process factor is associated with the characterization of material being dried and the physical conditions to which it is subjected. This is a very important consideration because it enables for the recognition and assessment of fire and explosion hazards. For this factor, the basis of safety is the prevention methods (e.g., drying in an inert atmosphere, elimination of the formation of explosive mixtures, and rigorous exclusion of all possible ignition sources).

The engineering factor relates to the plant layout, its location, the equipment used, and its engineering standards. Special attention should be paid to the design stage at which safety and reliability rely upon

the application of various codes of practice and standards. This is a basis for prevention of the fire explosions hazards. Within this factor, the philosophy of safety is based on the acceptance of the possibility of fire and explosion and provision of a method for protecting personnel and equipment from its consequences (protection method).

Usually, each drying plant being part of a chemical plant is under last protection layer that is emergency response that may essentially mitigate the impact of a fire or explosion to human, property, and environment.

Within the management factor, which usually accounts for majority of industrial accidents, the most important tool is a risk assessment, which examines the hazards associated with the drying operation and on that basis safety prevention and protection measures are designed. Besides these all operating, house-keeping, and maintenance procedures as well as training systems are defined by the management based on the risk assessment.

In view of the diversity of the nature of material handled and types of dryers, it is not possible to present any universal answer to the above-mentioned factors. However, many of the general values regarding the process, engineering, and management factors are commonly applicable to all types of dryers. These are given here, but one must remember that each material and dryer should be assessed individually, taking all steps possible to provide adequate safety. It is worth emphasizing that the application of different precaution methods in drying processes is a statutory requirement in many countries; e.g., the Factories Act of 1961 (in the United Kingdom) requires all practicable precautions to be taken to prevent an explosion wherever an explosion dust is involved.

Similar requirements are in Germany (Störfall VwV), the United States, and the other European Union countries. The EU ATEX Directives 94/9/EC is fully applicable to dryers [7]. An excellent guide to safety in drying operations has been prepared and published in a revised version by the Institution of Chemical Engineers of Great Britain [8].

48.2 THE PROCESS FACTOR

48.2.1 NATURE OF FIRE AND EXPLOSION HAZARDS IN DRYERS

The simultaneous presence of three basic conditions is required for fire and explosion to occur (Figure 48.2)

1. Fuel (e.g., flammable dusts, solvents, or hybrid mixtures)
2. Oxygen (air)
3. Ignition source with sufficient energy

When any of these three basic conditions is missing, no explosion can occur.

The majority of powdery, fibrous, and granular products processed in dryers contain fuels. Moreover, fine combustible dust, when dispersed in air, can form a reactive mixture leading to explosion, which may trigger others as dust is dislodged by a primary event. Frequently, the material can accumulate within a dryer in the form of deposits (in places unswept by a drying agent) or residues because the dryer may not have been cleaned before start-up. As a result of the accumulation, the overheating process may take place. The risk is substantially increased if a flammable solvent is being evaporated from the material or a mixture of combustible dust and flammable gas or vapor occurs (e.g., a hybrid mixture).

The air generally used as a drying agent contains enough oxygen for combustion and also all types of ignition sources may be present during drying. Moreover, drying is a very complex unit operation that may include other operations like pneumatic conveying, separation, storage, and direct/indirect heating

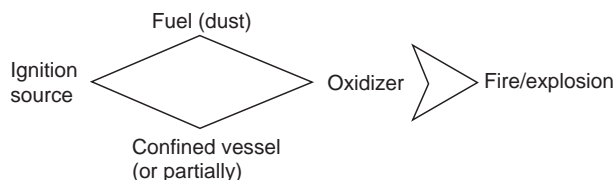


FIGURE 48.2 Fire and/or dust explosion components.

systems, which by the nature of the process may bring about a fire and explosion hazard. As a consequence, the following hazards may appear:

1. Fire (i.e., combustion of dust, which means a chemical reaction of a combustible and gaseous oxidant with resultant production of flame and burning gases)
2. A dust explosion (i.e., flame propagation through a confined dispersed dust with pressure generation by combustion)
3. A deflagrating or detonating explosion
4. An uncontrolled thermal decomposition of material producing extensive damage

Based on the nature of the reaction exhibited by materials when heated, they may be classified as follows:

1. Deflagrating or detonating explosives
2. Materials that exhibit violent exothermic reaction even in the absence of air (e.g., certain peroxides)
3. Materials that undergo exothermic oxidation or decomposition in layer, bulk, or fluidized form, or that can disperse into flammable clouds in air

The differentiation between a detonation type of combustion and deflagration is based on the velocity of combustion zone. According to NFPA Standard 69 (1986) edition the deflagration is defined as a propagation of combustion zone at a velocity that is less than the speed of sound in the unreacted medium. The opposite feature is characteristic for the detonation type of the velocity combustion.

Types 1 and 2 require a specialized drying technique and are not discussed in this chapter. The drying of solids of type 3 can be safely carried out by proper control of operating conditions and by providing adequate precautionary measures.

The hazards with combustible materials can be divided into:

- Ignition of dust clouds
- Ignition of hybrid mixture
- Ignition of dust layer or deposits
- Ignition of bulk powder

Ignitions of dust clouds are due to oxidation reactions whereas ignitions of dust layer or bulk powder are due to self-ignition or exothermic decomposition.

The optimum condition for ignition and subsequently for the combustion process is when the particle separation allows sufficient air to get to each particle, but the particles are close enough for the

heat release to support burning of adjacent particles. Such conditions are met mainly in a dispersal dryer in which dust clouds are formed. As a result of a dust cloud ignition, fire dispersion or explosion in a confined place or flash fire in open space may take place. The ignition of a dust layer and bulk powder results in burning by smoldering or flame. Transition from fire to explosion and vice versa can readily occur. An overall assessment of fire and explosion hazards in dryers is presented in Figure 48.3.

48.2.2 MATERIAL CHARACTERIZATION PROCEDURE

An extensive series of laboratory tests have been conducted in various countries to enable the investigation of fire and explosion characteristics of materials to be made. The testing apparatus and procedures differ from one country to the other, although there are some common points. The tests used to be made on a laboratory scale and the results obtained were extrapolated to full scale, always

giving some uncertainty. Also, the test procedures may not reflect exactly the condition in a particular type of dryer. Therefore, it is recommended that published data may be used carefully and individual tests on the particular dust concerned may be required.

The most comprehensive ranges of tests, which are already standard in their countries, have been made in the United Kingdom, the United States, and West European countries [9–14]. In addition to these testing methods, there are special methods that have been developed by specific industries for certain processes and products. As far as drying operations are concerned, ICI Ltd. (United Kingdom) [15] and CIBA-Geigy (United Kingdom) [8] have set up such procedures.

As a result of these established procedures, data regarding the explosion characteristics of many products already exist [14–16,19–21]. With the product not tested previously, the following characterization procedure is recommended (Table 48.1).

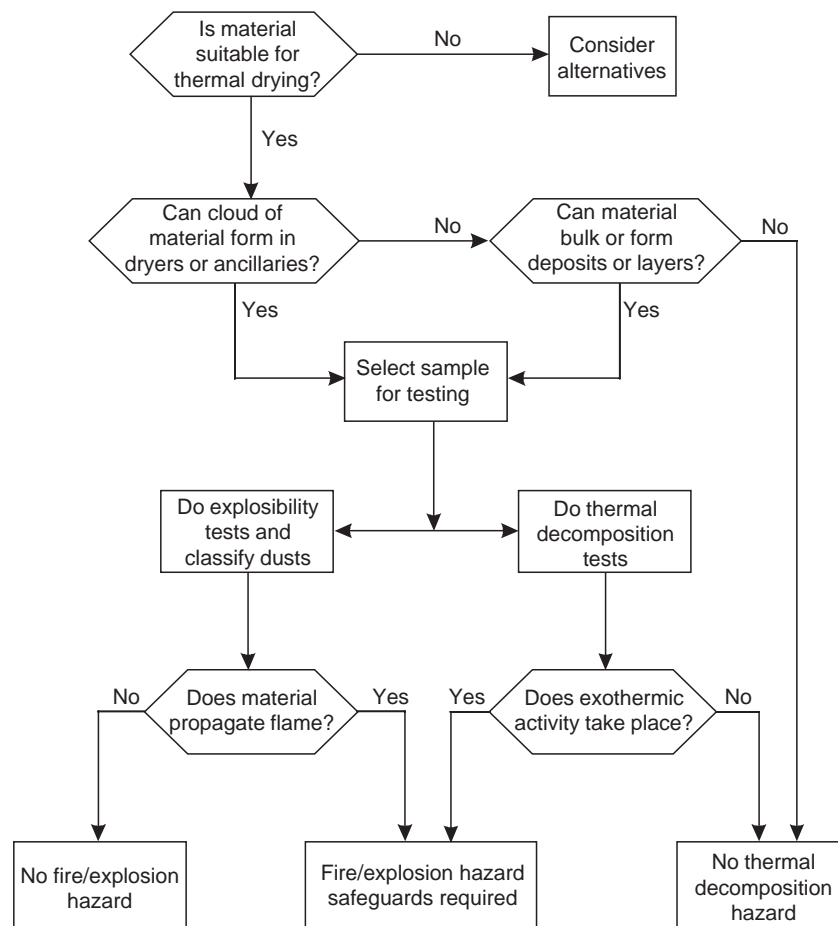
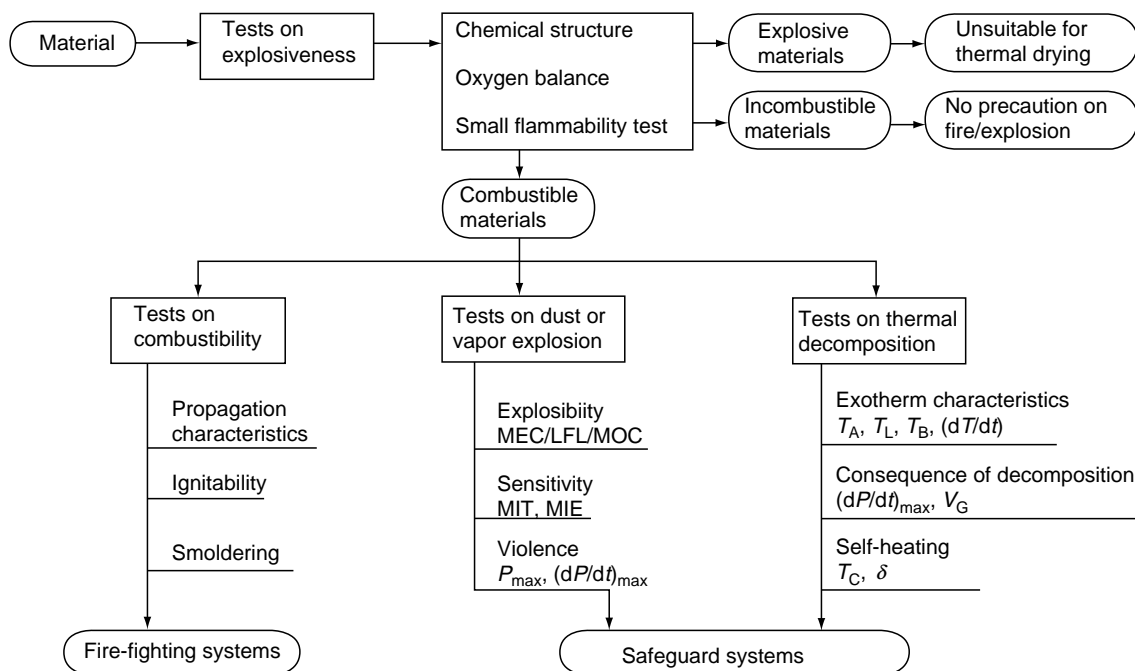


FIGURE 48.3 Overall assessment of fire and explosion hazards in dryers.

TABLE 48.1
Material Characterization Procedure



48.2.2.1 Explosiveness

In order to identify materials that may be accepted for drying, it should be determined whether or not they have properties associated with deflagrating or detonating explosives. Such an initial examination is based on the laboratory tests concerning the following:

1. Chemical composition
2. Oxygen balance
3. Small-scale flammability tests

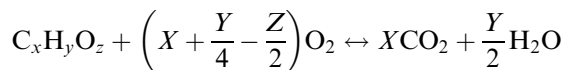
Chemical Composition—The explosibility of a material can be attributed mostly to the presence of some functional groups in its molecular structure. Some of these functional groups are:

1. Those affecting the explosive nature directly, e.g., aliphatic and aromatic nitro groups, nitrate ester, and nitramine
2. Those indirectly concerned with explosibility, e.g., azide, azo, nitroso, acetylene, peroxide, and perchlorate
3. Those that make some contribution to explosibility, e.g., hydroxyl, ether, amino, carbonyl, and sulfonic acid

It should be noted that this list is not exhaustive.

Oxygen Balance—An explosive material undergoes oxidative decomposition in the presence of oxygen under certain conditions of temperature, pressure, and composition, thereby releasing a lot of energy and a large volume of simple oxidation products. The oxygen availability within the compound is often a fairly good estimate of its explosive nature. Oxygen balance, computed from oxygen availability, when compared against known explosives, provides the indication of explosibility.

For a compound with carbon, hydrogen, and oxygen in the ratio given by $C_xH_yO_z$ and stoichiometry of combustion given by



the oxygen balance is defined as

$$\text{Oxygen Balance} = \frac{-16 \times (2X + Y/2 - Z) \times 100}{\text{molecular weight}}$$

Computed values for some of the known explosives are nitrobenzene, -162.6 ; trinitrotoluene, -74.0 ; and glyceryl trinitrate, $+3.5$.

If the oxygen balance of the compound is >-200 , the compound should be considered as being potentially explosive and these materials cannot be thermally dried.

Small-Scale Flammability Tests—The final examination is made by exposing a small-scale sample (1 to 2 g) to flame. Violent reaction will confirm the necessity of further explosiveness testing. A full account of the above-mentioned tests is available [8].

Additional information on explosibility of a material is provided by thermochemistry. All materials with endothermic heat of formation can release this stored energy upon decomposition and must therefore be regarded as hazardous substance [20]. A list of reactive chemicals and their associated hazards is given by Bretherick [21].

48.2.2.2 Combustibility

Combustibility properties are necessary to assess the type and magnitude of any fire associated with a material. These properties can be markedly affected by physical properties of the material and external factors. No universally accepted testing method is available. The basic data on combustibility may be obtained by a burning test and combustion test [8,9]. The essential information is required on the rate and type of exhibited propagation (flame or smoldering) and ignition temperature observed under one set of conditions may be changed substantially by a change of condition.

Based on the procedure developed by CIBA-Geigy [8], the material is rated from 1 to 6 depending upon its reaction to an electrically heated platinum wire. The sum of the two ratings is then taken as the overall index of the drying hazard and subsequently a hazard classification is determined (Table 48.2). Products with a hazard class Tr 3 should not be dried in heated vessels since safe handling of the dry product cannot be assumed.

TABLE 48.2
Classification of Drying Hazards by
CIBA-Geigy Ltd.

Total Rating	Hazard Class	Remarks
2–4	Tr 0	No fire hazard
5–7	Tr 1	Slight fire hazard
8–10	Tr 2	Fire hazard
11–12	Tr 3	High fire hazard

Source: From *Preventing of Fires and Explosions in Dryers—A User Guide*, 2nd ed., Institute of Chemical Engineers, Rugby, UK (1990).

48.2.2.3 Dust Explosion Characteristics

Many dusts, when dispersed in air, form a cloud or suspension. Clouds of combustible dusts are generally explosible and, when ignited, result in an explosion in a confined space (e.g., drying chamber) or a flash fire, developing little hazardous pressure, in an open space. For a dust to be explosible, the particle size must be small enough to give an adequate reactive surface area, and the dust cloud sufficiently dense and uniform to propagate the combustion reaction. Particles greater than about 500 μm in diameter are unlikely to cause explosions, although during drying a comminution process may take place that produces fine powder much smaller in diameter. Generally, the explosion of a dust cloud can occur only when all the following conditions are met:

1. Dust concentration lies within the explosive limits
2. Required oxygen concentration is available
3. An ignition source of sufficient energy is present

Flammable dusts mixed with air exhibit an explosive range having upper and lower limits (Figure 48.4). The upper limit for dusts is difficult to establish; the lower limit, called the minimum explosive concentration (MEC), as the lowest concentration of suspended dust capable of propagating flame, may be in the range of 0.02 to 0.085 kg of product/ m^3 of air for polyethylene and coffee, respectively. Such conditions may be met in all types of dryers except for tray and band dryers. The range of explosible concentrations is strongly influenced by many parameters, e.g., the particle size, particle shape, moisture content, the position and nature of the ignition source, and more. In all dispersal dryers, in which the material is dispersed in air, it is necessary to accept that a dust cloud of explosible concentration can form.

The required oxygen concentration is expressed by minimum oxygen concentration (MOC) to support combustion; this varies between 3 and 15% (v/v)

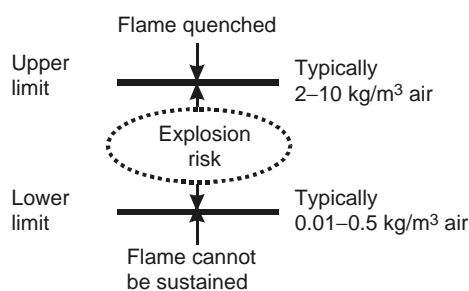


FIGURE 48.4 Dust explosivity limits.

as measured experimentally at 850°C, depending on the chemical nature of the dust, its particle size, moisture content, and temperature.

The sensitivity of a dust cloud to ignition is determined by the minimum ignition temperature (MIT) and minimum ignition energy (MIE). The MIT and MIE are determined by standard tests described in detail by Palmer [9] and Field [12].

The MIT is especially relevant in problems involving a relatively large heated area of a plant, e.g., the surface of a dryer, and represents a maximum temperature that should never be exceeded. Typical values have been determined to be 370, 500, and 575°C for sugar, cocoa, and coal, respectively.

The MIE is particularly relevant in ignition by sparks. Typical values have been shown to be 30, 120, and 50 mJ for sugar, cocoa, and coal, respectively, and are strongly dependent on the particle size: the finer the dust, the more easily it can be ignited [11]. The main application of MIE is in relation to the static electricity hazard. It has been reported that the majority of incidents occur with sparks having ignition energy below 25 mJ [13].

The course of an explosion of a combustible dust in a closed vessel is shown in Figure 48.5.

There are two main factors governing the violence of explosion: (1) the maximum explosion pressure P_{\max} and (2) the maximum rate of pressure rise $(dP/dt)_{\max}$. The severity of explosion of a dust cloud can

be described by measuring the pressure rise occurs during an explosion. As can be seen in Figure 48.5a, the pressure continues to rise inside the vessel until explosion pressure P_{\max} has been reached. In the rising portion of the pressure curve, an inflection point occurs at which the slope of the curve is the greatest. It gives the explosion rate of pressure rise $(dP/dt)_{\max}$. The P_{\max} for dryers operating at atmospheric pressure does not exceed 900 kPa and can develop rates up to 10 kPa/s. The effect of vessel volume, types of dust, and presence of flammable solvent is quantitatively shown in Figure 48.5b, Figure 48.5c, and Figure 48.5d, respectively.

The violence of explosion parameters has been determined for many years in the United Kingdom and the United States in the Hartmann bomb apparatus of 1.3-L mm capacity (Figure 48.6a) [9].

Today, an alternative method that is generally preferred, following extensive research and development in Switzerland, is the 20-L sphere test with a 10,000-J chemical igniter [11,14] (Figure 48.6b). Methods of qualifying explosion severity in the vessel have been described [22] and widely accepted in Europe [23].

These test series and data are used to determine an additional explosion parameter—the maximum rate of pressure rise normalized to a volume of 1 m³ (K_{st}). The explosion parameter is expressed in the form of the “cube root” law:

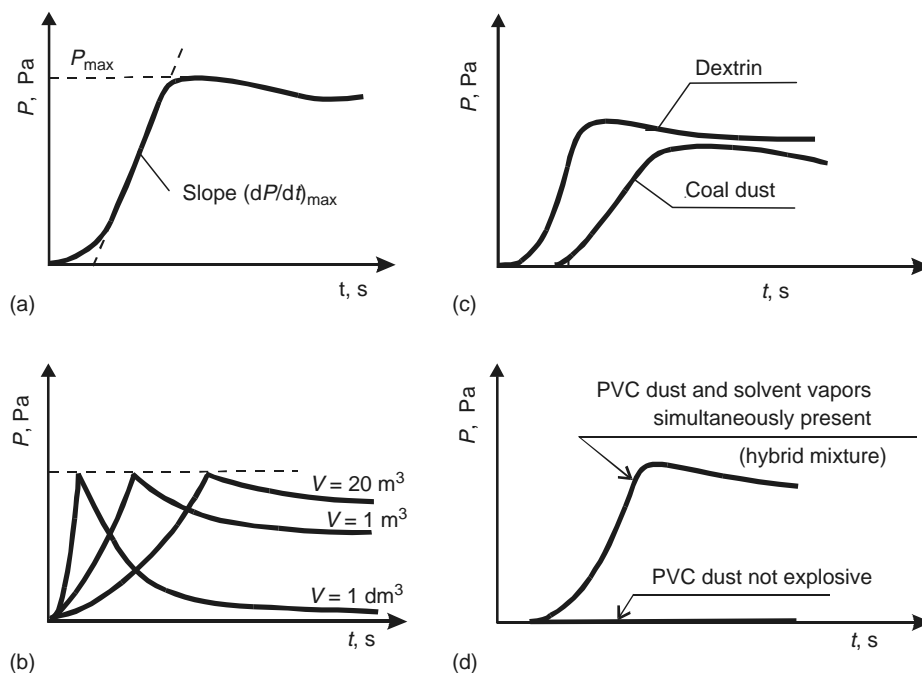


FIGURE 48.5 The course of explosion.

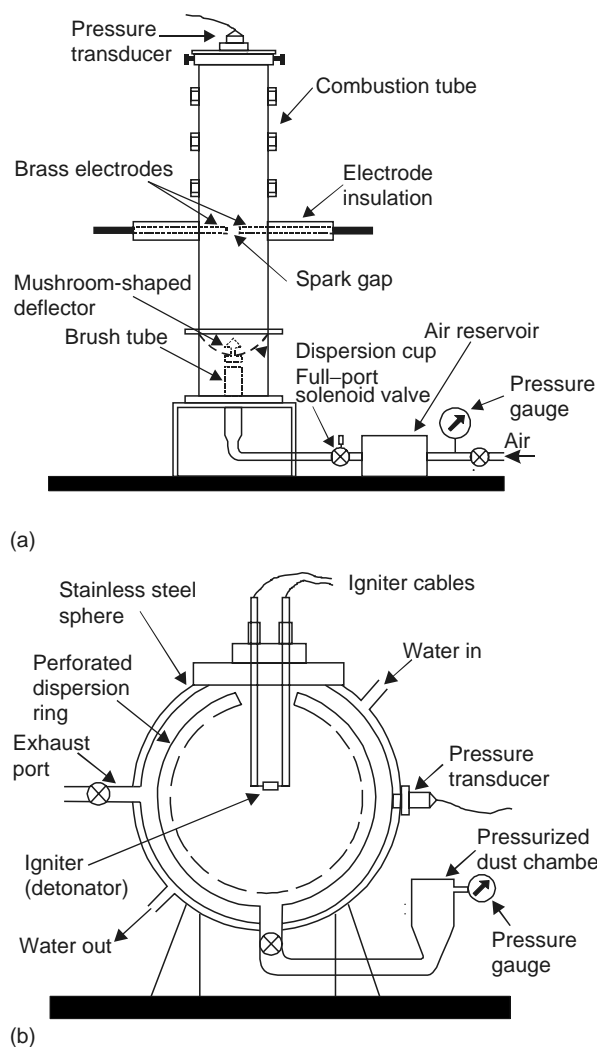


FIGURE 48.6 Apparatus for testing dryer explosion characteristics for organic powders.

$$(dP/dt)_{\max} V^{1/3} = K_{st}$$

where V is the volume of the vessel and K_{st} is the dimensional constant characteristic of a particular dust ($\text{kPa}\cdot\text{m/s}$). The maximum explosion pressure P_{\max} and the maximum rate of pressure rise $(dP/dt)_{\max}$ are used to design the explosion protection system and are discussed below.

Reported values for the explosion parameter K_{st} vary over a range for a particular material, e.g., from 2.8×10^3 to $34 \times 10^3 \text{ kPa}\cdot\text{m}\cdot\text{s}^{-1}$ for pigments and from 5.8×10^3 to $13 \times 10^3 \text{ kPa}\cdot\text{m}\cdot\text{s}^{-1}$ for milk. The differences are connected with the variations in chemical formulation of samples and differences in particle size. According to Bartknecht [14], the variation in moisture content may be less significant.

Experimental data on dust explosibility enable the classification of dust into dust explosion hazard classes. Such a classification applied in different countries is presented in Table 48.3. The determination of the dust explosion hazard class is required not only to assess dust explosion risk, but also to define statutory requirements as well as to assist in the identification of protective measures for different hazard classes. (Table 48.4), presents a selected data for dust explosibility [16-21].

48.2.2.4 Thermal Decomposition Characteristics

In almost all dryers, there is a buildup of deposits on the inner surface of the drying chamber or an accumulation of dust layer in some dead areas. The material may also be stored in bulk form at the bottom of some dryers and in bins, hoppers, silos, or bags. Ignition of such a material can occur not only by typical high-temperature sources (i.e., electrical sparks or incandescent particles), but also by

TABLE 48.3
Dust Classification Systems

Country	Test	Criterion	Dust Class	Reference No.
Great Britain	Horizontal tube <1300°C	Ignition and flame propagation	Group A (explosive)	9
	Vertical tube Inflamator		Group B (nonexplosive)	
United States	Physical properties of dust	Explosivity index IE	Group A (weak explosion); $IE = 0.1$ Group B (moderate); $IE = 0.1$ to 1.0 Group C (strong); $IE = 1.0$ to 10 Group D (severe); $IE = 10$	17
Federal Republic of Germany, Switzerland	Sphere test (volume 1 m^3)	Dimensional constant K_{st} in "cube root" law	St 0 (nonexplosive); $K_{st}/\text{kPa}\cdot\text{m/s}$ St 1 (weak); $0 < K_{st} < 200 \text{ kPa}\cdot\text{m/s}$ St 2 (moderate); $200 < K_{st} < 300 \text{ kPa}\cdot\text{m/s}$ St 3 (strong); $K_{st} > 300 \text{ kPa}\cdot\text{m/s}$	23

TABLE 48.4
Explosibility Characteristics of Various Dusts

Type of Dust	MEC, kg/m ³	MIT, °C		MIE, J	P_{\max} , kPa 10 ⁵	$(dP/dt)_{\max}$ × 10 ⁵ , kPa/s	Dust Classification	
		Cloud	Layer				IE	St
Metal powders								
Aluminum flakes	0.045	650	760	0.050	5.80	1400.0	>10.0	3
Magnesium	0.040	620	490	0.040	6.40	630.0	10.0	2
Silicon (96% Si)	0.110	790	—	0.100	6.40	840.0	0.9	1
Agricultural products								
Cocoa	0.045	470	370	0.030	5.40	211.0	13.7	1
Coffee	0.085	720	—	0.016	6.20	360.0	0.1	1
Grain dust (wheat, maize, barley, and oats)	0.055	430	230	0.030	9.20	140.0	9.2	1
Corn	0.055	400	—	0.040	7.80	413.0	6.9	1
Corn starch	0.045	400	—	0.040	7.30	516.0	9.5	1
Cotton linters	0.500	520	—	1.920	5.02	275.0	<0.1	1
Egg white	0.140	610	—	0.640	3.99	344.5	<0.1	1
Malt barley	0.055	400	250	0.035	6.54	303.0	5.5	1
Milk, skimmed	0.05	490	200	0.050	6.54	158.0	1.4	1
Potato starch	0.045	440	—	0.025	8.26	551.0	20.9	2
Rice dust	0.051	440	240	0.050	6.50	183.0	2.3	1
Soya protein	0.045	530	460	0.060	6.90	239.0	2.0	1
Sugar, powdered	0.045	370	400	0.030	7.70	352.0	9.6	1
Wood flour, pine	0.035	470	—	0.040	7.78	378.0	9.9	2
Pittsburgh coal	0.055	610	180	—	5.80	160.0	1.0	1
Plastics								
Cellulose acetate	0.040	420	—	0.015	5.85	248.0	>10.0	1
Epoxy	0.020	540	—	0.015	6.60	420.0	10.0	1
Nylon	0.030	500	430	0.020	6.70	280.0	10.0	1
Phenolformaldehyde	0.030	490	—	0.010	6.50	775.0	10.0	1
Polycarbonate	0.025	710	—	0.025	5.30	241.0	—	1
Polyurethane foam	0.025	550	—	0.015	6.60	254.0	—	1
Polyethylene	0.020	450	—	0.010	5.50	516.0	10	2
Polypropylene	0.020	420	—	0.030	5.30	356.0	10	1
Polystyrene latex	0.020	500	—	0.015	6.81	482.0	—	1
Polyvinyl acetate	0.040	550	—	0.160	4.75	689.0	—	1
Rayon	0.055	520	—	0.240	7.37	117.0	—	1
Rubber, synthetic	0.030	320	—	0.030	6.40	213.5	—	1
Urea molding comp.	0.085	460	—	0.080	6.13	248.0	1.0	1
Drugs and chemicals								
Aspirin	0.050	660	—	0.015	6.06	>689.0	2.3	2
Nitropyridone	0.045	430	—	0.035	7.64	>689.0	>10.0	2
Vitamin B	0.035	360	—	0.060	6.95	413.0	8.3	1
Adipic acid	0.035	550	—	0.060	5.78	>186.0	1.9	1
Benzoic acid	0.030	620	Melts	0.020	5.23	378.9	>10.0	1
Bisphenol A	0.020	570	—	0.015	6.13	585.6	>10.0	2
Diphenyl	0.015	500	—	0.020	5.64	689.0	>10.0	2
Salicylanilide	0.040	630	Melts	0.020	6.13	330.0	5.8	1
Sulfur	0.035	190	220	0.015	5.37	323.0	>10.0	1

low-rate atmospheric oxidation (self-heating) or thermal decomposition due to chemical reaction induced by exposure to heat.

These can take place when the rate of heat generation in the material Q_R (which strongly increases

exponentially with temperature according to Arrhenius dependence) exceeds the rate of heat losses Q_L (which increases only linearly with temperature) as shown in Figure 48.7. It may lead to runaway, increases in temperature of the material and

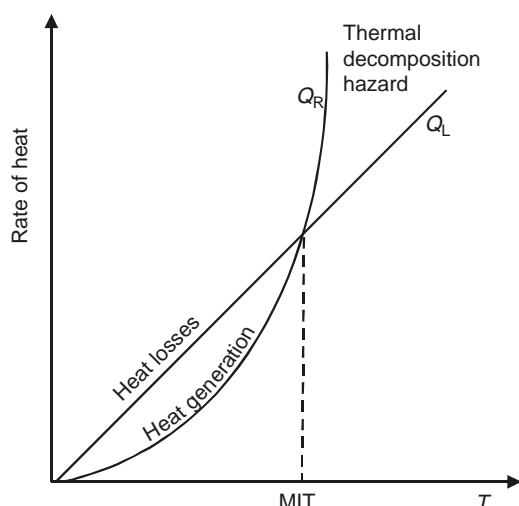


FIGURE 48.7 Thermal decomposition phenomenon.

subsequently to smoldering combustion that eventually may break into flame.

Characterization of an exothermal behavior of the material involves a determination of the temperature at which exothermicity is first detected and then the rate, magnitude, and nature of the subsequent decomposition. Such data can be produced by the application of techniques known as differential thermal analysis (DTA) and differential scanning calorimetry (DSC) [11].

These techniques have some limitations connected with the small scale of the experiments and they differentiate between conditions in the experiment and in the dryer. Therefore, ICI Ltd. of the United Kingdom has developed its own ICI Screening Test based on the measurement of the temperature at which exothermic activity is first detected during the heating

of the sample under controlled conditions (an exothermal onset temperature for the material in layer T_L ; in bulk T_B ; and aerated T_A) [15]. Test methods take into account more conditions that can be met in dryers (e.g., the effect of chemical composition of the material, physical properties, influence of external air, effect of bulk or scale, and the time cycle of the heat exposure).

For material with self-heating capabilities collected and stored in the bulk form, the most important is the discharge temperature, which cannot exceed the onset temperatures for self-heating T_C . Detailed data on the self-heating are given by Bowes [24].

There are also some important practical points of which one has to be aware:

- Contamination of the material by lubricating oil from the equipment (e.g., the atomizer in a spray dryer) decreases MIT
- During start-up and shutdown, some parts of the dryer may have a higher temperature than during normal operations
- Any smoldering layer may ignite into a flame when disturbed and as a consequence can result in a flash fire or an explosion; such deposits should be extinguished by a gentle water spray before cleaning operation begins

The summarized results of wide research made by Gibson et al. [15] for the exothermal onset temperatures for >200 samples of different organic chemicals are shown in Figure 48.8. The obtained data on the material in the form of bulk powders, layers, and aerated powders are used for the selection of safe operating conditions that simultaneously preserve good product quality.

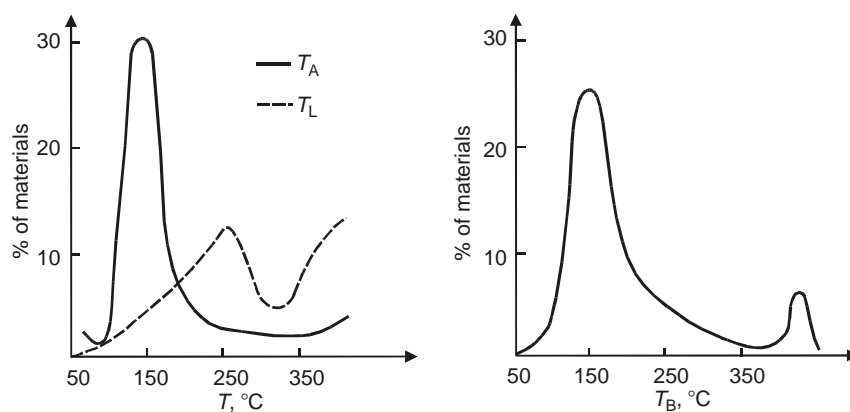


FIGURE 48.8 Distribution of material exothermal temperature T_A , T_L , T_B . (From N. Gibson, D. J. Harper, and R.L. Rogers, *Plant/Operations Progress*, 4(3): 181–189 (1985). With permission.)

48.2.3 HAZARDS WITH FLAMMABLE VAPORS

Thermal drying of any combustible solid or liquid produces a vapor that can form a flammable mixture with oxygen if only the vapor concentration is in the right range. Similar to dust explosions, a vapor explosion can occur under the following conditions:

- A vapor concentration is within lower flammability limits (LFLs) and upper flammability limits (UFLs)
- Minimal oxygen concentration (MOC) is reached
- An ignition source of sufficient energy is present

The presence of the flammable vapor in air makes the mixture more susceptible to ignition and the propagation of flame through the mixture of reactants is even more rigorous than that for dusts dependent on the overall conditions.

The general relationship between flammability parameters of an idealized flammable air mixture is shown in Figure 48.9, whereas Table 48.5 presents some flammability data for selected organic liquids.

The terms lower explosibility limit (LEL) and upper explosibility limit (UEL) are also used; they are synonymous with LFL and UFL, respectively.

The LFL and UFL can be determined experimentally and LFL, which is more relevant to drying for

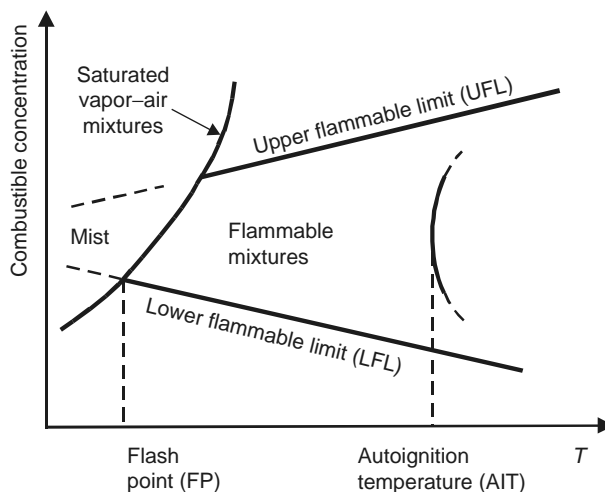


FIGURE 48.9 Flammability characteristics of an idealized vapor-air mixture.

most organic solvents in air, is around 1 to 2% by volume at room temperature. The flammability limits can also be calculated on the basis of the stoichiometric amount required to sustain combustion.

Adequate equations for flash calculation are set up in Table 48.6.

The LFL is largely independent of the oxygen concentration in the gas, whereas the upper limit is strongly dependent. These are also affected by tem-

TABLE 48.5
Flammability Data for Selected Organic Liquids

Compound	Limit of Flammability in Air, % v/v		Flash Point, °C	Boiling Point, °C	Autoignition Temperature, °C	Minimum O ₂ for Ignition (Average), % v/v ^a
	Lower	Upper				
Ethanol	3.3	19.0	13	78	363	10.5
Methanol	6.0	36.0	12	64	385	10.0
Isopropanol	2.0	12.7	12	83	399	—
Isobutanol	1.7	10.6	28	107	415	—
<i>n</i> -Butanol	1.4	11.2	29		343	—
Acetone	2.15	13.0	−20	56	465	13.5
Diethylether	1.9	36.0	−45	35	160	10.5
Ethyl acetate	2.0	11.5	−4	77	426	b
<i>n</i> -Hexane	1.1	7.5	−22	69	223	12.0
<i>n</i> -Heptane	1.05	6.7	−4	98	204	—
<i>n</i> -Octane	1.0	6.5	13	126	206	—
Gasoline	1.4	7.4	−38	38–204	180–456	—
Benzene	1.3	7.1	−11	80	498	11.2
Toluene	1.3	7.1	4	111	480	—
<i>m</i> -Xylene	1.1	7.0	27	139	527	—

^aThe data on minimum O₂ for ignition are for nitrogen as the inert diluent.

Source: From Fire Hazard Properties of Flammable Liquids, Gases and Volatile Solids, *NFPA No. 325M*, National Fire Protection Association, Boston (1977).

TABLE 48.6
Formulas for Calculation of Explosibility Parameters

Description	Equation	Remarks	Reference No.
Flash point of liquids, T_{FP} , °C	$T_{FP} = 0.683T_{BP} - 71.7, \text{ or}$ $T_{FP} = f(P_{FP}) \frac{P}{1 + (N - 1)4.76}$ $P_{FP} = \frac{P}{1 + (N - 1)4.76}; T_{FP} = f(P_{FP})$	<p>Recommended for hydrocarbons; P_{FP} = vapor pressure at closed-cup flash point;</p> <p>T_{BP} = boiling point, °C N—number of atoms of oxygen for combustion of one molecule of fuel</p>	25
Flammability limits of vapor, % v/v	$LFL = \frac{P_{FP}}{P} 100$ $LFL = \frac{0.55 \cdot 100}{4.76M + 1.19X - 2.38Y + 1}$ $UFL = \frac{3.50 \cdot 100}{4.76M + 1.19X - 2.38Y + 1}$	<p>M, X, Y = stoichiometric number of atoms of carbon, hydrogen, and oxygen, respectively, in molecule of fuel</p> <p>Recommended for hydrocarbons Deviation for nonorganic compounds</p>	26
Flammable vapor mixture LeChatelier's rule	$LFL_m = \frac{1}{\sum_{i=1}^n \frac{Y_i}{LFL_i}}; UFL = \frac{1}{\sum_{i=1}^n \frac{Y_i}{UFL_i}}$	<p>Y_i = mole fraction of component i n = number of combustible species</p>	
Mixture of flammable vapor (A) with nonflammable component (B)	$LFL_m = LFL_A \frac{\left(1 - \frac{X_B}{100 - X_B} 100\right)}{100 + LFL_A \frac{X_B}{100 - X_B}}$	<p>X_B = % v/v of nonflammable component B in mixture</p>	
Flammability limits of vapor Dependence on temperature T , °C Dependence on pressure	$LFL_T = LFL_{25}[1 - 0.75(T - 25)/\Delta H_c]$ $LFL_T = LFL_{25}[1 + 0.75(T - 25)/\Delta H_c]$ $UFL_P = UFL + 20.6(\lg P + 1)$	<p>ΔH_c = heat of combustion, kcal/mol</p> <p>P = absolute pressure, MPa</p>	
Minimum oxygen concentration, % v/v O_2	$MOC = LFL \left(M + \frac{X}{4} - \frac{Y}{2} \right)$	Stoichiometric relation	25
Minimum oxygen concentration: dependence on pressure P	$MOC_P = MOC - 1.5(\lg P + 1)$	P = absolute pressure, MPa	
Minimum nitrogen concentration, % v/v N_2	$MNC = \frac{(100 - LFL)}{4.76(M + \frac{X}{4} - \frac{Y}{2})LFL} - 1$		
Minimum explosible concentration for hybrid mixture	$MEC_H = MEC \left(\frac{x_G}{LFL} - 1 \right)$	<p>MEC—minimum dust explosible concentration x_G = total % v/v of flammable vapor</p>	14

perature (the flammability range increases linearly with the rise of temperature) and pressure (increased pressure widens the flammability limits).

If the air is diluted with an inert gas, e.g., nitrogen or carbon dioxide, the risk of ignition is reduced. The MOC for ignition varies from 10 to 13.5% by volume, depending on the type of the vapor and the inert gas. This feature has been used to develop self-inertizing dryers.

The most potentially hazardous situations are likely to occur when flammable substances are used at temperatures above the flash point, which corresponds to the vapor concentration at LFL. This situation frequently occurs for solvents encountered in drying because most of them have the flash point below room temperature.

Therefore, a careful control of the vapor or oxygen concentration is required even in a feed container when solvent-wet feedstocks are handled.

48.2.4 HYBRID MIXTURES

A combination of combustible dust and flammable gas or vapor called a *hybrid mixture* is of significance in certain drying operations. Such dust–vapor–air mixtures are much more susceptible to ignition and any resulting explosion is more violent than the dust-alone explosion. Research carried out so far has established that:

- An explosion may occur even if the concentration of vapor is well below the LFL and the dust–air mixture is nonexplosive
- A minimum vapor concentration of the order of 0.2% by volume may cause a marked increase of P_{\max} and $(dP/dt)_{\max}$
- Low energy ignition incapable of igniting a given dust–air mixture may easily cause an ignition of a hybrid mixture

The properties of hybrid mixture explosions are not yet available; therefore, until more information is available, all such mixtures should be treated with great concern.

48.2.5 IDENTIFICATION OF IGNITION SOURCES

All kinds of ignition sources can be present in dryers. However, for a particular type of dryer, there are some major sources with a higher risk of occurrence that should be well recognized, localized, and possibly eliminated. An identification of all ignition sources and their conjunction to hazardous areas is particularly important when a certain type of hazard takes place. For the areas where flammable vapors and dusts in

both dispersions and layers can be found, the most common and hazardous ignition sources include electrostatic discharges and electric frictional sparks. Electrostatic discharges can be reduced substantially by a careful design and application of the approved codes of practice [11,27,28]. The accumulation of electrostatic charge by product and equipment is in many cases directly associated with the process and operations involved in industrial production. It represents a source of ignition that can be hazardous not only under normal circumstances, i.e., without any kind of plant upset occurring, but also under abnormal conditions. Dust layers or bulk accumulates are especially sensitive to self-heating or exothermic decomposition. The risk may be reduced by ensuring proper operating conditions or by design like “deposit-free” spray dryer [29]. Since it is not possible to ensure that all ignition sources are eliminated, it is necessary to introduce a proper protection system.

48.2.6 SELECTION OF OPERATING CONDITIONS

The operating conditions for a chosen type of dryer can be selected taking into account the type of hazard that exists, the particular part of a drying system where the hazard takes place, and the appropriate fire and explosion characteristics.

A fire hazard may be safely avoided if the temperature of the material is lower than the relevant minimum ignition or exotherm onset temperature. The greatest risk exists in the part of the drying system where the product moisture content is lowest, such as the separation units (e.g., cyclones and bags). Where dust deposits can accumulate, the requirements are that the maximum permissible temperature of the plant surfaces should not exceed two thirds of the layer MIT. Also, for the dust layer, the time the material is exposed to heat should be taken into account as a controlling factor that will determine cleaning and removal of deposits.

Therefore, an automatic control system that monitors the temperature at selected locations is strongly recommended. The explosion hazards will be successfully eliminated if the dust concentration is less than the minimum explosible concentration. This is difficult to achieve for most dispersional dryers.

The onset exotherm temperature (e.g., layer or bulk) is to be lower than MIT (e.g., layer or bulk) and is proposed for the cases in which the change of material quality occurs before the ignition takes place. The summary of recommended conditions for safe plant operation is presented in Table 48.7.

The above proposal is based on the characteristic controlling parameters for a certain type of hazard

TABLE 48.7
Selection of Operating Conditions

Ignition Hazard	Basic Controlling Parameter	Recommended Operational Temperature T , Dust Concentration DC, Vapor and Oxygen Concentration VC and OC, Respectively	Type of Equipment
Dust–air mixture	Minimum ignition temperature MIT	$T = \text{MIT} - 50^\circ\text{C}$	Spray dryer, fluid dryer, pneumatic dryer
	Onset aerated exotherm temperature T_A	$T = T_A - (30\text{--}50^\circ\text{C})$	Fluid dryer, band dryer
	Maximum permissible oxygen concentration MOC	$\text{OC} \leq (50\text{--}70\%) \text{ MOC}$	Inside dryer and dust recovery system
	Minimum explosible concentration MEC	$\text{DC} < \text{MEC}$	It is not possible to achieve this for most dryers except for tray and band dryers
Dust layer or deposits	Onset layer exotherm temperature T_L	$T = T_L - 20^\circ\text{C}$	Band and tray dryers and deposits
Bulk dust	Onset bulk exotherm temperature T_B	$T = T_B - 50^\circ\text{C}$	Bins, hoppers, silos, and bags
Flammable vapor	Lower explosive limit LFL	$\text{VC} < 25\% \text{ LFL}$	Whole drying system
	Minimum oxygen concentration MOC	$\text{OC} \leq (50\text{--}70\%) \text{ MOC}$	Whole drying system
	Minimum ignition temperature MIT	$T < \text{MIT}$	Whole drying system

with a significant safety margin. Selection of a maximum drying temperature on the basis of the onset exotherm temperature has been verified extensively by Gibson et al. [15]; the temperatures specified do not place an unacceptable restriction on drying operations. They have continued to be efficient and economic.

48.3 THE ENGINEERING FACTOR

48.3.1 DESIGN CONSIDERATION

Within the design consideration factor, probably the most important section is plant layout. Taking into account the possibility of dust explosion and fire, a question arises whether a certain type of plant may be located in a particular place or whether some splitting of the process on particular operations may be required. The increased scale of many new industrial plants means that the volume of the plant providing a risk of explosion is much greater than it was only a few years ago.

Therefore, there is a tendency for a large plant to be situated in an open space. The best arrangement is to put the plant in a separate structure covered only with a light roof. If this is not possible, attention should be given to the situation of collection and storage, units and furnaces. All those units as far as possible should be outside the building or inside a fire-resistant enclosure. Also essential is the position of the central room and electrical switch rooms and how they will be affected by fire and explosion.

There are other engineering factors that affect the fire and explosion hazard, e.g., engineering standards of the structural steel and foundations, process equipment, heat exchangers, feeding system, fan and blowers, storage vessels, electrical equipment, instruments, and fire protection and safety equipment. Considerable assistance in design also can be obtained from relevant codes of practice. The responsibility for safe operation rests with the manufacturers of equipment and products as required by national law (e.g., Factories Act and Health and Safety at Work Act in the United Kingdom).

48.3.2 FIRE AND EXPLOSION PRECAUTIONS

The basis of safety in plants drying flammable materials can be prevention and protection of life and property to the maximum possible extent, assuming that explosion is inevitable. Fire and explosion prevention can be accomplished by

1. Maintaining the concentration of the flammable material beyond the range of flammability
2. Maintaining oxygen concentration within safe limits
3. Excluding all probable ignition sources

Fire and explosion protection can be achieved by

1. Proper venting
2. Containing the severity of explosion
3. Suppressing the fire or/and explosion

It is to be noted that there is no need to duplicate any method of protection or prevention. Methods of dust explosion prevention and protection are described in three Institute of Chemical Engineers (IChemE) guides [30–32] and in References 33 and 34.

48.3.3 FIRE AND EXPLOSION PREVENTION

For materials undergoing exothermic oxidation, the methods given below will be satisfactory for explosion prevention, whereas for materials undergoing exothermic decomposition resulting in the release of large volumes of gases, these methods can be used only to protect, not to prevent, explosion hazards.

48.3.3.1 Inerting

The principle of inerting is to create an environment in which combustion cannot take place. Some of the common inerting agents are N_2 , CO_2 , flue gases, and others. The presence of inert gases causes the concentration of oxygen in the gaseous mixture to be reduced below the minimum required for sustaining combustion, which is used to be within the range of 10% by volume. Inerting also decreases the maximum explosion pressure and retards the rate of pressure rise [14].

The MOC determined in the laboratory at 850°C is a very safe value for most of the dryer applications in which the operating conditions are never as severe. Actually, the use of this value results in an unnecessary high consumption of the inertizing agent. Hence, tests should be carried out at conditions that actually exist in the dryer. Some data on MOC are presented in Reference 9.

Recycling a part of the dryer exhaust gases not only improves the thermal efficiency, but also helps in reducing the oxygen concentration. Such dryers are called *self-inertizing*. An example of such a system is presented in Figure 48.10.

One more important point is that, during the start-up of a self-inertizing dryer, initially and for some time the oxygen concentration is above the minimum value desired, so the solid feed should not be started until the concentration reaches the desired limit. However, when organic solvents are involved, the drying air must be substituted by inert gas (e.g., nitrogen), which recirculates in a closed-cycle system (Figure 48.11).

48.3.3.2 Avoiding Formation of Dust Clouds

The only safe way of avoiding formation of dust clouds is to use air velocities at which no entrainment of dust takes place. In band and tray dryers, the control of velocity is easier, but this is difficult to fulfill in dryers that involve suspended-particle processing. In general, the nature of the drying operation is such that dust entrainment is not easy or feasible for various reasons.

48.3.3.3 Excluding All Ignition Sources

It is imperative to ensure that all the following probable sources of ignition are either removed or controlled to prevent explosion.

Direct Heating Systems—For dryers containing flammable vapors, direct heating systems are too dangerous to be employed. In such cases, a separate hot air generator should be provided.

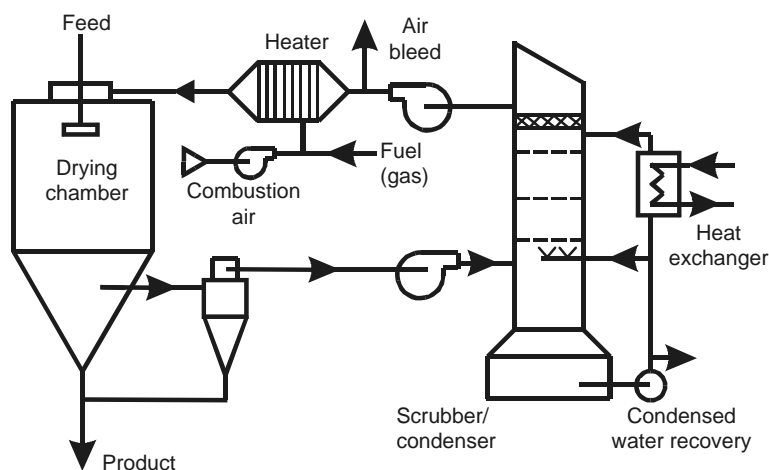


FIGURE 48.10 Semiclosed cycle (self-inertizing) spray dryer system.

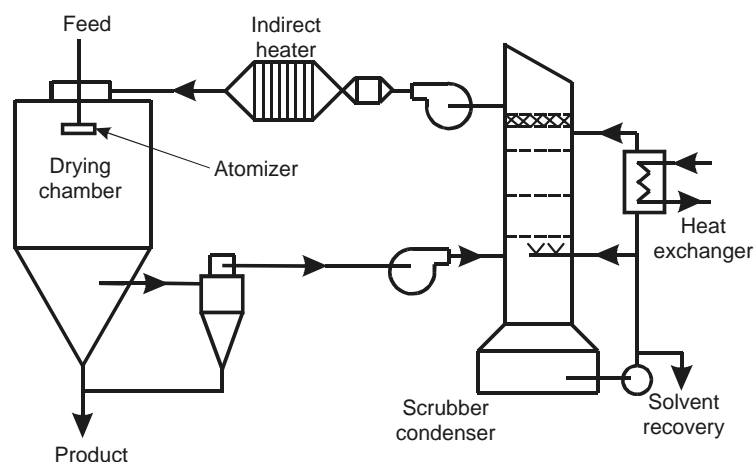


FIGURE 48.11 Closed-cycle spray dryer system.

Hot particles from a heater can also be a potential risk of ignition. According to accepted practice, the following precautions should be taken:

1. If recirculation of a part of the exhaust gas is considered, it is important to make sure that no dust particles enter the combustion chamber
2. The burners should be cleaned periodically, and efforts should be made to operate the burners at a proper air/fuel ratio to ensure complete combustion
3. The primary and secondary air fans should be located to take in dust-free air. If this is not practical, the inlet side of these fans should contain dust filters, which should be periodically checked and cleaned.

Electrostatic Discharge—Any metal or good conductor of electricity can store enough, electrical energy to exceed minimum spark ignition energy for dust and vapor clouds (25 mJ for most of the hydrocarbons). To prevent such charge accumulation, all the conducting elements should be grounded. Periodic checks are necessary to ensure proper inert grounding.

Similarly, nonconductors of high resistivity also pose a serious problem. In such cases, the resistance has to be reduced by incorporating some conductive elements. For example, the extremely high resistance of synthetic fibers used in bag filters may be reduced by adding some steel or carbon fibers.

Electrical Equipment—Contact switches, fuses, circuit breakers, and so on can discharge spark energy greater than MIE, thereby causing explosion in an environment containing flammable dust or vapor. It is necessary to make sure that such electrical equipment does not come in contact with dust or flammable vapor.

Frictional Sparks and Heating—For frictional sparks and heating, the probable sources of hazards are: (1) overheated bearings; (2) power drills and similar tools in which localized heating is likely; (3) frictional impact with shovels and scoops, which may cause ignition of compounds with low MIE, e.g., peroxides and sulfur; (4) rotating fan blades touching the casing; and (5) tramp metals and stones fed with the solid into the dryer. For all these hazards, standard procedures are available to take care of various problems [8].

Autoignition—It is worthwhile taking all precautions to ensure that at no stage of drying operation, including start-up and shutdown, does the material temperature exceed its autoignition temperature (AIT). The exhaust gas temperature is a safe way of controlling the material temperature in a dryer.

Also, ledges, corners, and crevices, e.g., should be avoided as far as possible to prevent dust layer formation. If unavoidable, periodic cleaning of the dust deposit is essential.

The decision tree for ignition sources assessment is shown in [Figure 48.12](#).

48.3.4 EXPLOSION PROTECTION

Explosion protection measures expect the explosion to take its course, but ensure that it does so safely without any effects on the personnel and equipment. The three possible ways of protection from a fire or explosion are: (1) venting; (2) containing; and (3) suppressing.

48.3.4.1 Venting

The principle of operation of venting is that at a predetermined pressure rise P_{st} the vent in the system opens up, thereby releasing the explosion products to a safe area. The course of vented explosion is

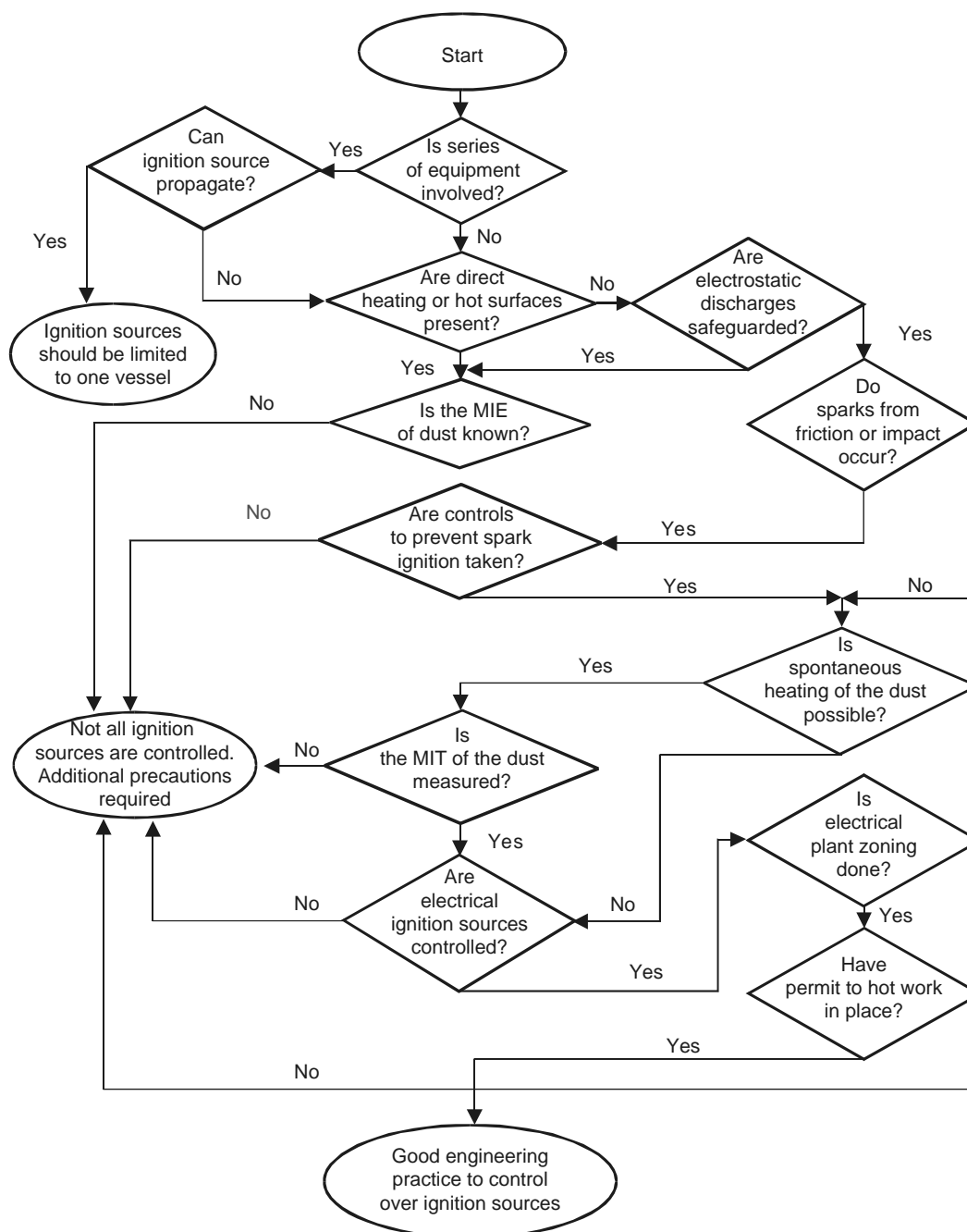


FIGURE 48.12 The decision tree for ignition sources assessment.

shown in [Figure 48.13](#). Venting is a common, often cheapest, way of protection from explosion. In case the explosion products are toxic and cannot be released in the immediate vicinity of the plant, the explosion products are carried to a safe area by ducts connected to the vents. Since most dryers operate at low or moderate pressures, a suitable vent-opening pressure rise is 10 kPa. It is useful to note

that, when vented ducting is used, the peak pressure can be increased. The increase in the peak pressure rise is directly proportional to the square of the length of the ductings.

The important criteria for vent design and selection are that the vent should open fully at the designated pressure with very little inertia and should be perfectly sealed when closed. The principles and

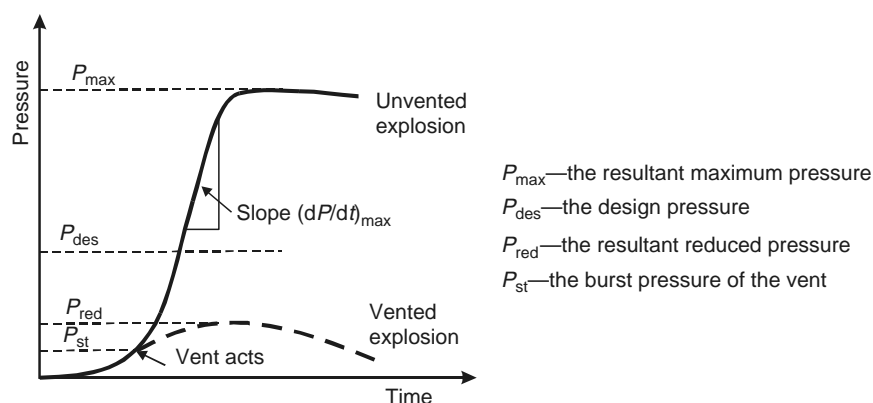


FIGURE 48.13 The course of vented explosion.

practice of venting have been widely discussed [9–15, 23,25,26,32–34]. The main question is vent design. Unfortunately, very little information is available on vent design for dryers. Moreover, high-temperature, flammable dusts, and turbulent contact between gas and solid further complicate the use of available information on vent design in general. Three criteria have been developed for vent area calculation: (1) vent ratio; (2) K factor; and (3) cube root law.

The vent ratio is defined as the ratio of vent area to the volume of the vessel. Based on the maximum rate of pressure rise in a 1.2-L Hartmann bomb, for vessels up to about 30 m³, the vent ratio can be selected from Table 48.8. For most carbonaceous dusts, 1:5 seems to be a safe vent ratio.

For volumes exceeding 30 m³, the vent ratio should be progressively reduced to a limiting value of 1:25 m^{−1} for a volume of 283 m³.

For estimating vent areas, the concept of K factor was developed by Simmonds and Cubbage [35]. They defined K as

$$K = \frac{\text{cross-sectional area of vessel}}{\text{vent area}} = \frac{A_C}{A_V}$$

TABLE 48.8
Recommended Vent Ratio

Maximum Rate of Pressure Rise, kPa/s	Vent Ratio, m ^{−1}	Dust Example
<35,000	1/7	Wood, coal, flour, and coffee
35,000–70,000	1/5	Plastics
<70,000	1/3	Magnesium and aluminum

The value of K depends upon the designed strength of the vessel and the vent-opening pressure. For a vent-opening pressure rise of 10 kPa and a maximum explosion pressure rise within the vessel of 20 kPa such that the designed strength of the dryer is 40 kPa (twice the maximum pressure rise), the value of K can be taken as 3. This value is recommended for the chemical plant handling dusts with dP/dt values up to 700 to 800 kPa/s measured in a 1.2-L Hartmann bomb [27]. If the designed strength required for the vessel decreases, the value of K has to be reduced. The variation of maximum pressure rise with K factor is given by Gibson and Harris [36].

In the United States, Germany, and Switzerland, the recommended vent areas are based on the experimentally measured dependence of the rate of pressure rise on the vessel volume, the “cube root law” defined in Section 2.2. From large-scale venting experiments, a series of normographs (or a tabular form) have been derived and incorporated in the German standard VDI 3673 [23]. Based on an original work of Bartknecht [14], the National Fire Protection Association of the United States [34] presents such a series of normographs that allow the vent area F to be determined for the products according to their explosion characteristics K_{st} for dust and K_G for vapor, vessel volume V , the vent opening pressure P_{stat} , and maximum pressure during venting P_{red} . Typical normographs are shown in Figure 48.14. An adequate equation for the sizing of a relief vent can be found in the monograph of Crawl and Louvar [26]. Table 48.9 presents equations for the vent area F_1 , required for volume V_1 , scaled up to that of F_2 for volume V_2 .

A wide variety of explosion relief venting is available to the industry. Bursting disks, displacement panels, and hinged covers or doors with spring-loaded or magnetic catches and automatic trigger vents are all in use. Selection of the specific type should be based on

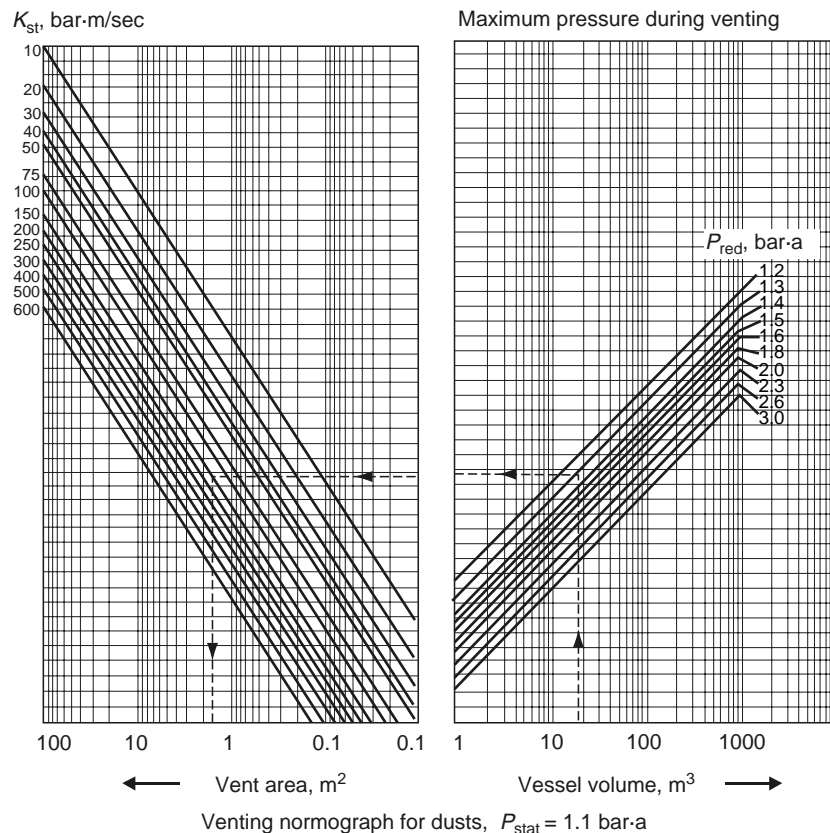


FIGURE 48.14 Normographs for the sizing of vent relief explosion. (From Explosion Venting, *NEPA No. 68*, National Fire Protection Association, Quincy, MA, 1988 (presently under revision by the NFPA Explosion Protection Systems Committee.)

cost, operating conditions, and vessel type. Figure 48.15 presents some commonly used explosion vents.

It must be recognized that during the normal course of explosion, a significant volume of flame will discharge through the vent opening. For this reason, the equipment utilizing venting should be located outdoors or close to external walls so that the vent duct will be as short as possible, avoiding any bends. Discharge must be to a safe area. Typical location of vents in a selected drying system is shown in Figure 48.16.

Ideally, the best location for an explosion-relief vent would be as close as possible to the source of ignition; however, predicting the ignition source is

nearly impossible. Realistically, the vent should be positioned where it is least affected by internal fittings, which could increase turbulence if the flame front has to pass through them. The vent may be on the side or the top of the equipment.

48.3.4.2 Suppression

Explosion suppression is a technique that is activated when the start of a dust explosion is detected (usually within the first 10 ms) and arrested during incipient stages, thus preventing the development of a pressure that could result in an explosion. The pressure/time curve for suppressed and unsuppressed explosion is similar to that for venting (Figure 48.13).

Commercially available equipment for detection and discharge suppression has been successfully applied even for the very large volumes typical of silos and spray dryers [32].

A typical suppression system operates on the basic principle shown in Figure 48.17. The detector 1 placed inside the dryer casing detects the pressure wave, before the outburst of flames, and induces the detonating cap valve 2 to blow the extinguish powder

TABLE 48.9
Scale-Up Equations for Vent Area Calculation

Scale-Up Equations	Basis	Recommended for
$F_2/F_1 = V_2/V_1$	Vent ratio	Dust explosions
$F_2/F_1 = A_2/A_1$	K factor	Mainly for vapor explosions
$F_2/F_1 = (V_2/V_1)^{2/3}$	Cube root law	Dust and vapor explosions

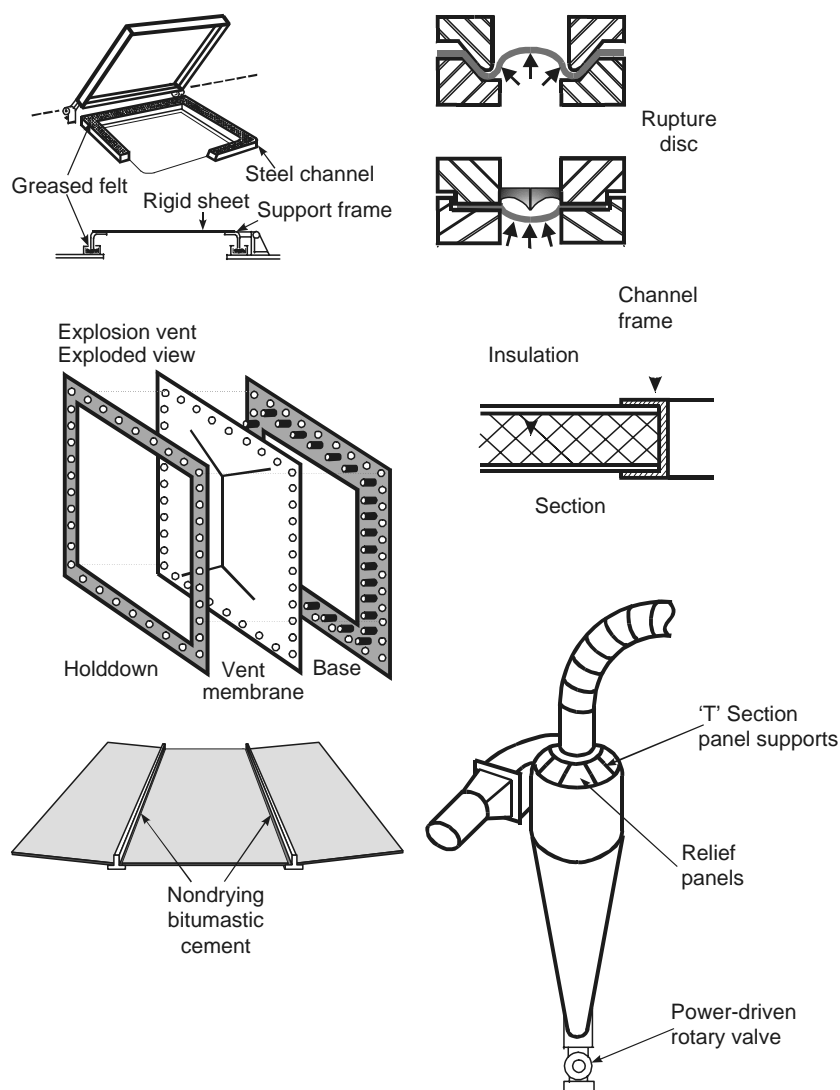


FIGURE 48.15 Types of explosion vents.

contained in the bin 3 evenly into the apparatus casing through the ball nozzle 4. The developing explosion is thus so rapidly being attacked that it cannot reach its fullest extent, and therefore causes a relative low pressure rise of only 0.1 to 0.6 bar. An efficient suppression system will keep the maximum explosion pressure rise below 10 kPa and the number of suppressors depends on the volume of the plant and airflow rate, and, according to Bartknecht [14], follows the cube root law ($n \sim V^{2/3}$). Suitable suppressants are halogenated hydrocarbons or dried powder, e.g., sodium bicarbonate, fluorides, cryolite, or ammonium phosphate. Generally, powder is more effective than halogenated hydrocarbons.

Automatic suppression systems are of particular use when the dust is toxic and safe discharge of the explosion products from a vent is not permitted. It is

also important to bear in mind that the suppressant is not effective against exothermic decomposition reactions resulting in large volumes of gas.

Generally, a limitation of this technique is the path along which the suppressants can be expected to travel with a possibly maximum value of 4 m [12], although Moore [37] reported a successful testing on a 250-m³ scale. As a general recommendation, it can be stated that the explosion-suppression systems are effective against most solvent-vapor explosions, hybrid-mixture explosions, fuel-droplet explosions, and dust explosions, provided that the explosibility constants K_{st} and K_G are <300 Pa·m/s. For more severe materials and for weak plants that are located within a building, e.g., a larger spray dryer plant, the explosion suppression and the explosion venting combination are alternatives.

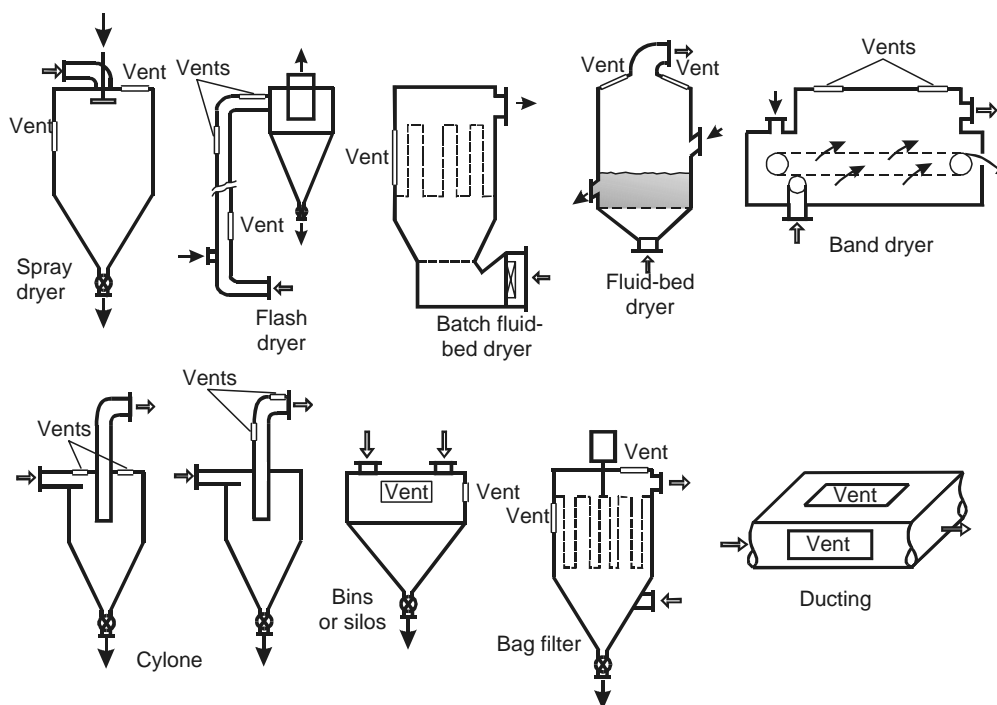


FIGURE 48.16 Location of relief vent in various drying systems.

48.3.4.3 Containment

Containment involves designing the drying system to sustain the maximum pressure rise. For dryers operating at or near atmospheric pressure, this may result in a vessel design pressure exceeding 900 kPa. As expected, this is an expensive solution and is worth considering only for small dryers and for some cases of vacuum drying.

48.3.4.4 Explosion Isolation Systems

Suppression or containment systems can be operated in conjunction with rapid-acting explosion isolation systems that isolate several interconnected vessels so an explosion in one vessel does not propagate through ducting to initiate an explosion in another vessel. Two types of isolation systems are used: (1) mechanical and (2) extinguishing, both based on a fast-acting knife gate valve principle. For details, please refer Reference 38.

48.3.5 FIRE DUST PROTECTION

Protection against fire can be obtained by all the means used for explosions and also by typical fire-fighting systems using appropriate extinguishers. The following methods may be applied:

- Gentle water spray
- Inert gas blanket
- Steam blanket
- Isolation and an external water spray

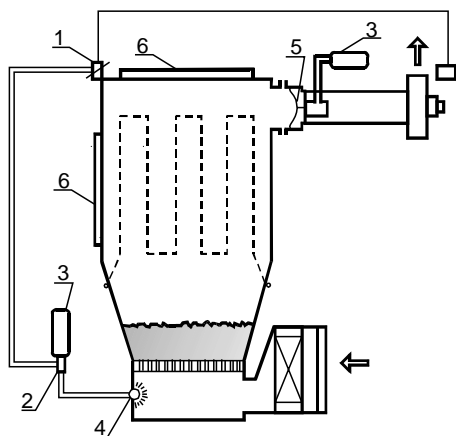


FIGURE 48.17 Suppression technique in a fluid-bed dryer. (1, detector; 2, cap valve; 3, bin; 4, ball nozzle; 5, screen containing powder; 6, explosion vents).

The selection of an extinguisher depends on factors such as the composition of the dust, its situation, and the presence of other hazardous materials. The most commonly used extinguisher, water, cannot be used for fire involving electric equipment or reactive metals (e.g., aluminum or magnesium). Gases or vapors such

as carbon dioxide, nitrogen, or steam may be used as extinguishing agents for dust fires, especially for dust that is contained in a relatively gas-tight volume. Such a vessel should be cut off from adjacent equipment. An alternative to extinguish fire is to allow it to burn itself out after isolating the vessel concerned, keeping a spray of cooling water on the outside of the vessel.

The application of any fire-fighting systems requires that the dust should not be disturbed into a cloud because burning material is present and may cause an explosion.

48.3.6 EXPLOSION HAZARDS AND PROTECTION IN ANCILLARY UNITS

Almost all dryers have to be provided with some type of heating system. Moreover, continuous-duty dryers require feeding and discharging arrangements as well as dust collection units. Therefore, it is necessary to consider some of the potential fire and explosion hazards associated with these systems and steps required to protect them from the hazards.

A comprehensive summary of explosion hazards and a respective safety measure are presented in

Table 48.10. It should be noted that these suggestions may be changed or modified depending upon the specific application and dust properties.

48.3.7 RELIABILITY OF THE PROTECTIVE SYSTEMS

An assured level of safety in a drying operation depends essentially on the reliability of the protective systems. This means that the system should function at the time of hazard under given conditions without failure. However, no single system is perfectly reliable. A hazard results when a protective system is called on to act during a dead period (the fraction of the time that a protective system is inactive).

On the basis of the hazard analysis (Hazan) [39,40], the mathematical formula for the hazard rate H caused by failure of the protective system is given by

$$H = f^n T^{n-1} [1 - \exp(-DT/n + 1)]$$

where f = the failure rate of the protective system to cope, T = the test interval between maintenance checks, D = the demand rate, and n = the number of systems.

TABLE 48.10
Explosion Hazards and Safety Measures in Ancillary Units

	Heating System		Feeding System	Dust Conveying and Recovery
	Direct	Indirect		
Major hazard	Flame and burning particle	No	Scrap metal and stones	Dust deposits
Ignition source	Hot particles	No	Sparks	Static electricity and self-ignition
Safety measures	Screen to stop particles	Pressure relief valves	Magnetic or pneumatic separators	Venting or suppression for toxic dust
For water-wetted dust	Automatic spark detection system Extinguishing system			
For flammable solvent	Not applicable	Inerting or suppression	Inerting	Inerting or suppression
Basic design	Correct air/fuel ratio	Interlocked with	Uniform feeding rate	Air velocity at
considerations	Interlocked with the airflow	the feeding system	Interlocked with heater	least 20 m/s in conveying system
			Inlet and outlet Interlocks	Bounding and grounding
			Effective seal	Incorporating carbon and metal filaments in fabric filter
Maintenance and operation	Purging before start	Inlet air filter cleaning	Good house keeping and inspection	Periodical cleaning and inspection
	Burners cleaning	Outside of heating tubes		

Using this formula and a typical figure on the relief valve failure ($f = 0.01/y$), test interval $T = 1$ y, demand rate $D = 1/y$, and the hazard rate $H = 0.00393$ (once in 254 y for $n = 1$). The value of demand rate D influences the extent of protective measures to prevent drying hazard, whereas the failure rate depends on the design and the maintenance of the protective system. When a single protective system is unreliable, giving an unacceptable hazard rate, the duplication or triplication of the system is recommended. The more accurate and reliable protection system may reduce significantly the safety margin in operating conditions, which may bring additional economic effects.

48.3.8 APPLICATION TO SOME COMMON DRYERS

A number of papers published so far on safety aspects refer to the spray dryer [41–44] and the fluid dryer [45–47]. Few data refer to other types of dryer [48,49], like the contact dryer, which generally can be safeguarded easier than convective dryers. Table 48.11 provides a summary of the probable hazards associated with a few industrially common types of dryers and some hints on the recommended safety measures to be taken. Once again, it is noted that this tabulated information on measures and hazards is not exhaustive. The suggested measures may have to be changed or modified depending upon the specific application at hand.

48.3.9 A NEW DEVELOPMENT IN SAFETY DRYER ASSESSMENT

Drying is an operation that may produce a combustible powder or dust that can cause an explosion under the drying conditions or produce a flammable solvent that may form an explosive vapor–gas mixture. As well as from fluids and gases, hazards can also result from solids. For such conditions, the machines, equipment and protection systems are under EU ATEX 10A Directive 94/9/EC, which is in force since 2003 [7]. There are two main requirements of that directive:

1. Categorization of the machines, equipment and assignment of an appropriate level of protection
2. Compulsory to carry out risk assessment to identify and evaluate any relevant hazard and on that basis eliminate or minimize the risk by an appropriate design measure and protective devices.

Dryers are ATEX category 1, therefore, level of protection must be very high. On the other hand, the

reliability of any safety-critical control system (protection system) according to standard IEC 61,508 [50] must represent a particular safety integrity level, i.e., “SIL.”

There are four SIL numbers from 1 to 4 expressing the range of probability of failure on demand. For category 1 (dryers) the overall protection system SIL3 is assigned, i.e., the probability of failure on demand is from 10^{-4} to $10^{-3}/y$. These are basic guidelines to the risk assessment and selection of particular protective devices.

The drying risk assessment methodology is discussed by Rogers et al [51] as well as by Markowski and Mujumdar [52]. Safety assessment of drying processes can also be done using the expert system [53].

48.4 THE MANAGEMENT FACTOR

The normal operation of the plant should be safe, efficient, and economical. Safety can be effectively achieved by correct operation and maintenance of the equipment. Hazardous situations can arise in the best-designed plant if it is improperly operated or inadequately maintained, or incorrect instructions are given. Human error accounts for over 90% of industrial accidents [51]. Within the management factors, the most important are the operating and maintenance procedures, which are common for all industrial equipment. Therefore, there is no special, different recommendation for drying equipment, although each type of equipment has its own specific characteristics, especially in terms of operating procedures. Housekeeping cannot be overemphasized either.

48.4.1 PLANT OPERATION

All employed dryers must be equipped with clear and comprehensive operating manuals prepared jointly by the equipment supplier and the plant operating management. The manual should be learned by the operators, who are to be trained in recognition of hazardous situations and risk assessment with the material being dried. Very useful techniques, known as Hazop (hazard and operability studies) and Hazan (hazard analysis), may be applied successfully in both new and existing processing plants [39,40].

The operating manual should contain the following:

- Prestart-up checklist
- Start-up procedure
- Normal running
- Normal shutdown
- Emergency shutdown

A special hazard is to be expected during start-up and shutdown when the process is not in equilibrium

TABLE 48.11
Summary of the Probable Hazards and the Recommended Safety Measures for Some Dryer Types

	Spray	Pneumatic	Fluid bed	Rotary
Dust cloud ignition	Rotary atomizer Yes Nozzle atomizer Yes (countercurrent)	Mostly in the upper section and in the dust recovery section	In the bed, no; just above bed and in dust recovery section, probably yes	Within the dryer due to fines generated as a result of attrition and breakage
Dust layer ignition	Always a probability; hammers or air blankets are used to prevent	Top bend of dryer is probable location; sometimes agglomerates settle at bottom of tube	In corners if bed is not properly fluidized; large agglomerates may not fluidize at all	Depending upon material characteristics
Bulk ignition	At bottom if discharge mechanism fails	Base of the cyclone; if product is discharged in the silo	On failure of the discharge mechanism; if hot powder goes to a silo	If drum stops rotating discharge mechanism fails; hot powder to a silo
Flammable vapor	Whole system	Whole system	Anywhere in the system	Whole system
Ignition sources	All those listed in Section 3.3	All those listed in Section 3.3; static electricity is more likely	All those given in Section 3.3; static is more likely	All those given in Section 3.3; friction is likely to be more at the seals
Safety measures	Specially for Flammable vapor and solid	For flammable vapors, it is the most feasible	Most practical for flammable vapor	Recommended for flammable solvent recovery
Inerting				
Venting		Top and bottom section of the vertical tube	On the roof for continuous duty and side vent for batch	Inlet and discharge hoods (vent area = dryer cross-section)
	At the roof or side doors (weight/area < 40 kg/m ²)			
Suppressing		Not useful	Suitability depends upon airflow rate and speed of injection	Use limited to small-length dryers
	Useful only for dryers with volume < 100 m ³			

Containing Process specification		Feasible Air temperature 50°C below MIT	Only for small units Bed temperature 20°C below MIT	Feasible Inlet air temperature 50°C below MIT
	Only for small units Inlet air temperature 50°C below minimum ignition temperature			
Dust cloud ignition	No, due to low air velocity except at discharge end	No, due to low air velocity; probable only during emptying trays	Weak explosions during drying and unloading are probable	Mostly depends upon type of material
Dust layer ignition	Material on band and that falls off edges of band	Product in the tray; product spills during charging and emptying	Not much of a change	On shaft and blades of mixer; on inner surface of trough
Bulk ignition	Discharge system fails; hot powder to a silo	Product directly bagged	If mixing stops; during discharging	If shaft stops rotating; if product goes to a silo directly
Flammable vapor	Normally not used	Whole system	Loading and unloading; during drying in the vacuum system	Whole system
Ignition sources	All those given in Section 3.3	All those given in Section 3.3; self- ignition is more likely	All those given in Section 3.3	All those given in Section 3.3; heating and frictional sparks
Safety measures	Not feasible	Feasible for flammable solvents	Can be done	Preferred for flammable vapor
Inerting	Required only when burners in the oven; 6 m apart on the roof	Feasible and most preferred	Required on the roof for cases of exothermic decomposition involving large volumes of gas	On the roof
Venting				
Suppressing	Can be used	Not feasible	Can be used	Only when large free space between mixer and roof
Containing Process specification	Not feasible Air temperature 20°C below MIT	Not feasible Oven temperature 50°C below layer ignition temperature	Easier to attain in cylindrical unit Operating temperature 50°C below bulk ignition temperature	Most suitable 50°C margin from cross-flow layer ignition temperature

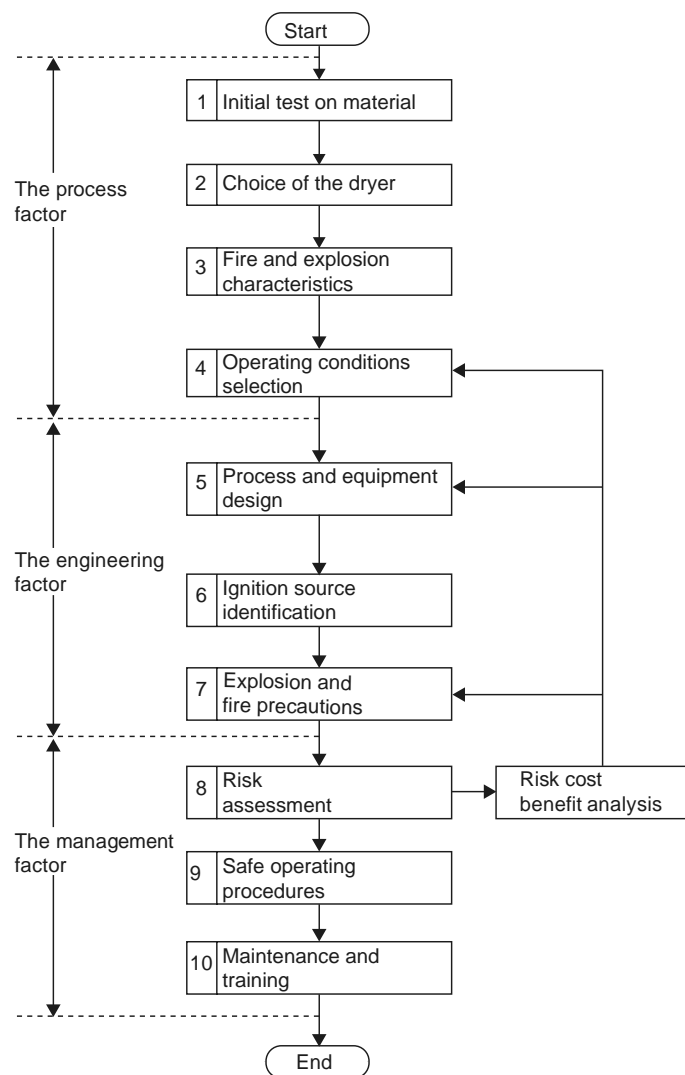


FIGURE 48.18 The procedure in safe plant dryer design.

and it is more likely that control may be lost. Very essential is online controlling equipment, especially for air and material temperatures and flow rates. For evaporation of organic solvents, vapor concentration monitoring is mandatory. Any instrument malfunctioning, dust leakage, or spillage, and any deviation from an acceptable range of parameters must be located and removed. Regular safety reviews should be carried out, particularly when any changes are to be made in equipment, feedstock, or operating practice.

A full account of the plant operation safety requirements can be found in Reference 8.

48.4.2 MAINTENANCE AND TRAINING

Fire and explosion hazard in dryers imposes extra requirements on plant maintenance. It is important to apply the general policy in a breakdown case or

prevention measures as well as routine maintenance and turnaround periods. Special attention should be paid to maintenance in all safety systems and in that equipment responsible for the ignition sources. A work permit should be followed in all maintenance work, especially performing hot work like welding, cutting, or soldering.

It is important to appreciate that the main danger is not necessarily from a primary explosion occurring within the drying plant. The secondary explosion, usually caused by dust deposition on surrounding surfaces, is always devastating; the importance of good housekeeping cannot be overemphasized.

Adequate training of management and operatives is necessary to run the plant safely, efficiently, and economically. The plant should allow not only for preparation of correct operating procedures, but also for staff action in emergency conditions.

48.5 THE PROCEDURE FOR SAFE PLANT DRYER DESIGN

On the basis of process, engineering, and management factors, a procedure of safe plant design can be proposed. This procedure may be divided into 10 stages (Figure 48.18) and can be used as a general guide with possible modifications for special conditions. The selection of the dryer type is based on the economics of running the dryer. The aspect of dust fire and explosion is not a restriction for the selection. However, if options are open, selection should be based on the explosive properties of the dust and the level of the fire and explosion hazard (the type of hazard, the volume of the dust–air mixture, quantity of deposit within the plant, and the design of the heat supply). It also should be noted that it is more economical to provide necessary safety measures against a hazard at the design stage so that safety can be built into the plant, rather than to “retrofit” safety equipment, which can be very costly.

If any change happens in the operation of the plant dryer, feedstock, or equipment, it is important to take it into account and consider the plant design process from the new starting point. Changes in the feedstock formula may have a substantial effect on its explosive properties and minor engineering changes may provide a greater accumulation of dust, which may eventually ignite, within the plant.

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49 Control of Industrial Dryers

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49.1 INTRODUCTION

Today, most industrial dryers are equipped with varying levels of automatic controllers. Often they use simple control strategies based, for example, simply on the exhaust-gas temperature for a direct dryer. Small-scale and slow drying operations are often controlled (or adjusted for process upsets) manually. Very high production units, those involving very rapid drying or units that produce products within stringent quality specifications, must be equipped with some degree of automatic control. Although commercial dryers currently use conventional control strategies, it is expected that within the next decade more and more industrial dryers will utilize model-based control (MBC), fuzzy logic control (FLC), or neural nets con-

trol when the dryer performance is highly nonlinear and difficult to predict with simple mathematical models. Some improvements in dryer controls became available because of the development of better sensors and analyzers, whereas others are by-products of new, more sophisticated, computer-based control techniques [1]. This chapter provides an introductory overview of both the conventional and the emerging control schemes for industrial drying. Examples are cited with reference to the more common dryers (e.g., spray, flash, fluid-bed dryers). Relevant information is also provided to the readers interested in intelligent control systems based on expert systems, fuzzy logic, or neural nets. It is inconceivable that within this decade equipment suppliers will market “smart” dryers that can adjust their operating parameters consistent

with the needs of product quality during drying. However, such a possibility exists for some dryer types in a longer term.

49.2 SCOPE

The aim of any industrial drying process is to produce a solid product of desired quality at minimum cost and maximum throughput and to maintain these consistently. Good quality implies that the product corresponds to a number of technical, chemical, and biological parameters, each within specified limits.

Thermal drying is an energy-intensive operation that accounts for up to 15% of all industrial energy usage. Moreover, conventional dryers often operate at low thermal efficiency, typically between 25 and 50% but which may be as low as 10% also. Application of automatic control technology to industrial dryers offers an opportunity to improve the dryer operation and its efficiency. Furthermore, the wide variety of dried products that are available to the consumer increases the concern to meet high-quality specifications. Therefore, the need for optimal-energy management, with the demand for high-quality products as well as the adoption of more strict safety and environmental regulations, catalyze the development of more advanced dryer-control strategies.

The control of dryers is probably one of the least-studied areas of process control and has not progressed concurrently with improvements in drying and dryer design. This may be attributed to various factors [1–3], including

- The lack of emphasis on product quality in the past
- An apparent lack of knowledge of the important role that the dryer control plays in product quality and drying efficiency
- The lack of a direct, online, reliable method for sensing product moisture content
- The complex and the highly nonlinear dynamics of drying processes, leading to difficulties in modeling the process adequately

The basic objectives of a dryer control system are:

- Maintenance of desired dried-product quality, irrespective of disturbances in the drying operation and variations in feed supply
- Maximization of throughput at optimal energy efficiency and minimum cost.
- Avoidance of overdrying and underdrying; underdrying may result in spoilage, in the case of grains and foodstuffs, whereas overdrying of

the product results in increased energy costs and reduced yields as the price of some products is based on a specific moisture content; it may also cause thermal damage to heat-sensitive products

- Reduction of fire hazard, defective product, and particle emission
- Suppression of the influence of external disturbances
- Stable drying process
- Optimization of the performance of the drying process

It is worth pointing out that drying is, in principle, an inherently self-regulating process and there is no need for automatic control as long as there are no input fluctuations and the mass balance and the process conditions are invariant. However, few real drying processes ever approach this ideal and changes do occur, justifying the need for regulation and control.

As the dryer represents a very complicated plant for automation with a great number of parameters affecting product quality, specifying a control system for the dryer requires the consideration of many process factors. These considerations include the process dynamics; the number of process variables that are to be controlled or monitored; product handling; operating ranges of temperature, pressure, humidity; air and product flows; initial and final moisture contents; and so on. Additional factors include the protection of critical process parameters with the use of interlocks; data acquisition systems; ease of calibration of control and sensing devices; and ease of maintenance and operating reliability. Overall reliability and cost of the control system are clearly important as well. If the process is inherently subject to fire or explosion hazards, then special considerations must apply; however, these are beyond the scope of this chapter.

49.3 DRYER CONTROL SYSTEM CHARACTERISTICS

The requirements and characteristics of any industrial dryer control system are [4]:

1. *Accuracy.* The exit product moisture content must be close to the desired value.
2. *Stability.* The system must not oscillate; otherwise, large fluctuations in output moisture content would occur.
3. *Speed of response.* Any disturbances (e.g., changes in input moisture content) should be quickly offset by the controller in order to provide acceptable upset recovery time and system stability.

4. *Robustness*. The control system should be able to operate successfully over a wide range of process conditions.

49.4 DRYER CONTROL SYSTEM SYNTHESIS

Prior to control system design, control synthesis must be performed. The synthesis of control configuration for a multivariable system involves the selection of controlled and manipulated variables, steady-state analysis, dynamic analysis, loop pairing (pairing manipulated inputs and controlled outputs), and the selection of best control configuration [5–7]. A good selection of the control scheme depends on studying the behavior of the system under different circumstances. This can be achieved by making some deterministic tests either in a real plant or in a physical model that properly and adequately fits the actual process.

In the language of process control, the variables associated with an industrial drying process can be divided into two general groups:

1. Input variables, which denote the effect of the surroundings on the drying process
2. Output or controlled variables, which denote the effect of the drying process on the surroundings

The input variables can be further classified into manipulated variables and disturbances or load variables.

Manipulated variables can be adjusted either manually or automatically. The most important manipulated inputs to a dryer are (depending on dryer type):

1. Heating rate (e.g., inlet air or steam temperature)
2. Solids feed rate (e.g., screw conveyor speed, discharge rate)
3. Airflow rate (for direct dryers)
4. Rotational speed (for rotary dryers)

Disturbances or load variables cannot be adjusted by a control system. The most common dryer disturbances are:

1. Ambient air temperature
2. Ambient air humidity
3. Feed moisture content
4. Feed composition

The output or controlled variables may be classified as measured output variables and unmeasured (or difficult-to-measure) output variables. Dryer output variables are:

1. Dried-product moisture content
2. Exhaust air temperature
3. The temperature of the air–product mixture (for spray dryers)
4. Exhaust air humidity
5. Product quality (color, flavor, textures, activity, etc.); these properties are generally difficult (or impossible) to measure online and the cost is often too high

In general, it is difficult to make online measurement of product quality; it is often an inferred entity based on experience. The moisture content can be a measured variable if a suitable moisture sensor is used in the process. Analyzers are available for moisture analysis, which can be adapted for automatic, closed-loop dryer control. The most successful of these units rely on infrared, microwave, or capacitance detection [1]. Care must be exercised in the selection to allow for changes in bulk density or for void spaces, which will introduce an error. In many industrial-drying plants, such sensors are not available because either they are expensive or they have low reliability. Consequently, in such cases the moisture content is an unmeasured output variable.

The relationship between output, manipulated, and load variables constitutes the process control system of the dryer (see Figure 49.1). In particular, the purpose of a dryer control system is to produce a desired output by changing the manipulated variables so as to compensate for changes in the main load variables (disturbances). Inputs are in the form of commands, which the output is expected to follow, and disturbances, which the automatic control is expected to minimize.

The most desirable drying process output variable to control is the product moisture content, but this is often difficult to measure directly. Often, the moisture content of the dried product can be inferred from the temperature and humidity of the exhaust gas, though care must be taken in applying appropriate heat and mass balances. Moreover, because measuring the temperature of the exit gas is simple, accurate, reliable, cheap, and has significant effects on the drying rate, dryer manufacturers have used this variable as the control output variable; and, in fact, the majority of existing dryers rely on the automatic control of this variable. However, due to the low correlation between the temperature and the actual product moisture content, using indirect control (by parameters such as temperature and humidity) usually results in poor control of the drying process.

In some special cases, this may even lead to serious failure if the exhaust temperature is used in isolation (without concurrent measurement of humidity). For

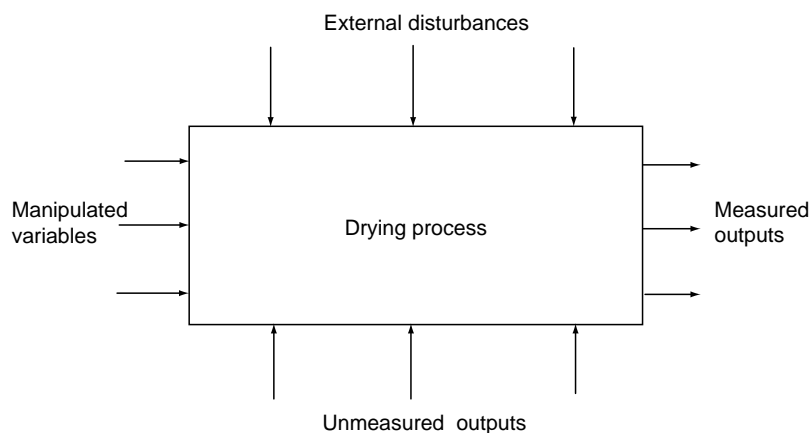


FIGURE 49.1 Schematic diagram of drying-process variables.

example, a higher exhaust-gas temperature will be interpreted by the controller as increased available dryer capacity, leading to an increase in feed rate. However, the elevated temperature can result from poor dryer performance (due to deposits, etc.). Thus the controller action can worsen the problem.

Direct control and online measurement of the solids moisture content would enable significant improvements in dryer control by providing an immediate measure of the moisture content at the dryer exit and by automatic compensation for factors that disturb the control action. In certain dryers, these are already implemented industrially (e.g., in paper dryers).

49.5 CONVENTIONAL AND ADVANCED DRYER CONTROL SYSTEMS

49.5.1 MANUAL CONTROL

In manual control, expert judgment by the operator is relied upon to judge the endpoint of the drying process. A dryer manual control scheme may be described by the following sequence [2,4]:

1. Turn on the dryer
2. Set the initial throughput
3. Measure the output moisture content and compare measurement with the desired value
4. Based on the difference between the desired and the measured moisture content value, make adjustments to the manipulated variables (e.g., energy input, feed rate) to maintain the desired moisture content

This type of manual control is simpler and less expensive, and less expertise is required than automatic control systems. It can be applied to small plants (mainly batch systems) and on easy-to-dry-

materials. However, manual control is not recommended in the case of large drying processes and where good control is required to stabilize the process against any disturbances. It is also labor-intensive.

49.5.2 FEEDBACK CONTROL

The principle of feedback control is one of the most commonly used control strategies in dryer control. The major function of a dryer feedback controller is to hold the controlled variable at a target value or set point. The control system receives a measured signal of the controlled output variable (i.e., moisture content) and compares it with the set point value, which generates an error signal. The value of the error is supplied to the main controller. The controller in turn changes the value of the manipulated variable in such a way to reduce the magnitude of the error. Usually, the controller does not affect the manipulated variable directly but through another device, known as the final control element, such as a control valve, motor, fan, or heater, depending on the application. Ideally, control will result in an exact correction in the process output variable, forcing it back to the desired value (set point). A typical feedback control loop is shown in [Figure 49.2](#).

Three basic types of feedback controller actions are available: (1) proportional, (2) integral, and (3) derivative action. Proportional action actuates the manipulated variable in direct proportion to the error signal. Integral action eliminates any steady-state residual errors or offsets and it moves the manipulated variable based on the time integral of the error. The purpose of the derivative action is to forecast fast changes in the error signal by using a control mode proportional to the time rate of change of the error signal.

In industrial dryer control applications, the three control actions described above can be used individually or in combined modes: proportional (P)

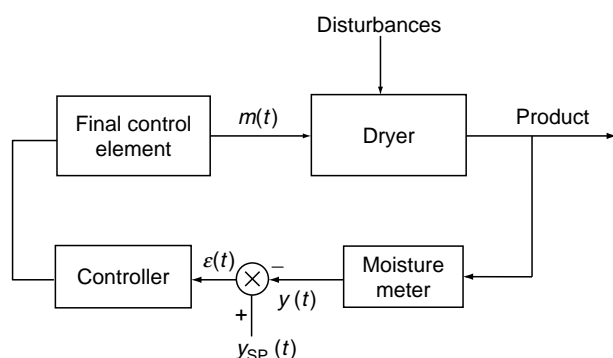


FIGURE 49.2 General structure of dryer feedback control system.

controller, proportional–integral (PI) controller, or proportional–integral–derivative (PID) controller. However, the derivative action is generally undesirable because of the problem of noisy signals [8,9].

The general three-action (PID) controller can be described by the following equation:

$$m(t) = K_c \varepsilon(t) + \frac{K_c}{\tau_I} \int \varepsilon(t) dt + K_c \tau_D \frac{d\varepsilon}{dt} + \text{BIAS} \quad (49.1)$$

where

m is the control action, K_c is the proportional gain of the controller, ε is the error signal, τ_I is the integral time constant, τ_D is the derivative time constant, and **BIAS** is the controller's bias signal (i.e., its actuating signal when $\varepsilon = 0$).

Typical feedback controllers in drying applications have been investigated and reviewed by Robinson [2], Marchant [4], and Whitfield [10].

49.5.3 FEEDFORWARD CONTROL

The residence time of the solid material may be relatively long in certain types of industrial dryers (e.g., rotary dryers, floater dryers for pulp sheets). Also, in certain dryers the thermal inertia of the heat-transfer mode may be long relative to the drying time (e.g., for drying of newsprint or contact dryers or Yankee dryers for tissue paper). In such cases, there is a significant time lag between a change being made to an input and its effect being felt on the output. If this dead time is large, it may result in inadequate performance of the feedback controller. To overcome this problem, a predictive type of control known as feedforward control is used.

In this control system, process disturbances are measured and compensation is made for them without waiting for a change in the output variable to indicate that a disturbance has occurred. The control scheme is implemented by measuring the disturbance

or load input (e.g., inlet moisture content) and the controller uses a system model to determine the relationship between the disturbances and the manipulated and controlled variables. The objective here is to keep the value of the controlled output variable at the desired level by eliminating the impact that the disturbances would have on the output. It is clear that the effectiveness of the feedforward control system depends on the accuracy of the system's mathematical model in predicting the response of the process to input and disturbance changes. The principle of feedforward control is illustrated in Figure 49.3.

49.5.4 FEEDFORWARD–FEEDBACK CONTROL

From the earlier discussion and from Table 49.1 [11], we would expect that a combined feedforward–feedback control system would retain the superior performance of the feedforward controller and the insensitivity of the feedback controller to uncertainties and inaccuracies. Figure 49.4 shows the main components of a feedforward–feedback control system; they include the feedforward controller incorporated with the process model, the feedback controller, and the dynamic compensator. The function of feedback element is to correct the action of the feedforward controller in the case of measurement and modeling inaccuracy. Courtois et al. [12] and Bruce and McFarlane [13] implemented the combined feedforward–feedback control system for mixed-flow grain dryers.

49.5.5 MODEL-BASED CONTROL

MBC has recently become a process-control approach of much interest. The basic idea is to use a dynamic model of the process in the control system. A comprehensive exposition of MBC is beyond the scope of this chapter; interested readers are

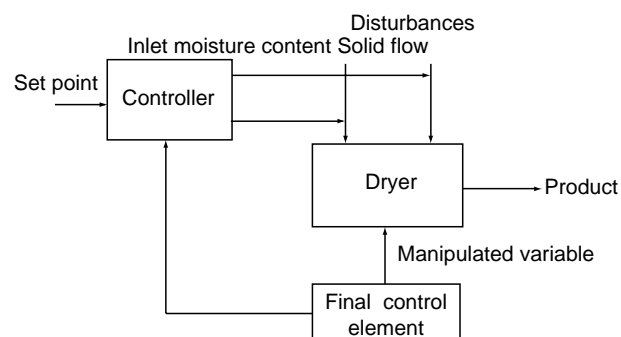


FIGURE 49.3 Dryer feedforward control system configuration.

TABLE 49.1
Relative Advantages and Disadvantages of Feedforward and Feedback Controllers [11]

Advantages

Disadvantages

Feedforward Controller

1. Acts before the effect of the disturbance has been felt by the system
2. Good for slow systems or those with significant dead time
3. Does not introduce instability in the closed-loop response

1. Requires identification of all possible disturbances and their direct measurement
2. Cannot cope with unmeasured disturbances
3. Sensitive to process parameter variations
4. Requires a process model

Feedback Controller

1. Does not require identification and measurement of any disturbance
2. Insensitive to modeling uncertainties
3. Insensitive to parameter changes

1. Waits until the effect of the disturbances is felt by the system before control action is taken
2. Unsatisfactory for slow processes or those with significant dead time
3. May create instability in closed-loop response

encouraged to read other references [8,14–17]. Some of the MBC strategies are:

1. Inferential control
2. Internal model control (IMC)
3. Dynamic matrix control (DMC)

Inferential control [11,17,18] is an early model-based approach for process control. This control strategy is useful when the main dryer controlled variable (i.e., product moisture content) cannot be measured directly due to some technical difficulties or due to insufficient economic justification for its measurement. Inferential control uses the values of measured outputs (e.g., product or gas temperature and humidity) together with the process model to

infer the value of the unmeasured control variable. These estimates are used to adjust the values of the manipulated variables in order to keep the moisture content at the desired levels (Figure 49.5). This control policy can also be used to counteract the disturbances as it is less expensive to infer these disturbances from other available process measurements rather than by measuring them directly.

The basic idea of IMC is to use a process model and to relate the controller settings to the model parameters in such a way that the selection of the specified closed-loop response yields a physically realizable feed-back controller [8,17]. IMC is advantageous because it can be adjusted to balance controller performance with control system robustness (when either modeling errors or changes in process dynamics occur). Clearly,

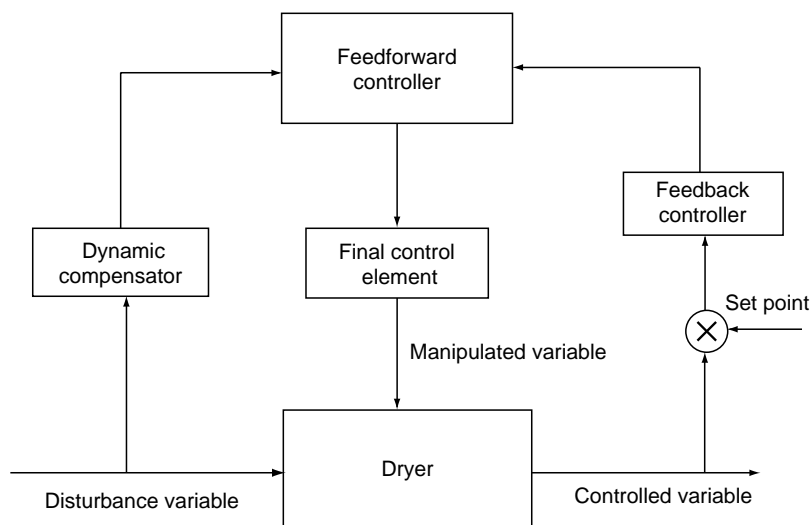


FIGURE 49.4 Feedforward–feedback control system.

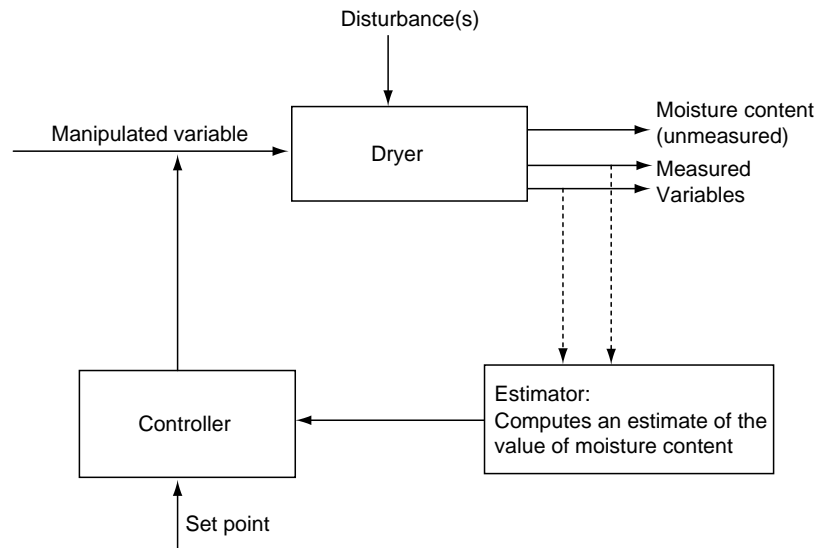


FIGURE 49.5 General structure of a dryer inferential control system.

the effectiveness of IMC depends on the availability of a reliable model for the dryer.

Panda investigated the performance of IMC in fluid-bed drying of sand particles, mustard seeds, and wheat grains [19]. The structure of the IMC system for the fluid-bed dryer is depicted in the block diagram shown in Figure 49.6. In this study, IMC uses a process-model transfer function (G_m) parallel to the actual plant transfer function (G_p). A filter is used in the control system to ensure robustness in performance. The exit-air temperature is used for set-point tracking by the IMC. If the system is performed without any oscillations, the overshoots will be tolerable, there will be no offset, and the control scheme will be effective and respond rapidly as described by Panda [19].

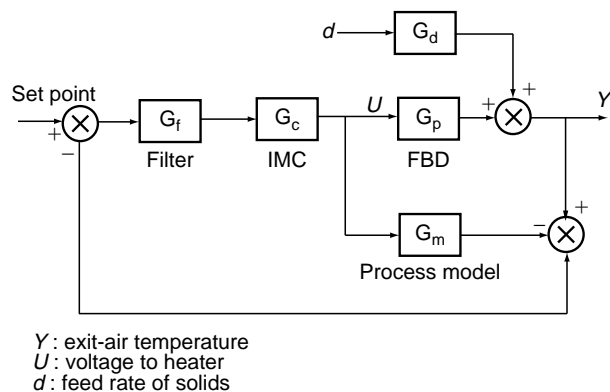


FIGURE 49.6 Basic structure of internal model control for a fluid-bed dryer.

DMC uses a direct, least-squares computational procedure to find the optimum values of future changes in the manipulated variables to match future output responses such that some performance index is minimized [8]. Panda applied DMC to control the fluid-bed dryer as mentioned earlier and proved that DMC gives better control in the presence of measurement noise [19]. The response in the exit-air temperature does not show much oscillation and settles quickly.

Al-Haj Ali et al. [5,6] developed different types of linear time invariant models by system identification, which adequately represent the fluidized-bed drying dynamics. MBC techniques such as IMC and model predictive control (MPC) were used for the designing of the control system. Simulations with multivariable MPC strategy provided robust, fast, stable, and non-oscillatory closed loop responses. A stationary form of Kalman filter was designed to estimate the particle moisture content (state observer). Performance studies showed that the Kalman filter provided satisfactory estimates even in the presence of significant noise levels and inaccurate initial states feed to the observer.

Kiranoudis et al. [20] developed a dynamic model for the simulation of conveyor-belt dryers and proposed a SISO (single-input, single output) control scheme for the regulation of material moisture content. In a subsequent work, Kiranoudis et al. [21] extended the dynamic model of this process to include MIMO (multiple-input, multiple output) scheme to control the material moisture content and temperature. In both works, PI controllers were appropriately tuned and nonlinear simulations were performed.

Trelea et al. [22] designed and tested a nonlinear, predictive, optimal control algorithm for a batch corn dryer. The control algorithm relies on the solution of a realistic, constrained, optimization problem and experiments showed that the algorithm was able to handle important disturbances and failures. Vasconcelos and Filho [23] developed a supervisory control strategy for the optimal operation of grain dryers. In this study, an optimization problem was developed to function as a supervisory control and predictive DMC algorithm was used for servo or regulatory control. The proposed algorithm presented satisfactory results for the load rejection and set-point variation.

Corrêa et al. [24] conducted control tests in a spouted bed dryer of paste using a generalized predictive control (GPC) algorithm. The GPC algorithm is based on the minimization of deviations between process variables and nominal (or reference) model variables using the recursive least-square method. The controller showed a stable behavior and good performance for both set point tracking and disturbance-rejection experiments. Didriksen [25] showed that the performance of a model-based controller was superior when compared with traditional feedback control in a simulation study based on industrial rotary dryer data. An augmented Kalman filter was used to estimate the disturbances. MBC had also been used by Altafini and Furini [26], Harn et al. [27], and Thyagarajan et al. [28].

49.5.6 MICROPROCESSOR-BASED CONTROL

Traditionally, analog instrumentation was used, and is still used for process control in some cases. However, the advent of the microprocessor in the early 1970s and the development of digital computers, coupled with significant reduction in their cost, have brought about the evolution of computer-based control systems and controllers capable of providing more than basic control.

Although the primary task of a computer-based controller is the implementation of a control algorithm (PID or more sophisticated algorithms), the presence of a computer makes it possible to achieve more than basic control and to assign a number of tasks that are useful in process control. The following characteristics of modern computer control system illustrate some of these tasks [8,11,18,29,30]:

1. Implement classical and advanced control algorithms
2. A single digital computer (or microprocessor) to service a number of control loops (time-shared basis)
3. Distribute data processing by which data can be collected from different process instruments

and processed for monitoring and control purposes; the computer system can also be connected to local analytical instruments (e.g., moisture and humidity meters), which usually have their own microcomputer

4. Provide static and dynamic displays on monitors or other visual display units
5. Provide mathematical functions
6. Provide data acquisition and storage for different process measurements, such as temperature, flow rate, pressure, humidity, and moisture content
7. Provide planning, supervision, optimization, quality control, and control of mode of operation

Figure 49.7 shows the structure of a microprocessor-based control system. The computer system collects data from the process measurements and calculates the values of the manipulated variable and implements the control action on the process, based on the control algorithm that is already programmed and stored in the memory of the computer. Signals are converted by digital to analog (D/A) and analog to digital (A/D) converters. The operator communicates with the control system with a keyboard, a monitor, and a printer or plotter.

Panda investigated a direct digital control (DDC) feedback loop to control a continuous fluid-bed dryer [19]. A PI control algorithm tuned with Cohen–Coon tuning rule gave satisfactory performance of the drying of sand, mustard, and wheat for different load variables, such as solid feed rate. A microprocessor-based control system for a column-type grain dryer had been developed by Jaaksoo et al. [31]. Input–output variables of the plant were specified and a two-input/two-output state variable model was obtained using experimental input–output data. The proposed control algorithm had proven to be reasonably suitable to realize a robust regulator for a grain dryer.

Sundaramoorthy and Rao designed and implemented a DDC scheme on a batch fluid-bed dryer, which was drying wet sawdust [32]. The system was described by a state-space model with parameters that were estimated from experimental data. The performance of the designed controller was checked by closed-loop simulation and then the scheme was implemented online using the heater power supply as the final control element to regulate the inlet-air temperature.

Barker and Christie described a microprocessor-based control system for the control of textile drying processes [33]. In addition to the control of the dryer outlet moisture content, the computer-based system was used to optimize the energy use of the dryer.

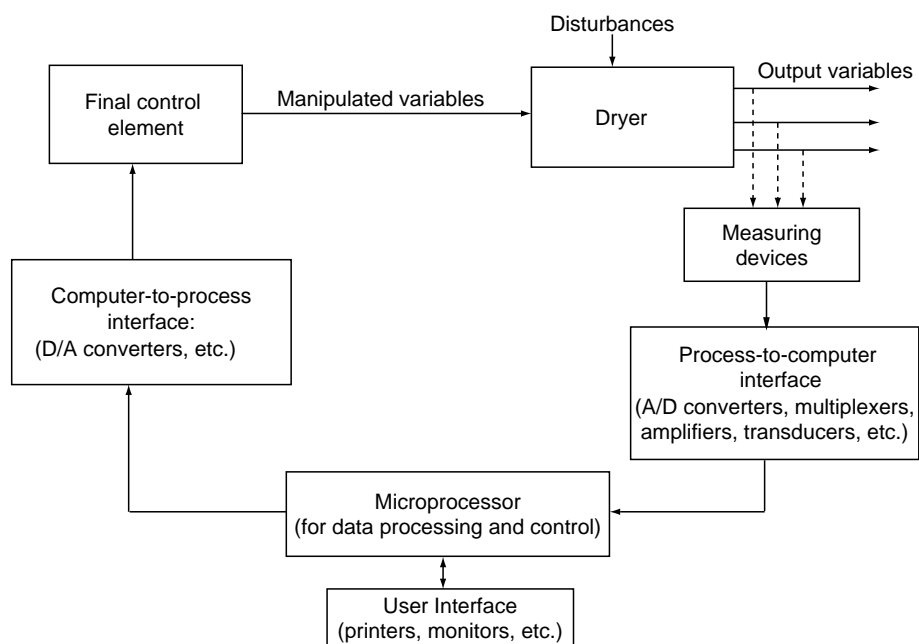


FIGURE 49.7 Microprocessor-based dryer control system.

Douglas et al. presented a computer-based control system of particulate dryers, including the required sensors, a centralized computer system, and the software [34]. The work included the development of a capacitance-type moisture meter. A continuous horizontal conveyor-belt dryer for drying pet food had been used as an illustration.

Microprocessor-based control system had also been implemented in a solar-tunnel dryer [35] and a pneumatic drum dryer [36].

49.6 TYPICAL INDUSTRIAL DRYER CONTROL SYSTEMS

In this section, we overview some control strategies for commonly used industrial dryers. It can be noted that different strategies are possible and, indeed, actually used in practice.

49.6.1 BATCH DRYERS

Batch drying is a time-dependent and repeatable process usually used for the drying of small volumes, sensitive products, or valuable materials (e.g., pharmaceutical products). Manual control of batch dryers requires higher labor costs per unit product throughput than continuous processes, necessitating the need of automatic control. The major benefits of automatic control of batch-drying processes are increased safety, increased production through a reduction in cycle time, and increased consistency of product quality. In

some special cases, preprogrammed temperature or airflow (or both) can be implemented in the interest of enhanced quality or energy savings.

Typical automatic batch-dryer control systems use the exhaust-air temperature as the controlled variable to determine when to end the drying process. Shinskey [37] and Fadum and Shinskey [38] have described an alternative control scheme employing inferential control. The system has proven to be an effective substitute for online moisture analysis in terminating drying. This method is based on the following equation:

$$T_{of} = T_{oc} + K(T_i - T_{oc}) \quad (49.2)$$

where T_{of} is the final outlet-air temperature corresponding to the desired moisture content, T_i is the inlet-air temperature, and T_{oc} is the constant outlet-air temperature. K is a factor that depends on particle size and drying rate. The control strategy is to shut down the process when the outlet temperature reaches the level defined in Equation 49.2. The moisture control scheme may be refined to include a dew-point loop to save energy during the falling drying-period rate.

Robinson described a control system based on a temperature-drop model to determine and control the product moisture content at any appropriate point inside batch dryers [2]. It is based on the model

$$M = k_1(dT)^q - k_2(D_t)^r \quad (49.3)$$

which relates product moisture content M to the temperature drop dT , which is defined as the difference between the temperature of the entering drying air and the temperature of the exhaust air. D_t is the drying time and k values and exponents are constant for a given product and dryer.

The temperature-drop method enables the calculation of drying rates and rate of change of drying rates for the use as controlled variables. As suggested by Robinson [2], such variables significantly improve the control of endpoint moisture content because they are independent of the dryer and process variables.

49.6.2 ROTARY DRYERS

Convective rotary dryers are normally controlled by the measurement of exit-gas temperature, which is used to regulate the inlet-air temperature; the pressure is adjusted by the exhaust damper setting, and the inlet-air rate is used to control the outlet-air wet-bulb temperature [39,40]. An alternative control scheme uses the air rate to control the difference between the dry- and wet-bulb exhaust temperatures; these temperatures serve as an indication of the rate of evaporation [39]. This modified control system is shown in Figure 49.8.

Douglas et al. have developed a control strategy to control the outlet moisture content and temperature by manipulating the airflow rate and rotational speed of the dryer, respectively [41]. A PI controller is used to achieve the control objectives. In steam-tube rotary dryers, the system is controlled conventionally by regulating the steam-tube pressure. This control

system produces acceptable results for a constant dryer load; but as load changes, underdrying and overdrying may occur. To overcome this problem, Fadum and Shinskey developed a compensated control system based on the fact that the relationship between pressure and steam flow must be maintained in order to control product moisture content [38]. They suggested that this relationship could be reduced to a linear approximation,

$$P_o = Rh + b \quad (49.4)$$

where P_o is the tube pressure set point, R is the ratio setting for desired moisture, b is the bias term, and h is the steam orifice differential pressure.

49.6.3 SPRAY DRYERS

The residence time in spray dryers is short and the small droplets offer a large surface exposed to heat and mass transfer, hence drying is rapid. Rapid drying leads to a short process dead time, allowing good automatic control. Usually, one of the control objectives is to keep the size of the droplets uniform. Three basic types of control systems are commonly used in spray-drying installations [42].

Control system A (Figure 49.9) features two quick-response control loops:

1. Control of exhaust-air temperature by feed-rate regulation
2. Control of inlet-air temperature by air-heater regulation

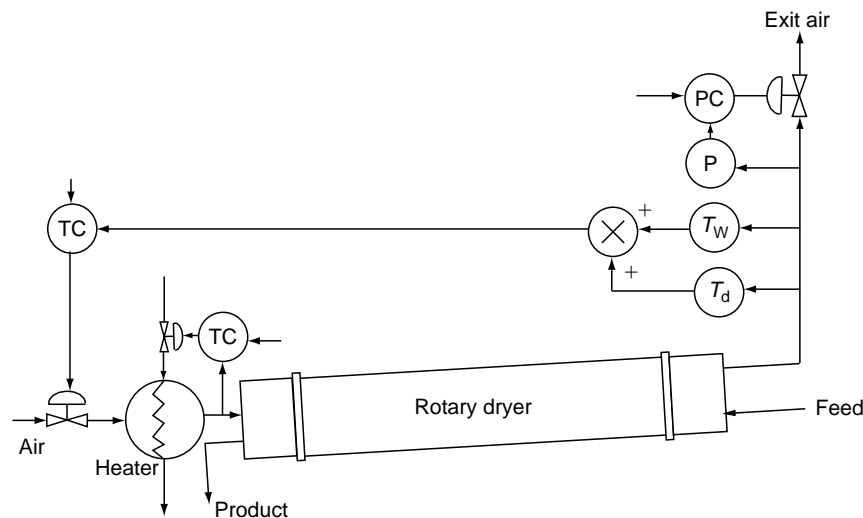


FIGURE 49.8 Rotary dryer control system.

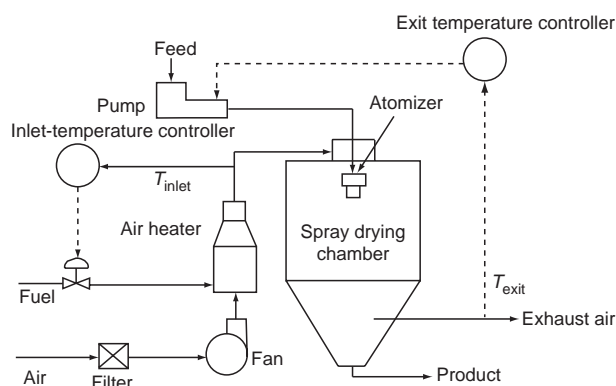


FIGURE 49.9 Control system A for industrial spray dryers.

Control system B (Figure 49.10) uses air-heater regulation to compensate for any deviation from the desired outlet-air temperature. This system also features manual regulation of the feed rate.

Control system C features two alternatives:

1. Outlet air-temperature control with feedback from moisture content measurement (Figure 49.11a)
2. Microprocessor-based control (Figure 49.11b) in which the outlet-air temperature is controlled by regulating the feed rate, with feedback from the moisture content measurement and feed-forward from the atmospheric humidity, feed specific gravity, and inlet air-temperature measurements

Control systems A and C are commonly used for rotary atomizers. Control systems A and B are applicable for nozzle atomizers, but control system B is

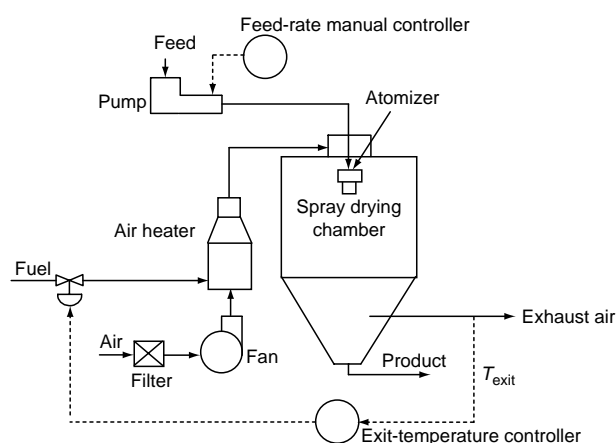


FIGURE 49.10 Control system B for industrial spray dryers.

preferred when wide variations in feed rate cannot be handled.

These control schemes are provided with safety systems that prevent any failure in the feed system in order to prevent the outlet-air temperature from rising above a specified safety level. These safety systems can shut down the air heater or pass water to nozzles positioned as a safety measure when a certain outlet-air temperature is reached [42]. Alarms might be added to detect the potentials for explosion, product plugging, flooding, high temperature, and loss of airflow.

49.6.4 CONVEYOR DRYERS

Various control schemes have been used in conveyor dryer systems. When a steam heater is used to heat up the makeup air, the temperature of the circulating air through the dryer is controlled by adjusting the steam-control valve [40].

Zagorzycki investigated the technical and economic feasibility of humidity control systems applied to conveyor dryers such as tobacco dryers [43,44]. The humidity at any point in the dryer is controlled by the adjustment of the volume of exhaust flow. In this case study, it has been shown that applying control technology to conveyor dryers offers substantial cost savings.

49.6.5 FLASH DRYERS

Traditional flash-dryer control systems include controlling the exhaust-air temperature by varying the inlet drying-air temperature (Figure 49.12). This control system performs adequately in the absence of process disturbances, but poor control and overdrying or underdrying occur in the case of load changes (feed moisture content) because of the short residence times of both solids and gases. Alternatively, the flow rate of wet solids can be controlled with a small response time [40]. If the product is proved to form deposits, special care is needed in interpreting the control action.

Shinskey conducted a thorough study of adiabatic dryers, including flash dryers, and, using material and energy balances, established the following relationship for product moisture content [37,38]:

$$X_p = K \ln \left(\frac{T_i - T_w}{T_o - T_w} \right) \quad (49.5)$$

where X_p is the product moisture content, T_i is the inlet-air temperature, T_o is the outlet-air temperature, T_w is the wet-bulb temperature, and K is the constant for a particular dryer and product.

From Equation 49.5, it is clear that, in order to control the moisture content in the dryer, the ratio of

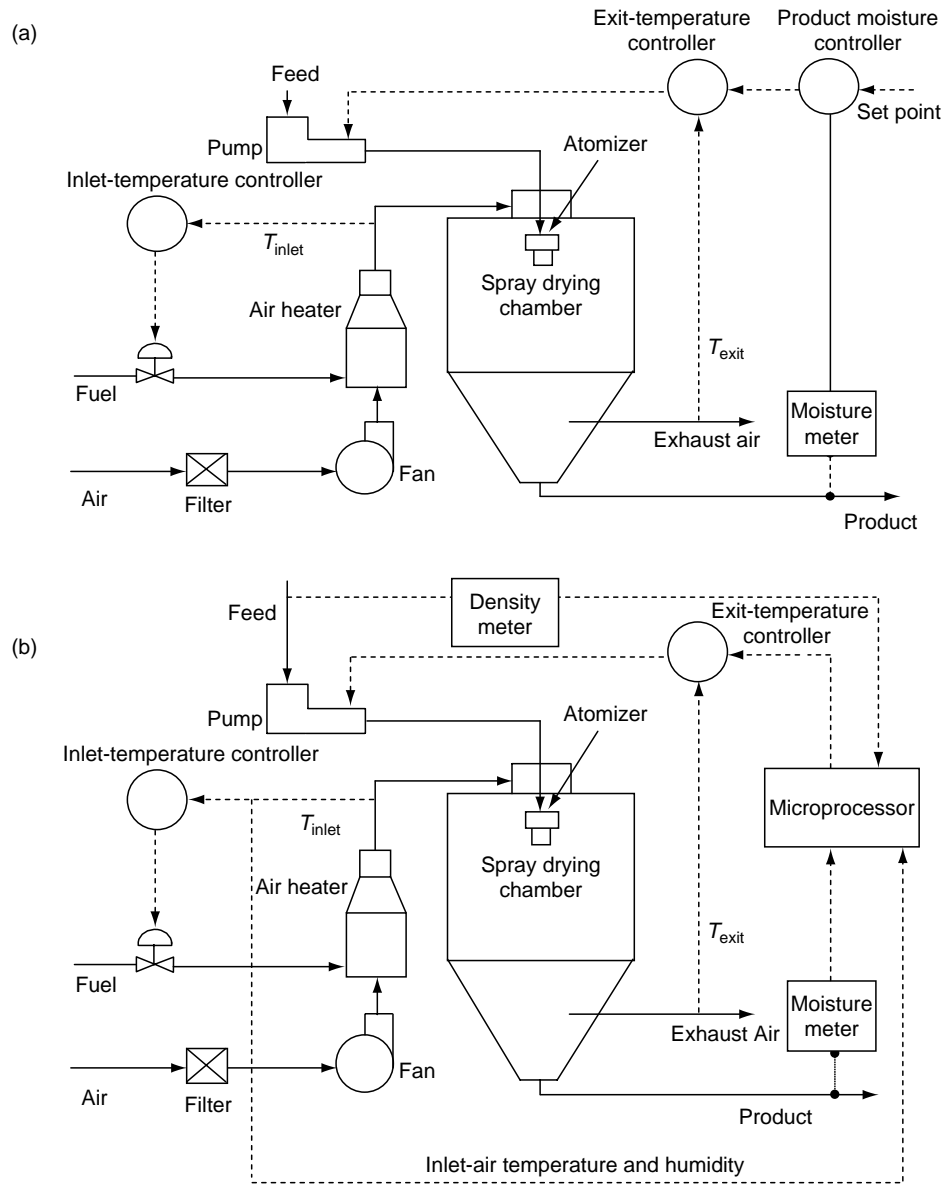


FIGURE 49.11 Control system C for industrial spray dryers: (a) outlet-temperature control with feedback from moisture content measurement; (b) microprocessor-based control system.

the driving force at input and output must be kept constant by varying both input and output temperatures as the load varies. Equation 49.5 can be rearranged to give

$$\frac{T_o - T_w}{T_i - T_w} = K = \text{constant} \quad (49.6)$$

or

$$T_o = KT_i + (1 - K)T_w \quad (49.7)$$

which gives the required output temperature to keep a constant moisture content for a given input temperature and wet-bulb temperature.

However, as it is difficult to measure the wet-bulb temperature in the case of hot air contaminated with solids, a program relating T_o to T_i for a particular ratio of temperature difference $T_o - T_w / T_i - T_w$ has been established. Linearization results in the following equation:

$$T_o^* = RT_i + b \quad (49.8)$$

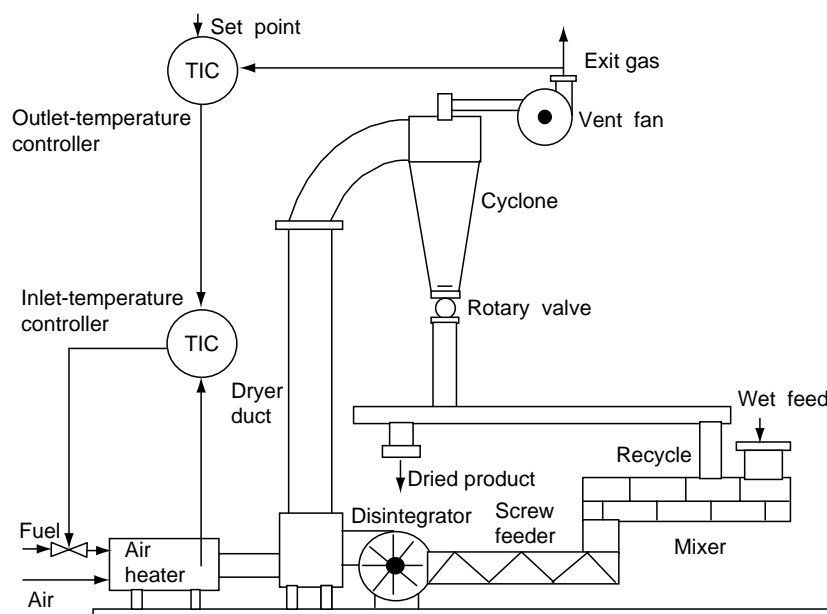


FIGURE 49.12 A conventional flash-dryer control system.

where T_o^* is the desired outlet-air temperature, R is the slope of T_o vs. T_i curve, and b is the T_o axis intercept or bias.

A control system employing this concept is shown in Figure 49.13.

49.6.6 FLUID-BED DRYERS

Among the available control strategies, the control system based on exhaust-air temperature as the con-

trolled variable and inlet-air temperature as the manipulated variable is found to be adequate for most fluid-bed dryers. This control strategy (Figure 49.14) performs quite satisfactorily under any load changes because of the fluid-bed buffer capacity [40]. However, as this control system responds to exhaust temperature but not to the absolute humidity of the airstream, it is limited by the “reverse action” [1]; an increase in humidity of the entering air can cause a reduction in inlet-air temperature instead of

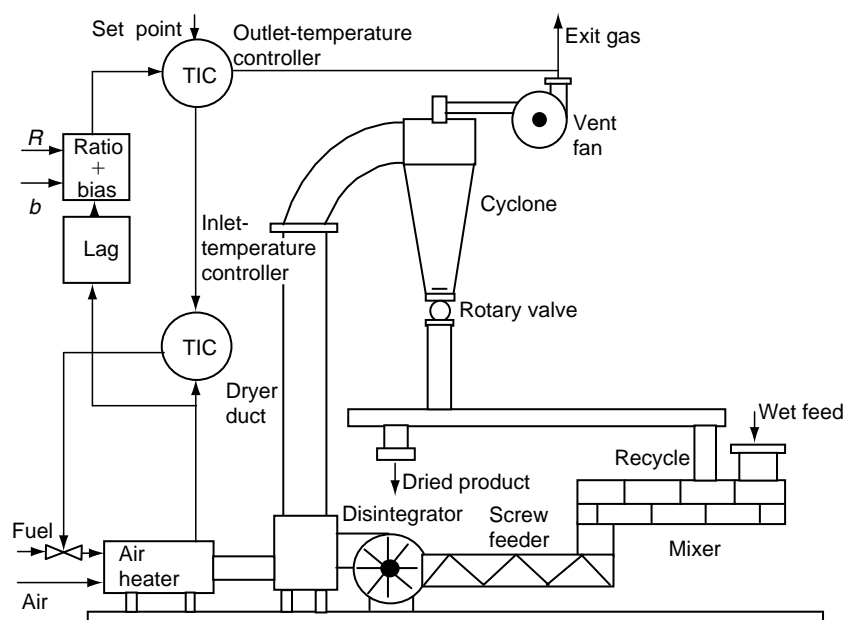


FIGURE 49.13 Compensated flash-dryer control system.

increasing it. This is because the effect of increased humidity is to reduce the drying rate, which represents less heat supply from the air, and therefore raises the outlet temperature. The controller (TIC) in Figure 49.14 compensates for this rise in temperature by lowering the inlet temperature, which further reduces the drying capacity of the air.

An alternative fluid-bed dryer control system has been developed based on the temperature-difference technique [45]. This technique is based upon the fact that the temperature difference between the material and the wet bulb is a function of the moisture content of the material at the instant of temperature measurement.

Alden et al. extended the method by estimating the equilibrium relative humidity based on the wet-bulb temperature (T_w), the partial pressure of water at the solid's temperature, and the saturated vapor pressure of water at T_w [46]. Thus, the moisture content of a given material can be inferred from the estimated temperature difference (ΔT) and the equilibrium relative humidity. This method has been successfully implemented to control the endpoint in batch fluid-bed dryer for aqueous and aqueous-alcohol granules. An empirical method was used to determine the desired ΔT value by frequent sampling of the dried material and moisture content determination. The real-time ΔT was estimated using a computer program based on the temperature measurements; this

value was then compared with the desired value and used as the drying-endpoint indicator.

49.6.7 CONTINUOUS CROSS-FLOW GRAIN DRYERS

Traditionally, continuous cross-flow grain dryers have been controlled manually by adjusting the unloading auger revolutions per minute (rpm), which controls the residence time of the grain in the dryer. However, this control often leads to overdrying and excessive stress cracking of some of the grain.

Feedback control based on the exhaust-air temperature was the first automatic control method used in cross-flow grain-drying plants. However, due to the uncertain relationship between the outlet-air temperature and the outlet grain moisture content, this control strategy had proven to be ineffective [47,48].

Due to the above-mentioned difficulties with manual and temperature-based feedback controllers, the control systems based on product moisture content have been investigated and applied in many grain-drying installations in recent years. Forbes et al. compared different control strategies based on the product moisture content for commercial corn-drying units [49]. Three control schemes were studied: (1) a model-based feedforward controller for which the corn-drying process is represented by an exponential-decay-type model with the corn-drying characteristics lumped into a single parameter; (2) a feedback

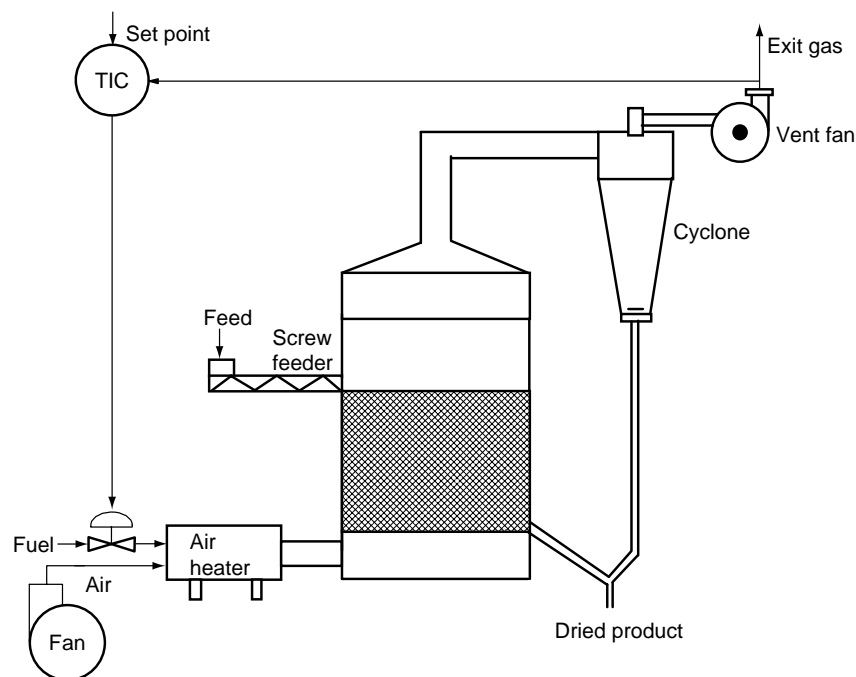


FIGURE 49.14 A conventional fluid-bed dryer control system.

controller; and (3) a feedforward lead/lag controller. Using simulations, it has been proven that MBC system functions better than either PID feedback control or lead/lag feedforward control. This was attributed to the fact that MBC is able to avoid instabilities caused by the frequent and large variations in the feed moisture content without detuning.

A more effective control strategy for cross-flow grain dryers is the feedforward-feedback controller [47,48]. The control algorithm consists of a model-based feedforward controller with feedback correction and dynamic compensation. In this system, the controlled variable is the grain outlet moisture content, the manipulated variable is the auger rpm, and the major load (disturbance) variable is the grain inlet moisture content. The control system consists of a microcomputer, a commercial moisture meter, a tachometer, and the control/dryer model software, which is shown schematically in Figure 49.15. The function of the feedback trim is to force the outlet moisture content back to the desired value (set point) in the case of any changes in the minor load variables (e.g., grain test weight, wind effect), inaccuracy of the process model, or imprecise grain inlet moisture measurement. The dynamic compensation is accomplished by using a pseudo inlet moisture content consisting of a weighted average of the present inlet moisture content and the inlet moisture content of the grain currently in the dryer. The control algorithm has been successfully tested on several commercial cross-flow maize dryers for which the average outlet maize moisture content was controlled to $\pm 0.5\%$ of the set point.

49.7 INTELLIGENT CONTROL SYSTEMS

The control strategies of industrial dryers, which are described earlier, are all based on some mathematical equations or algorithms. Classical control theory gives differential equations or transfer functions, and a modern control theory gives a first-order vector-matrix differential equation based on the state-space method. In these approaches, one has to represent the knowledge and prior information about the system into some analytical structure. However, many difficulties are encountered when designing and applying a dryer control system based on these approaches. These difficulties arise due to one or more of the following reasons:

1. The drying process is complex, time-variant, and nonlinear.
2. Some drying variables (e.g., product quality and color) cannot be measured directly. Other measurements (e.g., moisture content) may be inconsistent, imprecise, incomplete, or not totally reliable.
3. Dryer models are generally approximations to the real process and may require large computing time.
4. Difficulties are experienced when the process operates over a wide range of conditions and suffers from stochastic disturbances.
5. Often, it is not possible to adequately represent the system characteristics such as nonlinearity, time delay, time-varying parameters, and overall complexity.

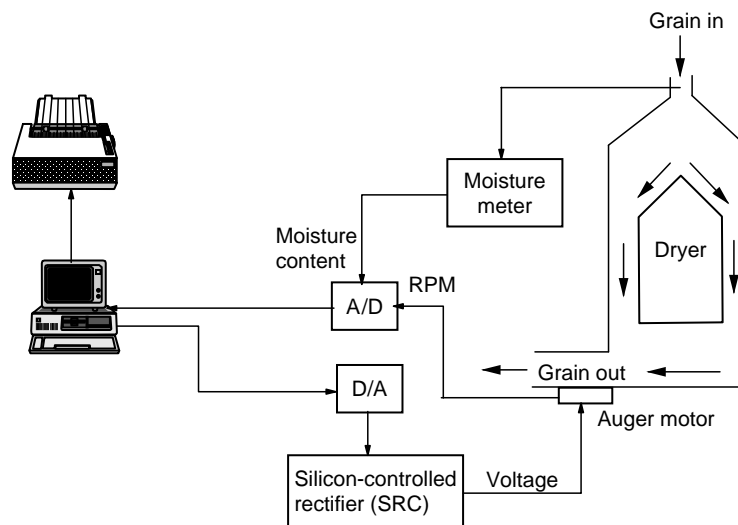


FIGURE 49.15 Cross-flow grain dryer control system.

6. Dryers contain controlled and manipulated variables that exhibit interacting effects.

It is clear that, in order to tackle these difficulties, there is a growing need for the reevaluation of the conventional control methods used in drying operations. This leads us to search for a more general concept of dryer control, one that includes higher-level decision-making, planning, sequencing, learning, logic, and reasoning. These goals present another dimension to the control of an industrial process, i.e., the human element.

The process operators' control strategy is based on anticipation, knowledge (know-how), and experience, and can be considered as a set of heuristic decision rules or "rules of thumb." In order to recognize the human's basic elements and to include the operator's control actions in an automatic scheme, a good user-system interface is required. A useful way to combine this user-system interface with process control is to use intelligent control systems.

Three approaches dominate the real-time intelligent control field: (1) expert systems, (2) neural net controllers (neurocontrollers), and (3) fuzzy logic controllers [50]. These intelligent control systems are based on two types of information processing: symbolic and subsymbolic processing [15,51].

The symbolic process consists of production rules, semantic networks, frames, and objects. Systems built on this principle are called expert systems and are useful for reasoning about process state (temperature, pressure, etc.) and process structures. The subsymbolic process is based on a model of biological neural system. Systems modeled on this principle are called artificial neural networks (ANNs) and are useful for

reasoning about process trends and the complex causal interaction of process variables and states.

Fuzzy logic systems grew out of a desire to quantify rule-based expert systems. Fuzzy set theory had provided us with an effective framework for dealing with fuzzy information and for translating control strategies based on an expert knowledge into an automatic control strategy.

Although not common in the control of industrial dryers, it is anticipated that complex dryers will find useful applications for such controls within the next decade. The following sections present a simple outline of these control strategies and references pertaining to their basic principles and some applications (not necessarily in drying). The interested reader is referred to the literature cited for details.

49.7.1 RULE-BASED EXPERT CONTROL SYSTEMS

An expert system is an intelligent, computerized, knowledge-based system that uses symbolic processes and inference procedures to simulate the decision-making process that an expert performs to solve a problem.

As illustrated in Figure 49.16, an expert control system consists of the following [51–54]:

1. A knowledge base of domain facts, rules, and heuristics associated with the process. These can be obtained from the knowledge of the plant functions, engineering principles, statistical information, and from observing the skilled human operators. The latter can be achieved by means of interviews, questionnaires, and online recording of human-initiated control actions.

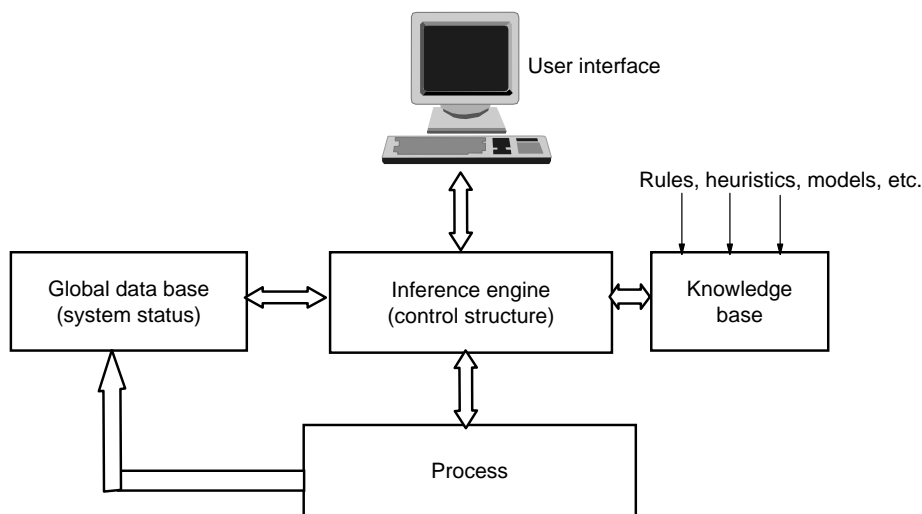


FIGURE 49.16 Structure of an expert control system.

2. An inference engine (or control structure) uses inference procedures to draw conclusions and to infer the correct control action based on the information stored in the knowledge base and the current state of the process.
3. A global database for keeping track of the system status, the input data, and the relevant history of the process.
4. A user interface to provide communication between the user and the program.

Expert systems can be employed in many process control-related usages, such as [15,51,53,55]:

1. Process optimization, process management, trend analysis, alarm processing, control-system design, and adaptive control
2. Enhancement of classical controller performance by sensor failure identification, valve saturation, and process constraints
3. Fault detection, diagnosis, and troubleshooting
4. Supervisory control of simple PID-like controller and startup or shutdown procedures

49.7.2 FUZZY LOGIC CONTROL SYSTEMS

The FLC is a knowledge-based control strategy that uses fuzzy linguistic variables into its rule set to model a “human-operator-like” control approach to cope with the uncertainty in process dynamics or the control environment. In this system, fuzzy logic is used to convert linguistic variables into precise numerical control actions.

The use of fuzzy logic in automatic control was suggested by Zadeh in an attempt to design controllers for complex or ill-defined dynamic systems [56]. Most early applications came out from the researches done by Mamdani and his colleagues [57–59]. Definitions of some terms used in FLC systems are listed in Table 49.2.

In a typical FLC system, the relation between the input x and the output $f(x)$ of the process can be described with a set of linguistic rules of which the typical form is

$$R_i: \text{ If } x \text{ is } A_i, \dots, \text{ and } y \text{ is } B_i, \\ \text{ then } z = f(x, \dots, y) \text{ is } C_i \quad (49.9)$$

where x, \dots, y are linguistic variables representing the process-state variables and z is a linguistic variable representing the process-control variable; A_i, \dots, B_i , and C_i are linguistic values of the linguistic variables x, \dots, y , and z , respectively.

Figure 49.17 shows the basic configuration of an FLC system, comprising four components: a fuzzification interface, a knowledge base, a decision-making logic (control algorithm), and a defuzzification interface.

The literature on FLC has been growing rapidly in recent years, making it difficult to present a comprehensive listing of the wide variety of applications that have been made. Among these are an experimental warm-water plant [65], temperature control of stirred tank reactor [66], activated sludge wastewater treatment process [67], cement kiln [68,69], and startup of catalytic reactor [70]. In drying applications, FLC systems have been developed for simulated rotary dryers [71–73], deep-bed grain and food dryers [74–76], fluidized-bed dryers [77,78], and drum dryers [79].

49.7.3 NEURAL NETWORKS CONTROL SYSTEMS

ANNs are computational paradigms that function in an analogous way to biological neural systems [64,80–82]. They consist of massively interconnected, simple, processing elements (also called neurons or nodes). The strength of the connection among these neurons is characterized by its assigned weight. These weights are adjusted with a training algorithm in order to reach a desired input/output.

ANNs are of interest to the control community because they have the potential to treat many problems that cannot be handled by traditional control techniques. Bhat and McAvoy [83], Psychogios and Unger [84], and Ydstie [85] are among many researchers who have investigated the use of ANNs in process control. A handbook by White and Sofge

TABLE 49.2
Fuzzy Logic Control Defining Terms [60–64]

Fuzzy sets	Sets that do not have a crisply defined membership, but rather allow objects to have grades of membership from 0 to 1.
Fuzzy logic	Kind of logic using graded or qualified statements rather than ones that are strictly true or false. It is much closer in spirit to human thinking and natural language than the traditional logical systems. Basically, it provides effective means of capturing the approximate, inexact nature of the real world.
Linguistic variables	Ordinary-language terms that are used to represent a particular fuzzy set in a given problem, such as positive big (PB), positive medium (PM), positive small (PS), zero (ZO), negative small (NS), negative medium (NM), and negative big (NB).

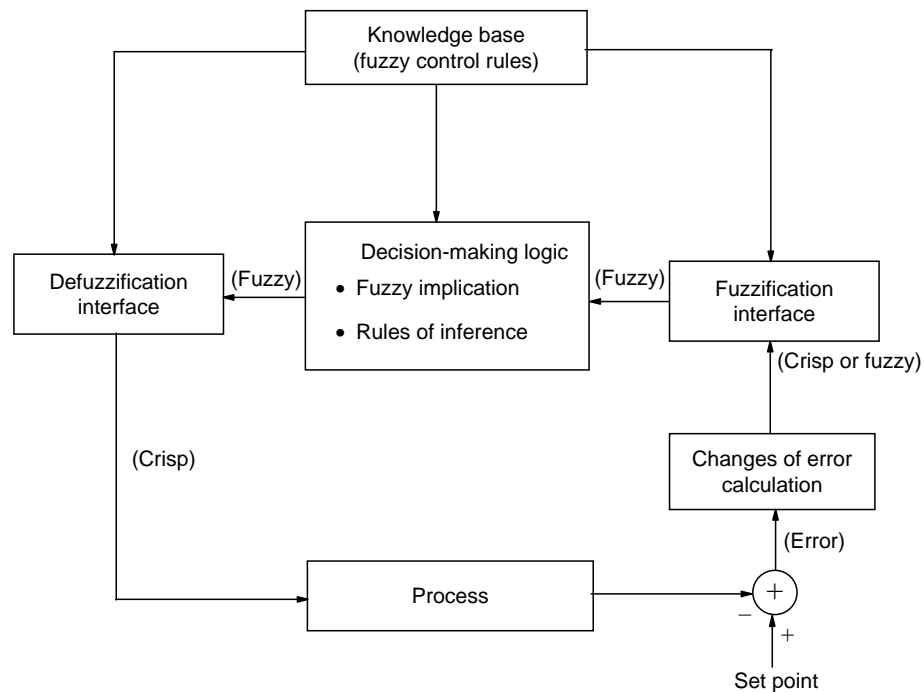


FIGURE 49.17 Basic fuzzy control system configuration.

[86], a book by Miller et al. [87], and an article by Hunt et al. [88] give good reviews of the application of ANNs in process control. Duchesne et al. [89] and Yliniemi [90] have investigated the use of neural networks and hybrid neural-fuzzy control systems in rotary dryers.

In addition to process control, ANNs can be used in many potential applications related to industrial dryers' design, operation, and control, such as:

1. Process modeling
2. Qualitative interpretation of process data for the purpose of control, extraction of control rules for fuzzy logic controllers, and the like
3. Detection of sensor failure
4. Provision of inferred values for signals that are difficult to measure in practical situations
5. Estimation of model parameters (e.g., mass diffusivity) from experimental data

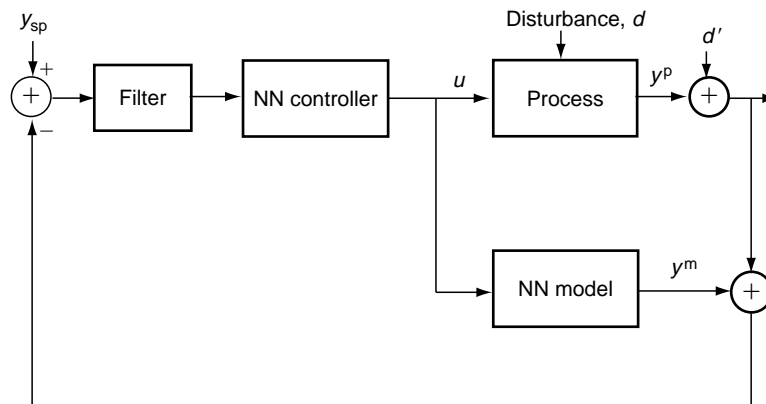


FIGURE 49.18 Neural network internal model control structure.

Figure 49.18 illustrates a neural network IMC system in which both process model and process control are realized using neural networks.

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50 Solid–Liquid Separation for Pretreatment of Drying Operation

Mompei Shirato and Masashi Iwata

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50.1 INTRODUCTION

In many industrial processes, solid–liquid separation operations are combined with drying operations. Figure 50.1 illustrates a process combining drum filters, a rotary kiln, a centrifuge, a disk filter, a spray dryer, and a conveyor dryer. In this combined operation, the solid–liquid separation enhances the quality of the product and decreases the thermal energy of

the drying operations. The final product may be either the solids in the filter cake or the substances dissolved in the filtrate, and these require subsequent drying in order to obtain the final product. Thus, solid–liquid separation and drying are complementary operations in many industrial fields. This chapter will focus on the industrial practice and equipment for solid–liquid separation as a pretreatment process of drying operations. First, the classification and

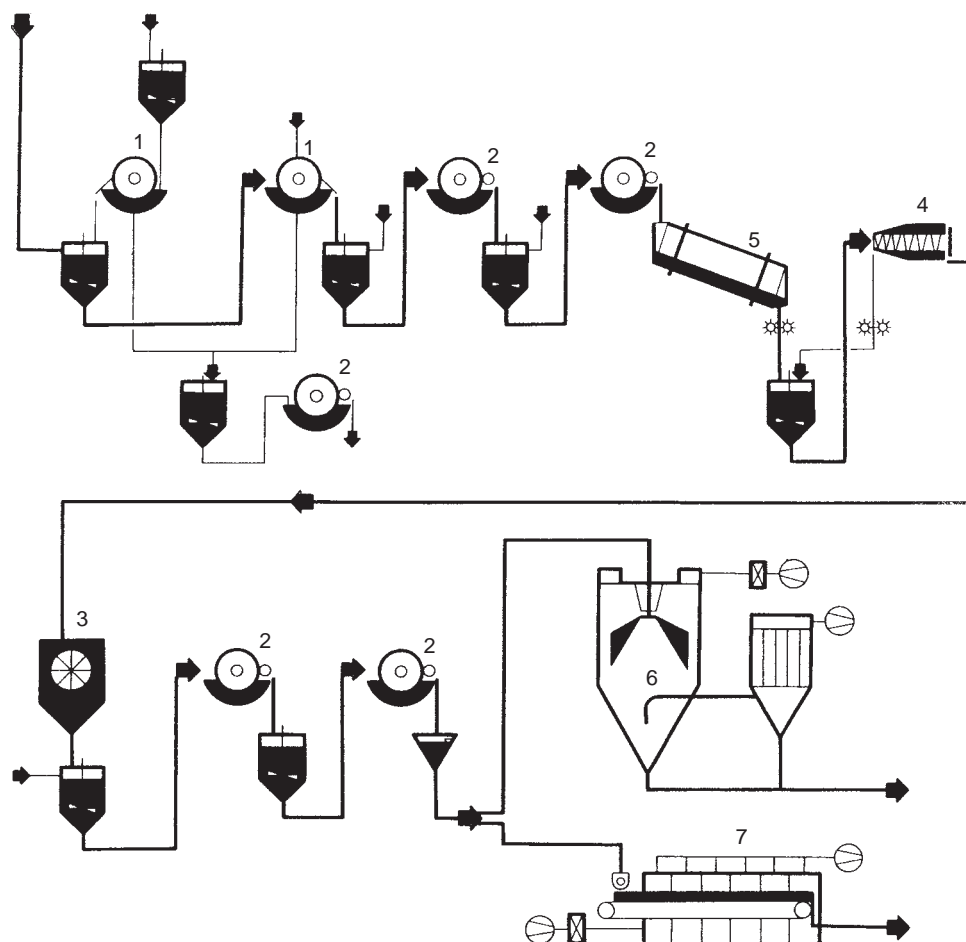


FIGURE 50.1 A process combining solid–liquid separation and drying (1, drum filter; 2, drum filter; 3, disk filter; 4, centrifuge; 5, rotary kiln; 6, spray dryer; 7, conveyor dryer). (From Mitsubishi Kakoki Kaisha, Ltd., Catalog 7-1-1 02–01B-P. With permission.)

selection of nonthermal dewatering equipment are outlined then, batchwise and continuous filters, filtering centrifuges and expression devices are discussed as they are commonly used in combination with dryers.

50.2 AN OUTLINE OF SOLID–LIQUID SEPARATION OPERATION

The diversity of substances and the competitive efforts have evolved into many variants for solid–liquid separation equipment. Before categorizing the equipment for pretreatment of the feed material to the drying operation, it is useful to look at the entire field of solid–liquid separation. It can be divided as shown in Table 50.1 [1,2].

Conditioning is utilized to alter the properties of the slurry so that it is easier to separate. For example, particle size may be enlarged by chemical treatment, producing flocculation or coagulation. In solids con-

centration, part of the liquid may be removed by thickening or hydrocycloning so that the load on the filter is decreased.

Filterates can be divided into two broad categories: cake and depth filtration. Cake filtration can be further classified into centrifugal, and pressure, vacuum, and gravity (PVG) operations. In cake filtration, the particles from a slurry are blocked at the surface of a supporting porous medium whereas the fluid passes through. In depth filtration, the particles are captured in the interstices of the medium and no cake is formed on the surface of the medium. In many processes, a stage of depth filtration precedes the formation of a cake. The first particles may enter the medium and, with very dilute slurries, there may be a time lag before a cake begins to form. Smaller particles enter the medium while larger particles bridge the openings and start the buildup of a surface layer. In general, depth filtration is used for removing small quantities of contaminants. Cake filtration is

TABLE 50.1
Stages of Solid–Liquid Separation

Conditioning	Chemical	Flocculation Coagulation		
	Physical	Crystal growth ^a Freezing and other physical changes Filter aid addition Hydrocycloning		
Solids concentration	Thickening ^a Clarification			
Solids separation	Cake formation	Press, vacuum, and gravity (PVG) filters Centrifuges	Batch ^a Continuous ^a Filtering Sedimenting	Perforated bowl ^a Solid bowl ^a
	No cake formed	Deep granular bed Cartridges		
	Filtrate	Polishing	Membranes Ultrafiltration	
Posttreatment	Cake	Washing Deliquoring	Displacement ^a Reslurry ^a Drainage ^a Mechanical expression ^a	

^aCommonly combined with drying operation.

primarily employed for more concentrated slurries. In actual processing operations, both PVG filtrations and centrifugation are encountered. Many separations can be achieved equally well with either PVG or centrifugal equipments. Cost, operating, and maintenance considerations determine the actual choice.

Posttreatment is also a very important factor in the selection of equipment. For removing the impurities from a filter cake, cake washing is performed by permeating the wash liquor through the cake or by reslurring the filter cake. In general, a low moisture content of the cake is attained by squeezing mechanically or hydraulically, by blowing with air, or by aiding the drainage on a rotary vacuum filter by means of steam. However, capillary pressures for 10- μ m particles are so high that it is difficult to blow air or steam through a cake with such particles. If a rotary vacuum filter were chosen, there would be a limitation on moisture reduction for cakes with small particles unless an additional unit for mechanical expression was available. In general, it is essential to balance the additional cost of producing a dry cake against the thermal requirements of dryers.

50.3 INITIAL CHOICE OF EQUIPMENT

The initial choice of the general class of equipment should be tentatively made based on the rate of cake growth. Specific equipment types are related to cake-formation rate as given in Table 50.2 [1,2].

Fast-filtering slurries are most often found in mineral processing and as products of crystallizers. They settle rapidly and form a thick cake in a few seconds. The rapid rates require filtration with continuous vacuum filters or, for small-scale production, with batch filters. Medium-filtering slurries are those that can be kept in suspension by gentle stirring. With these slurries, a 3- to 5-cm cake is formed with vacuum in about a minute. Rotary drum, Nutsche, or batch pressure filters can be used, depending on the production scale. If a high degree of washing is required, a plate-and-frame type may be most cost-effective. Slow-filtering slurries require pressure filters, generally of a batch type, to increase flow rates. Dilute slurries include those that do not form a cake that is quick enough to be discharged from a continuous filter. The selection depends on such factors as the scale of production, the water content of the cake, and the amount of washing that is required.

50.4 CAKE FILTRATION

50.4.1 THEORETICAL ASPECTS OF CAKE FILTRATION

In cake filtration, once a layer of solid particles has formed on the filter medium, its surface becomes the actual filter medium. As the solids get deposited they add to the thickness of the cake whereas the clear liquid passes through. The cake is therefore composed

TABLE 50.2
Guide to Filter Selection

Slurry Characteristics	Fast Filtering	Medium Filtering	Slow Filtering	Dilute	Very Dilute
Cake-formation rate	cm/s	>1.0 cm/min	0.12–0.62 cm/min	<0.12 cm/min	No cake
Normal concentration	>20%	10–20%	1–10%	<5%	<0.1%
Settling rate	Rapid, difficult to suspend	Fast	Slow	Slow	—
Leaf test rate, kg/(h · m ²)	>2500	250–2500	25–250	<25	—
Filtrate rate, cm ³ /(min · cm ²)	>20	0.8–20	0.04–0.8	0.04–8	0.04–8
<i>Vacuum filters</i>					
Multicompartment drum	○	○	○		
Singlecompartment drum	○				
Hopper dewaterer	○				
Top feed	○				
Scroll discharge	○	○			
Tilting pan	○	○			
Belt	○	○			
Disk		○	○		
Precoat				○	○
Batch vacuum leaf		○	○	○	○
Batch Nutsche	○	○	○	○	○
(vacuum and pressure)					
<i>Pressure filters</i>					
Plate-and-flame		○	○	○	○
Vertical leaf		○	○	○	○
Tubular		○	○	○	○
Horizontal plate	○	○	○	○	○
Continuous-pressure precoat				○	○
Cartridge					
<i>Centrifuges</i>					
Solid bowl	○				
Filtering	○				
Sedimenting		○			

of a bulky mass of particles, among which irregular small channels run. The flow of liquid through the channels is always laminar and can be described by the following Ruth's filtration equation:

$$\frac{dV}{Ad\theta} = \frac{p}{\mu[\alpha(W/A) + R_m]} \quad (50.1)$$

In this expression, the differential or instantaneous rate of filtration per unit area ($dV / Ad\theta$) is given as the ratio of a driving force, pressure p , to the product of viscosity μ , and the sum of cake resistance $\alpha(W/A)$ and filter medium resistance R_m . The mass of dry cake W is related to the volume of filtrate V by a simple material balance, thus:

$$W = cV = \frac{\rho s}{1 - ms} V \quad (50.2)$$

where c is the mass of dry cake solids per unit volume of filtrate, ρ is the density of the filtrate, s is the mass

fraction of solids in the slurry, and m is the mass ratio of wet cake to dry cake.

α represents the average specific resistance of the cake, which is a constant for filtration under constant pressure. In the usual range of operating conditions, it can be represented as a function of the filtration pressure p in the form

$$\alpha = \alpha_0 p^n \quad (50.3)$$

where α_0 is a constant determined largely by the size of the particles forming the cake; n is the cake compressibility, varying from 0 for rigid, incompressible cakes, such as fine sand and diatomite, to 1.0 for very highly compressible cakes. For most industrial slurries, n lies between 0.1 and 0.8. R_m represents the resistance of unit area of the filter medium, but includes other losses in the system across which the liquid pressure drops.

The mutual effects of the operating variables can be seen in Equation 50.1. When the cake is composed

of hard granular particles that make it rigid and incompressible, an increase in pressure results in no deformation of the particles or their interstices, i.e., $n = 0$. If filter medium resistance is neglected, Equation 50.1 becomes

$$\frac{dV}{Ad\theta} = \frac{p}{\mu\alpha_0(W/A)} \quad (50.4)$$

For incompressible cakes, therefore, the flow rate is directly proportional to the area and pressure and inversely proportional to the viscosity, to α_0 , and to the total amount of cake (or filtrate).

When the cake consists of extremely soft, easily deformed particles, such as metal hydroxides, n approaches 1.0. Equation 50.1, with the filter-medium resistance again neglected, reduces to

$$\frac{dV}{Ad\theta} = \frac{1}{\mu\alpha_0(W/A)} \quad (50.5)$$

Thus, the rate is independent of pressure for very compressible cakes.

The effect of pressure shown earlier is modified in most industrial filtrations in which cake compressibility usually lies between 0.1 and 0.8. Furthermore, the resistance of the filter reduces the effects of the respective variables. It has been found, however, that an increase in pressure causes a nearly proportionate increase in the flow rate in the filtration of granular or crystalline solids. Flocculent or slimy precipitates, on the other hand, have their filtration rates increased only slightly by an increase in pressure. Some materials have a critical pressure above which a further increase results in an actual decrease in flow rate.

Cake thickness is an important factor in determining the capacity and design of a batch filter and it determines the cycle of operation. In constant-pressure filtration, the integration of Equation 50.1, with the medium resistance neglected, yields

$$\left(\frac{V}{A}\right)^2 = K\theta \quad (50.6)$$

$$K = \frac{2p(1 - ms)}{\mu\alpha\rho s} \quad (50.7)$$

where K is Ruth's constant-pressure filtration coefficient. V/A is proportional to W/A , and thus to the thickness of the cake. It may be noted, as a consequence of these relationships, that the filtration time for a given quantity of filtrate is proportional to the square of the thickness of the cake at the end of the filtration. Maximum filter productivity $W/A(\theta + \theta_d)$ under batchwise constant-pressure filtration, with the medium resistance neglected, is

obtained when the filtration time θ is equal to the downtime θ_d , which is the sum of the time required to remove the product cake and to prepare the medium for the next cycle. The greater the resistance of the filter medium and the longer the preparation downtime, the longer is the optimum filtration time and the thicker is the optimum cake.

The effect of viscosity is as indicated by Equation 50.1 the filtrate flow rate at any instant is inversely proportional to the filtrate viscosity.

The effect of temperature on the filtration rate of incompressible solids is principally evident through its effect on viscosity. Higher temperatures permit higher filtration rates. Compressible sludges are affected in more complicated ways by temperature increases, but the general effect is inclined to be increased filtration rate.

Particle size significantly affects cake and medium resistances. Even small changes in particle size have an effect on the coefficient α_0 for cake resistance in Equation 50.3, and larger changes affect the compressibility n . Decreased particle size results in lower filtration rates and higher moisture content of the cake, and sometimes in better washing efficiency. It is important, therefore, that close control be kept of the particle size in the feed to the filter. Degradation of particle size by violent agitation or pump action must be avoided. Conditioning of a slurry by both physical and chemical means can make the separation easier by increasing the particle size. Changing the temperature, recrystallization, aging, and freezing are sometimes effective physical techniques. The addition of certain electrolytes such as alum, lime, and iron salts may effect particle-surface changes and act as coagulants or flocculants. The addition of high-molecular-weight polyelectrolytes is often a successful pretreatment for entrapping fine particles forming a flocculation structure. Control of pH may affect the surface charge or zeta potential of the particles.

When a filter cake is washed, the rate of wash throughput is generally the same as the final rate of filtration, if (1) the washing pressure is the same as the final filtration pressure, (2) the wash liquid and the filtrate have similar physical properties, (3) the wash liquor does not interact with the filtrate, and (4) there is no rearrangement of the cake.

50.4.2 BATCH FILTER

50.4.2.1 Filter Press

There are two basic types of the filter press: (1) the plate-and-frame type and (2) the recessed-plate type. The plate-and-frame filter press (Figure 50.2) consists of a skeleton framework made up of two end supports

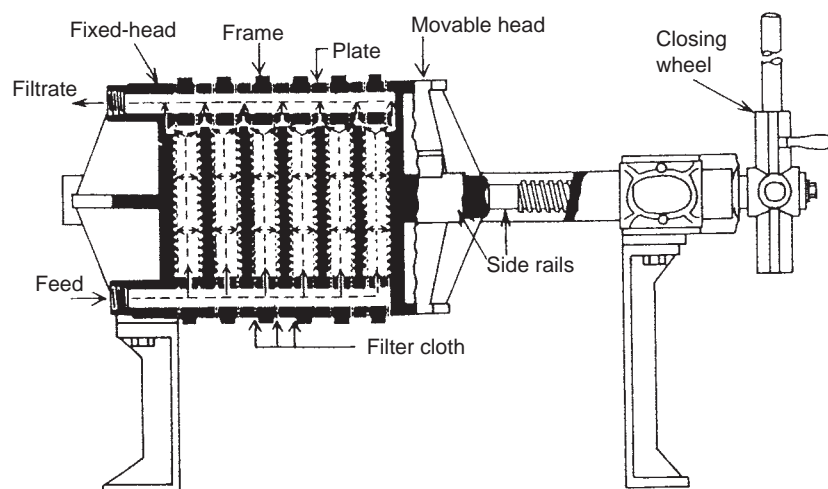


FIGURE 50.2 Plate-and-frame filter press. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

connected by two horizontal parallel bars. On the bars, a varying number of filter chambers are assembled that consists of medium-covered plates alternating with frames that provide space for the cake. The chambers are closed and tightened by a screw or hydraulic ram, which forces the plates and frames together, making a gasket of the filter cloth. The charge enters the filter press under pressure and fills each chamber approximately simultaneously. The liquid passes through the filter medium, which in turn retains the solids. The clear filtrate is removed at a discharge outlet. The filter cake builds up until the frames are full—judging by filtering time, decrease in feed rate, or rise in backpressure. Once the frames are full of cake, filtering is stopped and wash liquid is applied if necessary; deliquoring the cake may follow.

The recessed-plate filter (Figure 50.3) has no frame; each plate serves as both plate and frame. The feed slurry enters through a large central port, and filter cake begins to form in the recess in each side of the plate. In order to minimize the strain on the filter cloth, the depth of the recess is usually about 16 mm, giving a maximum cake thickness of about 32 mm.

The advantages of a filter press are adaptability to high pressures, production of a dry washed cake, durability, easy alterability of media and frames, and flexibility. Disadvantages include high labor cost (unless mechanized), leakage, and exposure of product to air during discharge. The provision of mechanized systems to move the filter plates and to open and close the filter press at the end and commencement of the filtration cycle can reduce the operating workforce requirements.

50.4.2.2 Tank Leaf

The tank-leaf pressure filter (Figure 50.4) features a number of leaves (filtering elements) suspended within a closed vessel. Common variations involve the arrangement of the tank and leaves into various vertical and horizontal combinations. Vertical tanks normally contain vertical leaves whereas horizontal tanks may have either vertical or horizontal leaves. The leaves are hollow shells fitted with support grids and covered with a metal gauze or woven fabric. They

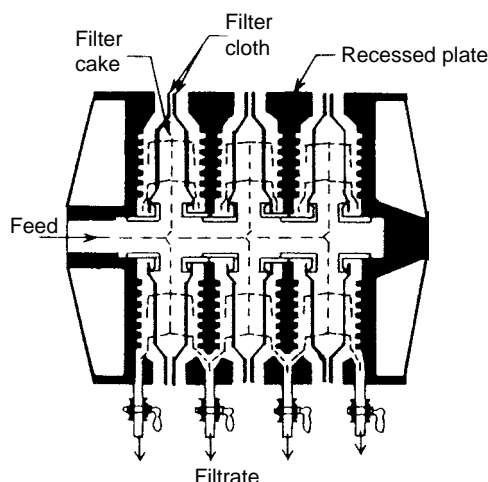


FIGURE 50.3 Recessed-plate filter press. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

may be stationary or rotating and may be square, rectangular, trapezoidal, circular, a segment of a circle, or elliptical in shape. They have filtering surfaces on both faces. Spacing varies from 50 to 150 mm center to center, but is rarely less than 75 mm. Drainage outlets may be located at the bottom, top, and center or in various combinations. Each leaf design lends itself to several discharge methods. For wet discharge, elements may be cleaned by a stationary, rotating, oscillating, or traveling sluice; or the element may be rotated past a fixed sluice. Sluice action may be coupled with sparge or vibration. Dry discharge may be accomplished manually or mechanically, using sudden shock, vibration, air, or centrifugal spinning. With some filters, the leaves are rotated slowly against a brush to assist in cake release. In other designs, the tank or leaves are

sometimes rotated 90° after filtration and before discharge. Filtration is accomplished with the leaves in a horizontal position, and discharge is carried on with the leaves in a vertical position to take advantage of gravity.

50.4.2.3 Plate Filter

The horizontal multiple-plate pressure filter (Figure 50.5) consists of a number of horizontal, circular, drainage plates and guides placed in a stack in a cylindrical shell. Flow is downward; the cake builds up on the filter medium, which rests on a plate. A scavenger plate permits more complete recovery. The horizontal filter is good for applications in which small quantities of cake or intermittent flow rates are involved and in which cleanliness or sterile conditions are essential, as in food and pharmaceutical industries. Disadvantages are high labor cost and size limitations. Normally, plates are designed for a maximum pressure of 0.35 MPa.

50.4.2.4 Tray Filter

The tray filter (Figure 50.6) consists of a series of trays in a horizontal tank open at one end. Each tray is normally drained through an individual outlet to a manifold outside the tank. Piping is arranged so that the liquid may be transferred from the bottom of the tank to the top tray and to the other trays by overflow. The tray filter, having an enclosed tank, provides an efficient drying cycle. This allows the almost complete recovery of solids and filtrate. Because of the horizontally formed, tray-held cake, the filter may be run intermittently without the cake moving or falling off the tray. Batches may be run up to the cake-holding capacity of the tray. If required, solids may be dissolved or melted in the tray. The primary advantage of the tray filter is that the complete recovery of solids and liquid is practical. There are three disadvantages: (1) filtration occurs on the upper surface of the tray only, (2) it is size-limited, and (3) operating costs are higher than for two-sided filtration units.

50.4.2.5 Tank Tube

A tank-tube filter (Figure 50.7) is essentially the same as vertical-leaf type, except that hollow tubular elements are mounted in a pressure vessel. This type may be competitive with the vertical-leaf type, when a dry-cake discharge is not required. Filtrate flows through the filtering tubes into a discharge manifold whereas solids deposit outside. The tubes may be made of porous carbon, porous ceramic, plastic, sintered metal,

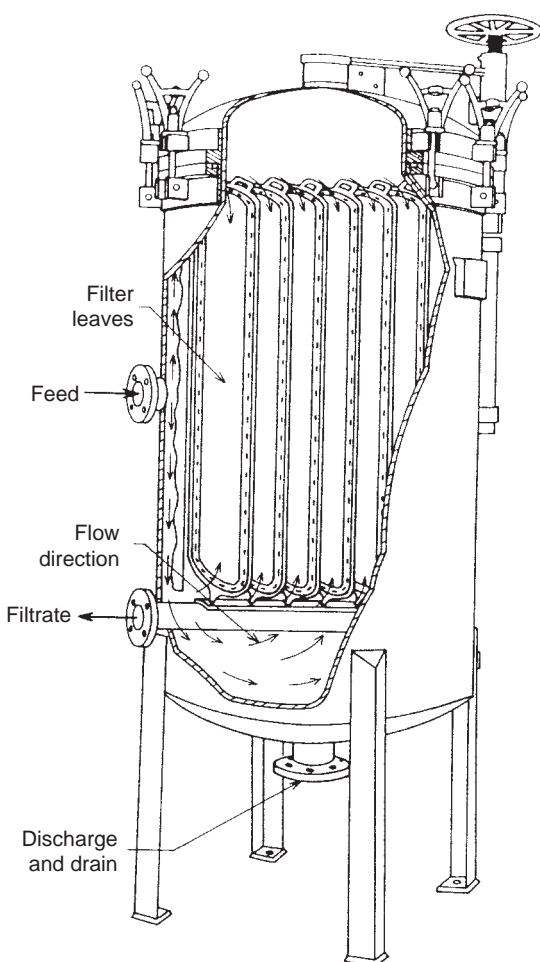


FIGURE 50.4 Vertical tank, vertical leaf filter. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

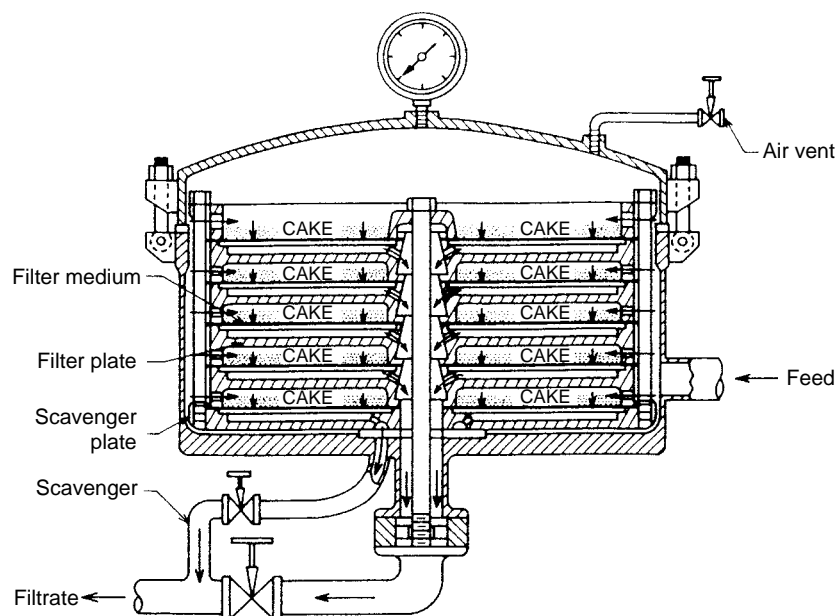


FIGURE 50.5 Horizontal plate filter. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

or screen-wrapped perforated pipe. Some types of tubes may be fabric-covered; others are made flexible so that mechanical movement or backpressure can be used to detach the cake. A major advantage of this type over the vertical-leaf type is that a relatively high-pressure backwash can be utilized for discharging the cake and cleaning the medium. Its tubes must be uniformly cleaned by backwashing to prevent the ensuing problems. Dry discharge is not generally feasible, except for some designs. The construction of the tube must be able to withstand repeated, reverse-flow, backwash action.

50.4.2.6 Nutsche

Similar in operation to the well-known Buchner funnel, the Nutsche filter is usually constructed as a cylinder divided into two parts by a horizontal filter plate covered with a convenient medium. Slurry is placed in the upper chamber and filtration is accomplished by gravity, vacuum, or pressure. A major field of application is for small-scale or pilot-plant production rates on free-filtering materials. A Nutsche filter has the advantages of simplicity of construction and operation, effective displacement washing, suitability for testing procedures and pilot-plant studies, suitability for free-flowing crystalline materials, and suitability for corrosive materials. Its use demands a high labor cost for cake removal, and it occupies a large floor space per unit filtering area. It may be constructed of a wide range of materials including reinforced plastic or ceramics.

In recent years, this unit has frequently been preferred because of its great adaptability to various filtration processes. With the intention of decreasing high labor requirements, several developments have occurred that have automated the operation. [Figure 50.8](#) is a new type of Nutsche filter, which automatically carries out cake filtration, dewatering,

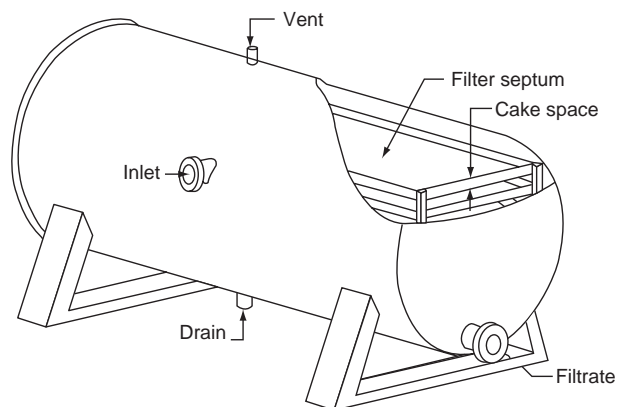


FIGURE 50.6 Horizontal tray filter. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

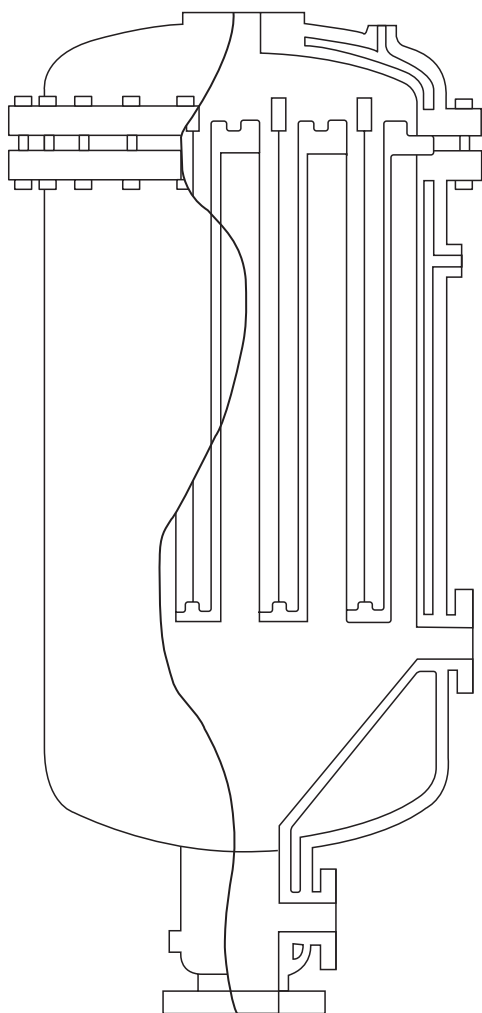


FIGURE 50.7 Tubular filter. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

washing, drying, and discharge. During filtration, cracks may occur in the filter cake; they can be smoothed out by an agitator or discharge device.

50.4.2.7 Vacuum Leaf

The vacuum leaf is an open-leaf filter operating under suction. An early version, the Moor filter, consists of a number of rectangular leaves manifolded together and carried by an overhead crane. The leaves are dipped successively in a slurry tank, where vacuum filtration takes place; a holding tank, where cake washing occurs; and a receiving container, where cake discharge is performed, usually by backblowing. If the slurry is agitated sufficiently to prevent cake formation, it can be used as a thickener. Elements can

be made in tubular form for clarification filtration. The advantages of this filter include a wide range of available construction materials; low cost in large units; low labor cost for easily discharged cake; and easy inspection of elements, filter medium, precoating, and cake formation. Its disadvantages include a limitation on volatile liquids, temperature controls imposed by its open construction, and a pressure differential limitation imposed by its use of a vacuum.

50.4.3 SEMICONTINUOUS FILTERS

50.4.3.1 Horizontal Pan

This filter (Figure 50.9) consists of a series of open pans located in a horizontal plane rotating about a central, vertical axis. The pans form numerous wedge-shaped sections, which slope toward and connect directly to a common filter valve under the center of the unit. Feed is applied by a pipe and a weir box distributor above the unit. The filter cake may be washed with spray weirs after initial dewatering takes place. Countercurrent washing is possible. Cake discharge is accomplished by separately driven scrolls or paddles, or simply by tilting each sector at the completion of filtration (tilting-pan filter). The filter is suitable for granular-product dewatering with or without washing and for very high solid loads and high hydraulic requirements. Its disadvantages are that it utilizes one-sided element filtration, it requires a large space, it is not suitable where the media may be easily blinded, and its cost is relatively high.

50.4.3.2 Automatic Filter Press

Basically, this unit is a filter press with an automatic cake-discharge system. With the conventional filter press, hours of work are required, especially for cake discharge. In order to minimize these shortcomings, yet to maintain and further develop the advantages, filter press automation was pursued by the following three means:

1. Automatic cake discharge by design of automatic filter-plates shifting and filter-cloth vibrating mechanism
2. Automatic filter-cloth washing to eliminate clogging on the cloth
3. Filtration capacity increase by reduction of the miscellaneous time required for processes other than filtering

Classification of automatic filter presses now in practical use are as follows: recessed-plate type and plate-and-frame type; lateral type and vertical

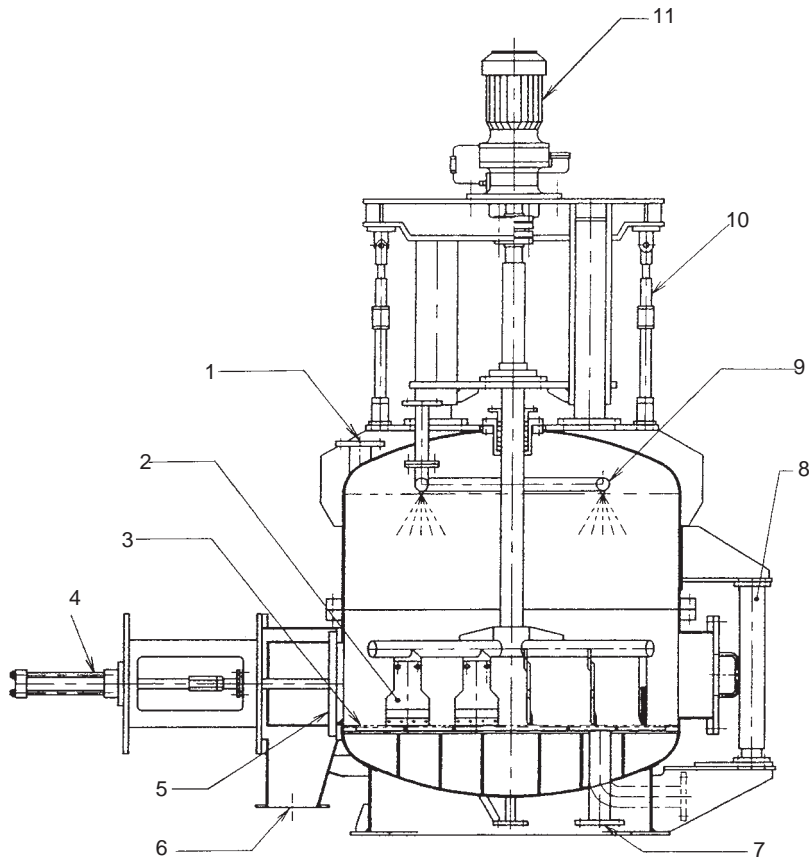


FIGURE 50.8 Nutsche filter (1, feed; 2, impeller; 3, filter plate; 4, cake-discharge door drive; 5, cake-discharge door; 6, outlet; 7, filtrate; 8, filter opening cylinder; 9, washing device; 10, impeller drive cylinder; 11, impeller drive). (From Mitsubishi Kakoki Kaisha, Ltd. With permission.)

type; fixed filter-cloth type and filter cloth-traveling type.

Most automatic filter presses are of the recessed-plate type because of its simple structure and high pressure. With the lateral type, filter chambers are formed when perpendicular filter plates are placed together in a horizontal direction. The lateral, fixed-cloth filter press has a cloth hung in an inverted V shape between the plates so that the cloth becomes inclined when the filter plates are opened, prompting the removal of cake on it (Figure 50.10). The cloth-traveling type (Figure 50.11) moves an endless filter cloth between the plates to peel off the cakes.

With the vertical type (Figure 50.12), the plates are placed one upon another vertically to form filter chambers, and only the filter cloth-traveling system is used. At the end of the cycle, the plates are separated and the endless filter cloth put into motion. Cake is discharged as the cloth leaves the filter chamber. Fresh medium is introduced into the filter, and the used sections are washed and cleaned while the next cycle is in progress.

A flexible membrane is utilized for pressing and deliquoring the cake by hydraulic pressure. Such a filter produces a cake with low residual moisture content; the blinding of cloth is minimized, and it is flexible in its usage. High cost is its primary disadvantage.

50.4.4 CONTINUOUS FILTERS

50.4.4.1 Drum Filters

A standard drum filter (Figure 50.13) consists of three main parts: (1) a drum with an automatic filter valve, (2) a slurry reservoir with an agitator, and (3) a scraper for cake discharge. The drum, or cylinder, having a porous wall and covered on the outside with filter media, rotates about a horizontal axis with a portion submerged in the slurry. The filter operates continuously through stages of cake formation, washing, drying, and discharge. A vacuum is the normal driving force, though enclosed pressurized units are sometimes built. The drum is usually

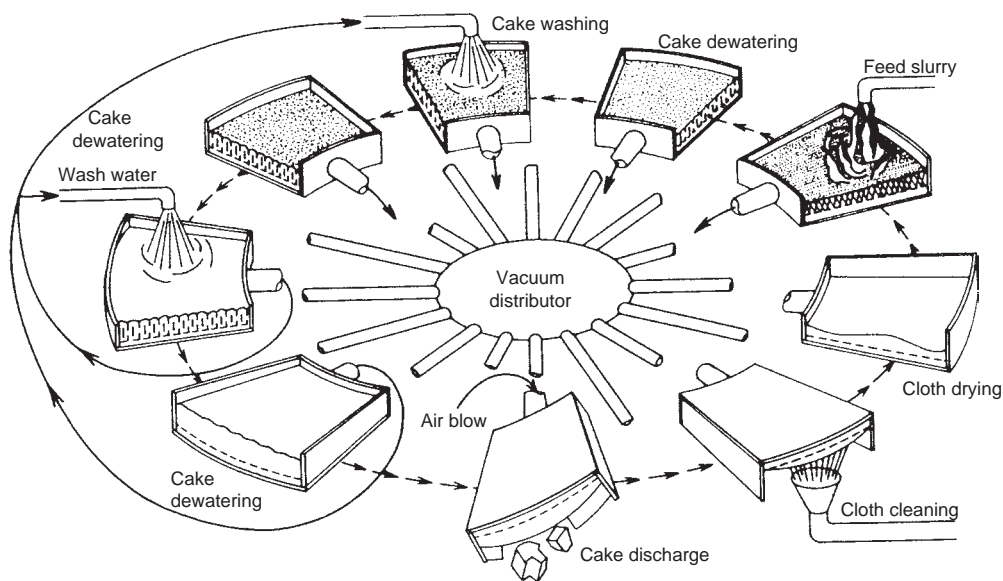


FIGURE 50.9 Tilting-pan horizontal filter. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

subdivided into a number of separate compartments so that the various stages of filtering, washing, drying with air or steam, and discharging with an air blow, can be carried out. Rotational speeds usually vary from about 0.25 to 3.0 rpm. As each compartment is submerged in the feed slurry, cake formation takes

place. Some drum filters are furnished with a single internal compartment.

Various systems of cake removal are in use, all of which can be assisted by air blowback. In addition to simple scraper discharge, string or coil, belt, and roller discharge methods may be employed. In the

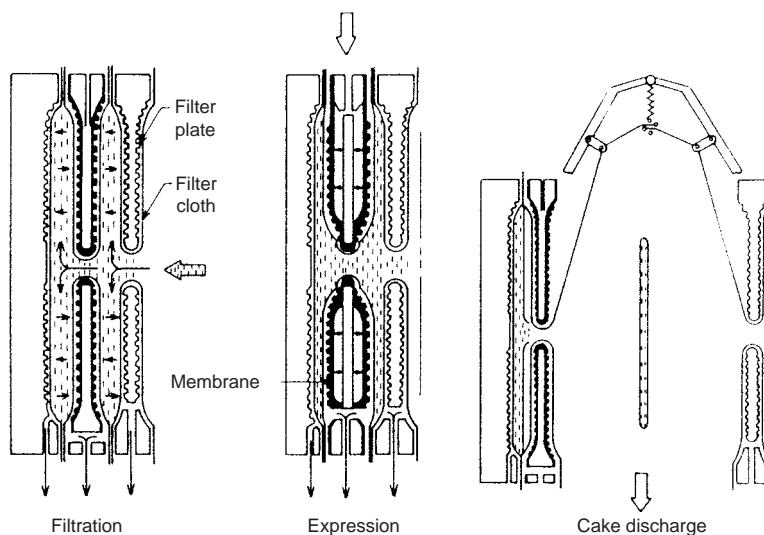


FIGURE 50.10 Lateral, recessed-plate, fixed filter cloth, expression-type automatic filter press. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

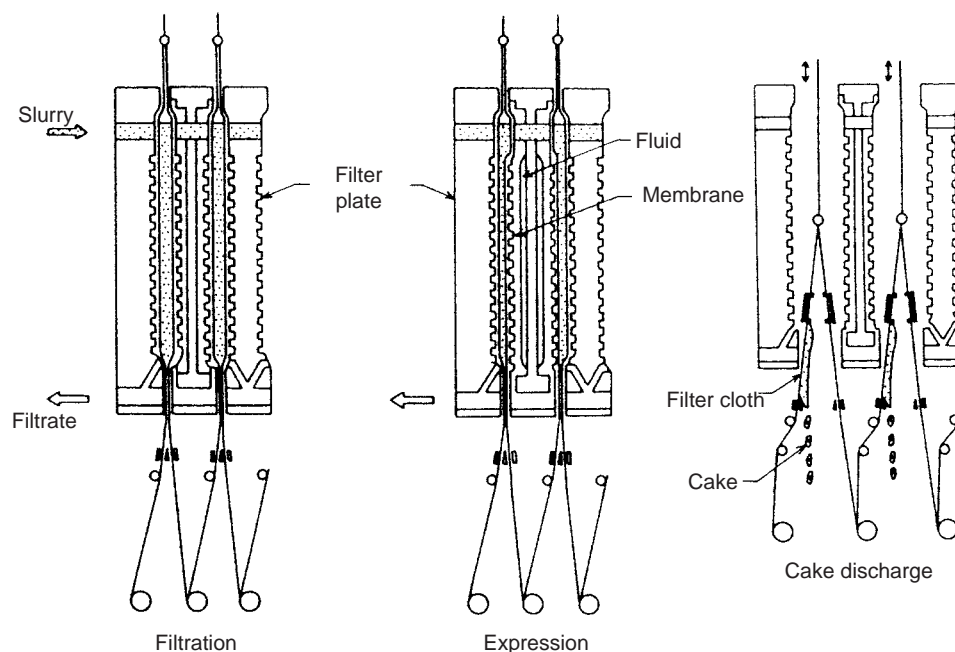


FIGURE 50.11 Lateral, recessed-plate, separate filter cloth-traveling, expression-type automatic filter press. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

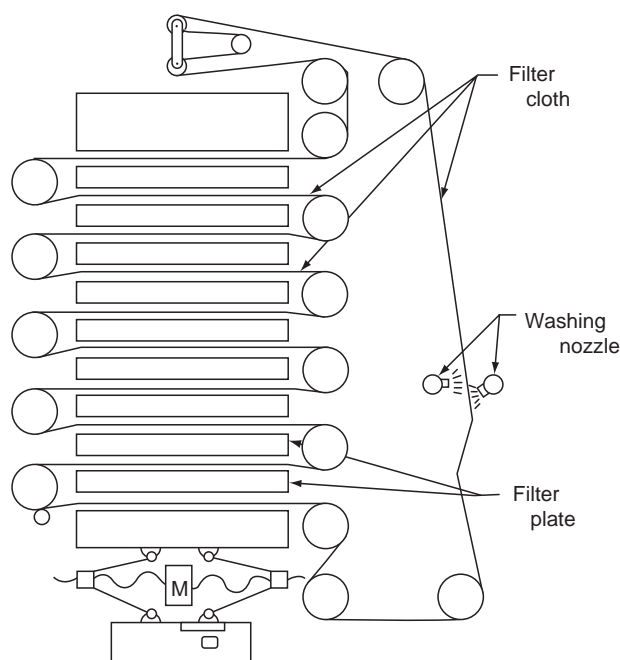


FIGURE 50.12 Vertical, recessed-plate, endless filter cloth-traveling, expression-type automatic filter press. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

string discharge drum filter (Figure 50.14a), endless strings are wound over the filter medium to lift off and remove the filter cake. The strings normally pass from the drum over a discharge roll where, because of the small radius of the roll, they make a sharp turn and dislodge the cake.

In belt discharge drum filters (Figure 50.14b), the medium is passed on an endless belt over the drum for filtering and then over a small roller to dislodge the cake. The medium may be cleaned as it returns. With roller discharge (Figure 50.14c), a roller is rotated close to the surface of the drum. The filter cake passes to the roller near the discharge point where the vacuum is removed from the compartment, and then scraped off.

String discharge removes some cakes very well. However, thin cakes cannot be discharged by this method and some noncohesive materials may not be lifted off. Belt discharge, like the automatic press, allows for good cleaning of the media. It is relatively expensive and requires care in keeping the media aligned. Roller discharge depends upon the adhesion to the roller instead of the media. It is simple and inexpensive when the cake has properties such that it can be used.

Although rotary drum filters are normally employed for materials of moderate to low resistance, they can be used as precoat filters for highly resistant slurries. A thick coat of several centimeters of filter

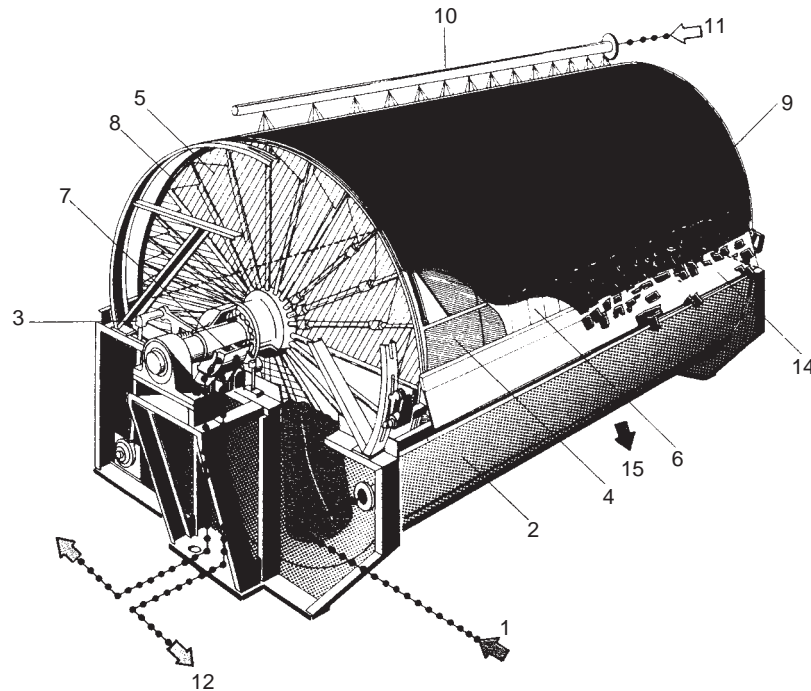


FIGURE 50.13 Scraper discharge, rotary drum filter (1, slurry; 2, filter trough; 3, pendulum agitator; 4, filter cells; 5, drum; 6, filter cloth; 7, control head; 8, filtrate pipes; 9, filter cake; 10, washing device; 11, wash liquid; 12, mother filtrate; 13, wash filtrate; 14, discharge device; 15, solid). (From Mitsubishi Kakoki Kaisha, Ltd. Catalog 7-1-1 02-01B-P. With permission.)

aid is deposited on the drum prior to the start of filtration. The precoat is then gradually cut off as the solids are deposited, thereby retaining a low overall flow resistance. An additional problem of bottom-feed drum filters is their inability to handle very fast-settling slurries that cannot be kept in suspension. This limitation may be overcome by using top-feed drum filters.

For high capacity per unit area and superior cake-washing capacity, the single-compartment drum filter may be used. For reasons such as safety and prevention of vapor loss, enclosed vacuum or pressurized types may be utilized.

The advantages of drum filters in general are that they are capable of providing continuous filtration, washing, and partial drying. With relatively large

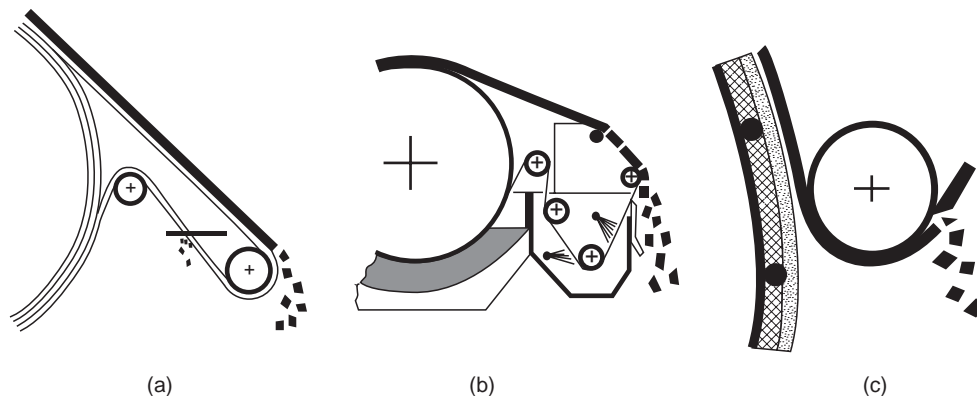


FIGURE 50.14 (a) String discharge, (b) belt discharge, and (c) roll discharge drum filters. (From Mitsubishi Kakoki Kaisha, Ltd. Catalog 7-1-1 02-01B-P. With permission.)

particles, it is possible to produce nearly dry cakes by using steam; and they permit wide latitude in operating conditions through the control of submergence, vacuum, and rotational speed. The disadvantages are that they have a relatively high cost, especially in small sizes; they are not applicable to cakes that build up slowly; they are not applicable to batch-type processes; the precoat type may necessitate the high cost of precoat filter aids; and pressure units are normally limited in size by cost; and, in addition, depend upon the efficacy of cake removal from the tank in dry form.

50.4.4.2 Disk Filter

The vacuum disk filter (Figure 50.15) consists of a series of circular filter elements mounted on a rotating, horizontal, hollow shaft at regular intervals. Each element, or disk, is divided into sectors ribbed on both sides to support filter cloth. A vacuum is applied to each row of sectors by an automatic valve. Filtrate and air are removed through radial pipe nipples connected to a manifold, which leads through the hollow shaft to the valve. The operation of the disk filter is very similar to that of the drum filter. The disk filter provides a much greater filtering area for a given floor space and the least cost per unit area. Only a small amount of wash can be applied, however, because of the vertical elements. Other limitations of the standard disk are poor cake-dewatering, poor discharge,

and unsuitability for precoat operation. Cloth washing is virtually impossible.

50.4.4.3 Band or Horizontal Belt

This type of filter (Figure 50.16) consists of an endless, reinforced, rubber belt mounted in tension. Covered with the filter cloth, the belt passes over suction boxes between drums and pulleys. The slurry to be dewatered is fed onto the upper side of the belt at one end. A vacuum applied through the suction box dewaterers, washes, and discharges the filter cake, which is discharged as it passes over the end pulley. The speed of the belt is adjusted to produce a cake of suitable thickness. The belt filter has all the advantages of the horizontal pan filter. The filter cloth is conveniently washed while it returns to the start of the filtration cycle. Disadvantages are the high initial cost and large floor space per unit filtration area. Other limitations are similar to those for the horizontal pan filter, except that speed is not limited by the method of cake discharge.

50.5 CENTRIFUGAL FILTRATION

In centrifugal filters, separation is effected by directing the solid-liquid mixture on to the inner surface of a perforated, rotating bowl. In fine separations, the inner surface of the bowl will be lined with an appropriate filter medium.

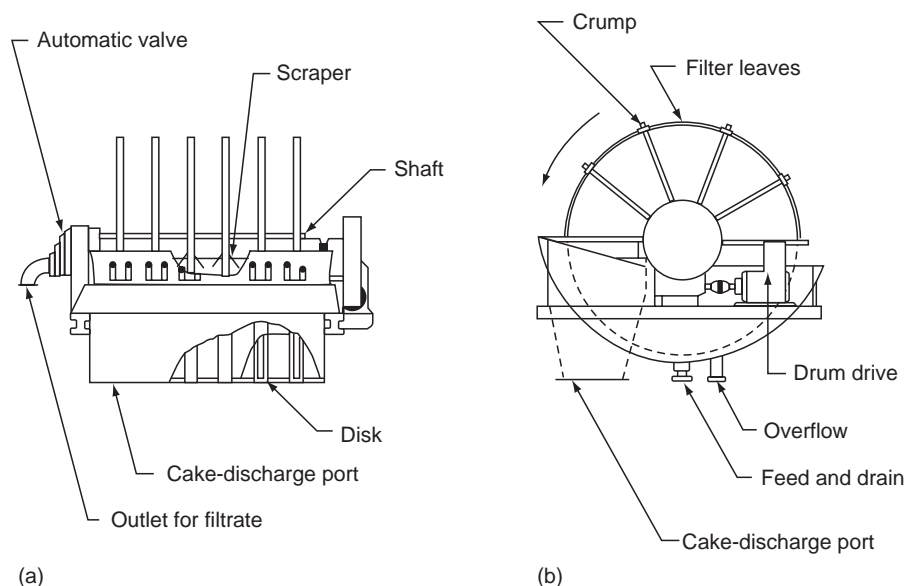


FIGURE 50.15 Vacuum disk filter: (a) side view, (b) top view. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423.)

During the filter cake-formation period, the filtrate passes radially outward through the filter medium and the bowl. Cake removal from the bowl may take place continuously or batchwise; in the latter case, slurry feed to the centrifuge is terminated during the cake removal. The theory of centrifugal filtration is well summarized in the literature [3–5].

50.5.1 BATCH-DISCHARGE CENTRIFUGES

Batchwise centrifuges have retained a considerable share of the market because of their operational flexibility, which permits qualitatively effective process control. In continuous centrifuges, relatively limited and only slightly variable times are available for main filtration, product washing, and dry centrifuging. All operations are carried out at a constant drum speed. In batch centrifuges, the duration and the drum speed can be adjusted for each operation. This facilitates the qualitative optimization of the processed products and adjustment to varying product parameters.

The peeler centrifuge is typical of the automatic batch-operated machine. Each batch is subdivided into the necessary operations: filling, centrifuging, washing, spin drying, and peeling. The horizontal peeler centrifuge, as shown in Figure 50.17, is designed to discharge solids at any speed. The solids slide down a chute. In most cases, discharge occurs at the same speed as used for final cake drying so that energy and time losses, which occur when the speed is drastically reduced, are avoided. Horizontal peeler centrifuges are designed for large throughput capacities and are widely used in the chemical and raw material industries.

The increased capacity of peeler centrifuges has been obtained by the use of a siphonic action to enhance the effective pressure differentials, as shown

in Figure 50.18. Here the filtrate is not centrifuged through the drum perforations immediately after penetrating the filter cake, as occurs in standard peeler centrifuges, but collects behind the filter cloth. This chamber is connected to a ring cup that behaves like a siphon, through boreholes, and is gas tight. The filtrate flows through the holes in the drum base into the ring cup by a hinged peeler pipe and is discharged from the ring cup. The liquid level can be adjusted at random by means of the peeler pipe, as shown in Figure 50.19. If the liquid level in the ring cup has a greater radius than the filter medium in the drum, then suction, caused by level differences and the effective centrifugal field on the back of the filter medium, occurs and further assists filtration (Figure 50.19c). This suction can reach its limit very quickly, which is the vapor pressure of the filtrate; with a level differential of 2 cm and centrifugal acceleration of 500 g, a suction effect of 10-m column of water is achieved, which considerably accelerates filtration. Conversely, a retarding effect can be achieved by setting the liquid level of the ring cup radially within the filter medium (Figure 50.19b). In this way, filtration can be delayed during the filling stage to prevent imbalances, or else the contact time of a wash or extraction medium can be extended. The rotary siphon is also used for backwashing the residual layer that remains after peeling and which is thus regenerated.

50.5.2 CONTINUOUS-DISCHARGE MACHINES

Typical example of continuous, solids-discharge machines are shown in Figure 50.20, which depicts pusher centrifuges. Both single-stage and multi-stage rotors may be employed. In these units, the slurry is

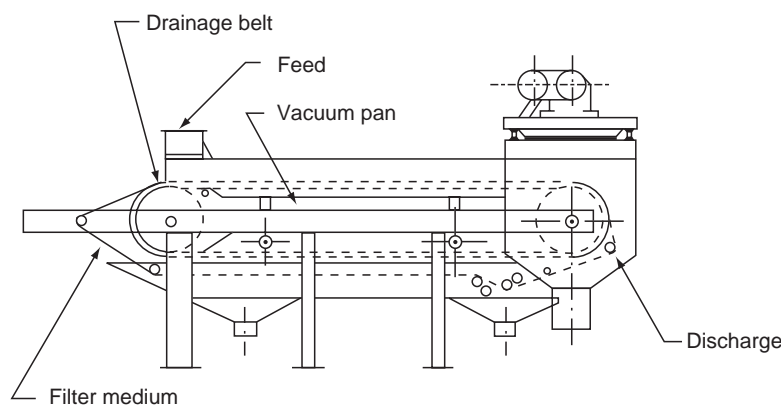


FIGURE 50.16 Horizontal belt filter. (From M. Shirato, T. Murase, E. Iritani, F.M. Tiller, and A.F. Alciatore, in M.J. Matteson and C. Orr, Eds., *Filtration—Principles and Practice*, Marcel Dekker, New York, 1987, pp. 299–423. With permission.)

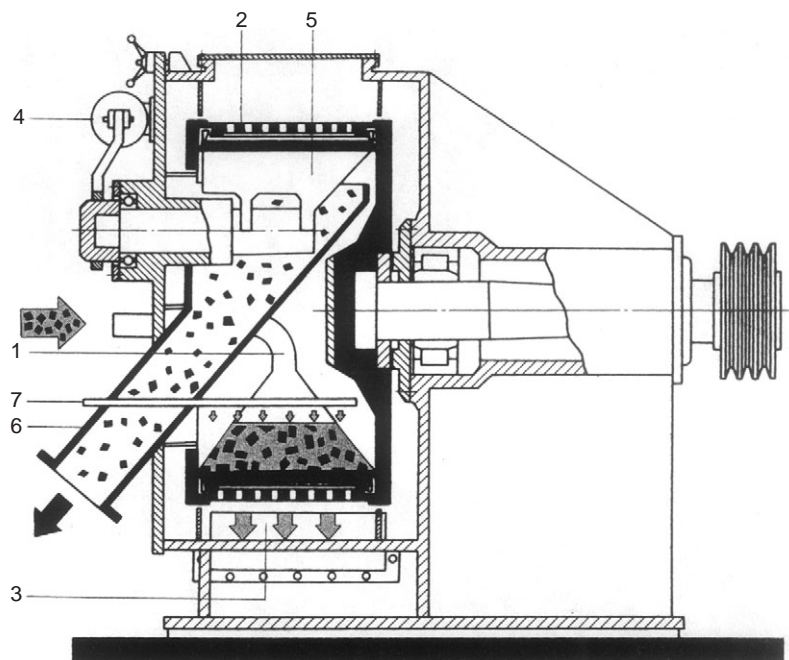


FIGURE 50.17 Horizontal peeler centrifuge (1, feed pipe; 2, slotted-screen basket; 3, liquid outlet; 4, peeling knife drive; 5, peeling knife; 6, solids chute; 7, wash pipe). (From Mitsubishi Kakoki Kaisha, Ltd. licensed by Kracess Maffei Process Technology AG. With permission.)

directed toward the back of the rotating bowl, which contains a rotating and reciprocating pusher plate. During the backstroke of the latter, filtration occurs in the space created behind the cake; the forward

stroke then causes the cake to be pushed toward the outer, lipless edges of the bowl. The cake progresses toward the outlet in a series of steps; each step distinguished by an annular band or ring of cake. Feed

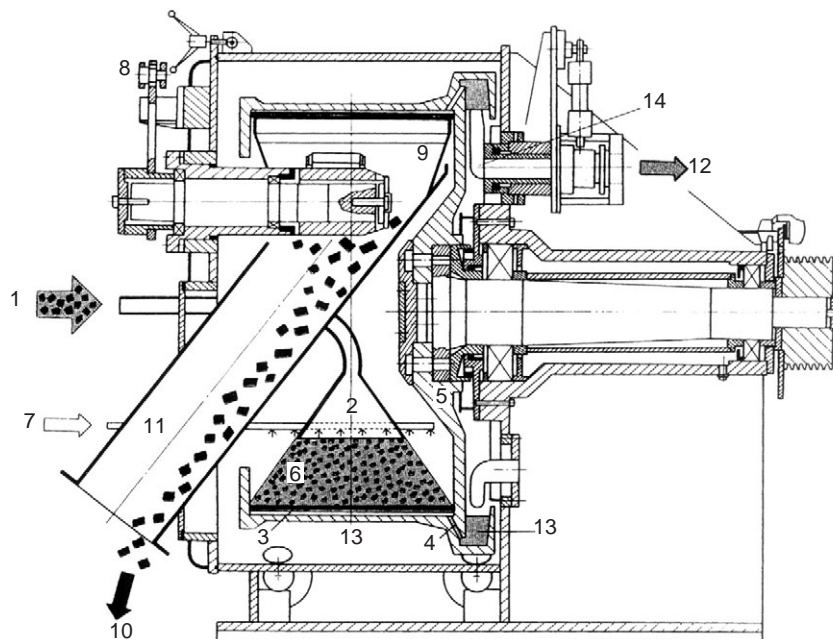


FIGURE 50.18 Peeler centrifuge with rotary siphon (1, slurry; 2, feed pipe; 3, cake; 4, siphon holes; 5, basket; 6, slurry; 7, wash pipe; 8, peeling device; 9, peeling knife; 10, solid; 11, solids chute; 12, filtrate; 13, liquid layer; 14, siphon pipe). (From Mitsubishi Kakoki Kaisha, Ltd. licensed by Kracess Maffei Process Technology AG. With permission.)

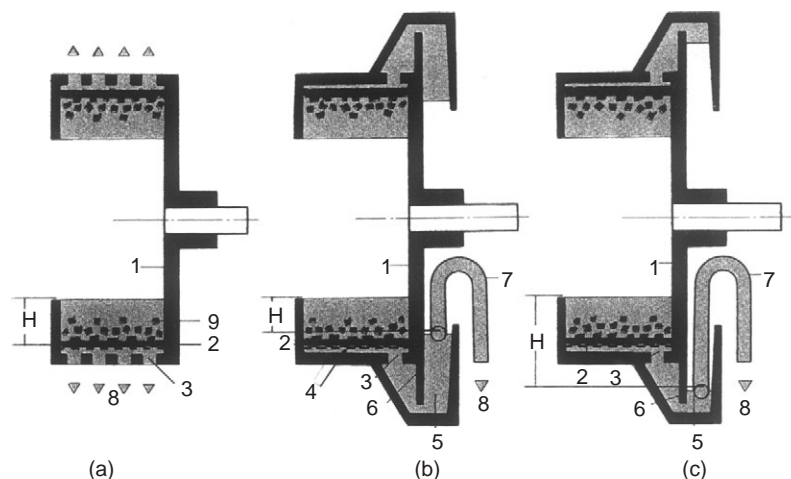


FIGURE 50.19 Mechanism of siphonic action: (a) standard centrifuge, (b) negative siphon, (c) positive siphon (1, basket; 2, medium; 3, hole; 4, filtrate chamber; 5, siphon chamber; 6, partition plate; 7, siphon pipe; 8, filtrate; 9, cake). (From Mitsubishi Kakoki Kaisha, Ltd. Catalog 7-1-1 02-01B-P. With permission.)

rates have to be controlled in order to prevent the flow of unfiltered slurry over the surface of the filter cake, in which event the surface of the cake becomes scored with deep rivulets.

50.6 EXPRESSION EQUIPMENT

Expression is the separation of a liquid from a two-phase solid–liquid system by compression, due to movement of the retaining wall rather than the pumping of the solid–liquid system into a fixed chamber as in filtration. In filtration, the original mixture is sufficiently fluid to be pumpable; in expression, the material may appear either entirely semisolid or slurry. Based on an expression theory [6,7], the time for expression is also proportional to the square of cake thickness.

Various types of equipments are available. The principal batch presses are the box, platen, pot, curb, cage [5], and tube press. A filter press with a compression device is widely used in various fields. For continuous operation, the screw press, the roller mills, and the belt press are in common use. In connection with drying operations, the following devices are explained.

50.6.1 FILTER PRESS WITH COMPRESSION DEVICE

The filter presses shown in Figure 50.10 through Figure 50.12 are all recessed plate–type units having a diaphragm fitted inside the filter plate. The diaphragm plate shown in Figure 50.10 consists of a solid body plate onto which the flexible diaphragms are heat-laminated. Thus, each diaphragm section acts as a single element in assembly. The surface of the diaphragm is provided with drainage pipes, which

lend support to the filter medium and facilitate movement of the filtrate and washing during processing. In general, the separation cycle of a filter press with a compression device consists of a first-step filtration under a relatively low filtration pressure, followed by a second-step filtration and filter cake consolidation, both under a high expression pressure. The optimum time for filtration and consolidation can be determined by the constant-pressure expression theory [6,7]. At a predetermined point in the separation, the feed to the system is interrupted and the diaphragm is inflated by pumping compressed air (or water) into the cavity between the membrane and the plate. The action of the diaphragm plate in filtration and squeezing periods is shown in Figure 50.10. Expression pressures presently applied are between 0.6 and 2.5 MPa.

50.6.2 SQUEEZER WITH MOVABLE PLATE

Figure 50.21 shows a simple filter-bag squeezer, in which the bag filled with slurry is squeezed between two plates by the thrust of a hydraulic cylinder. The separation process proceeds in six steps: (1) setting the movable plate in a fixed position, (2) fastening the bottom end of the filter bag, (3) filling the filter bag with slurry, (4) closing the top end of the filter bag, (5) squeezing the slurry by the movable plate, and (6) discharging the cake. This machine is good for applications in which cleanliness is crucial, as in food industries, especially in the extraction of natural seasoning from fish powder or shiitake mushrooms and in squeezing the bean paste for use in Japanese confectioneries.

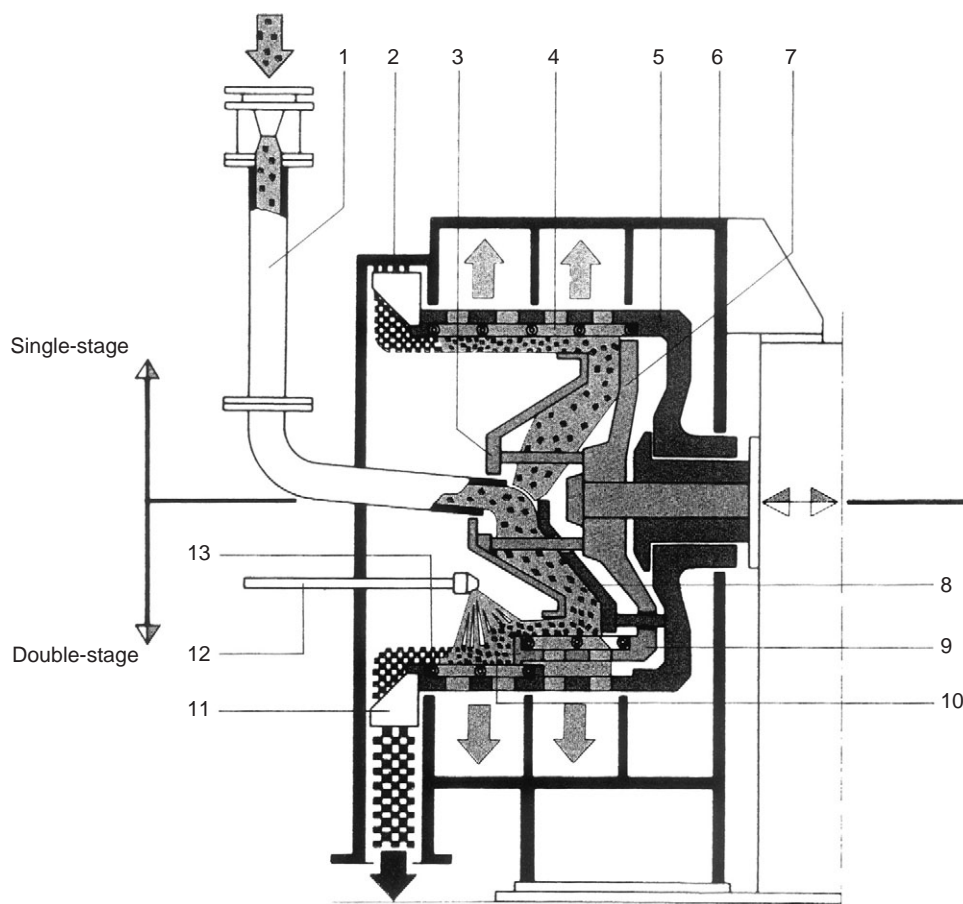


FIGURE 50.20 Pusher centrifuge (1, feed pipe; 2, solids collector; 3, feed cone; 4, slotted screen; 5, basket; 6, filtrate collector; 7, pusher plate; 8, pusher plate; 9, basket; 10, pusher ring; 11, scraper; 12, wash pipe; 13, basket). (From Mitsubishi Kakoki Kaisha, Ltd. licensed by Kracess Maffei Process Technology AG. With permission.)

50.7 CONCLUSION

This chapter summarizes the solid-liquid separation operations commonly used for the pretreatment of drying operations. We focus on the practical aspects of cake filtration, centrifugal filtration, and mechanical

expression. The choice of equipment depends on the objective of the separation, the properties of the slurry, and the scale of production. The details of solid-liquid separation theories are omitted. The reader is referred to the references for further information.

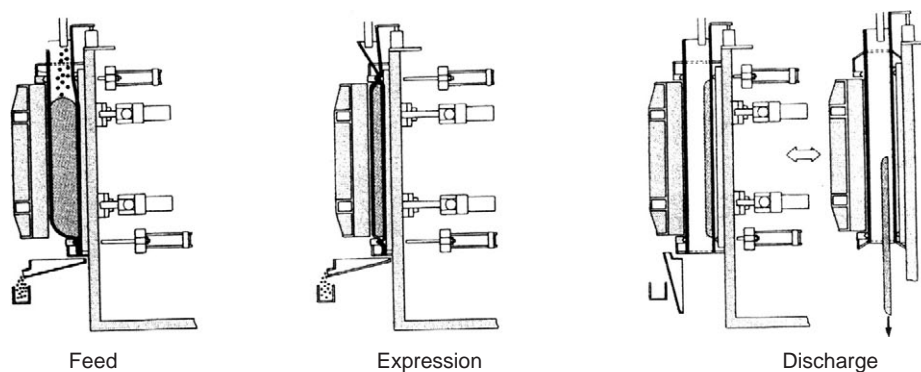


FIGURE 50.21 Squeezer with movable plate. (From Kurita Machinery Mfg. Co. With permission.)

NOMENCLATURE

A	filter area, m^2
c	the mass of dry cake solids per unit volume of filtrate, kg/m^3
K	Ruth's constant-pressure filtration coefficient, m^2/s
m	ratio of wet to dry cake mass
n	compressibility coefficient in Equation 50.2
p	applied pressure, Pa
R_m	medium resistance, m^{-1}
s	mass fraction of solid in original mixture
V	volume of filtrate, m^3
W	mass of dry cake, kg
α	specific filtration resistance, m/kg
α_0	empirical constant in Equation 50.2, $\text{m}^{1+n}\text{s}^{2n}/\text{kg}^{1+n}$
θ	filtration time, s
θ_d	filtration downtime, s
μ	viscosity of liquid, $\text{Pa} \cdot \text{s}$
ρ	density of liquid, kg/m^3

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51 Industrial Crystallization

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51.1 INTRODUCTION TO CRYSTALLIZATION

Crystallization from solution is known to produce particles of high purity and of an approximately uniform size. The major fields of application are in the chemical, pharmaceutical, and food industries. However, in some cases, difficulties in separation and drying of the crystals turn into the obstacles of an otherwise ideal method.

This chapter is an overview of industrial crystallization from the point of view of facilitating the downstream processes for product-quality improve-

ment. Some aspects of process considerations, equipment as well as the advantages and limitations of the technique are also discussed.

Over the past decades, many handbooks on crystallization have been published, such as Mullin's *Crystallization* [1], Randolph and Larson's *Theory of Particulate Processes* [2], the *Crystallization Technology Handbook* edited by Mersmann [3], the *Handbook of Industrial Crystallization* edited by Myerson [4], Tavare's *Industrial Crystallization* [5], Davey and Garside's *From Molecules to Crystallizers* [6], and Söhlneel and Garside's *Precipitation: Basic Principles and Industrial Applications* [7]. Each of these works

contains an introduction to the fundamentals of crystallization.

The easy dewatering and drying of crystals requires that the crystals be of a uniform and sufficiently large size. This is, therefore, the major theme of this chapter.

51.1.1 SOLUBILITY AND SUPERSATURATION

The amount of solute required to obtain a saturated solution under given conditions is called the solubility. The variation of solubility according to temperature, pH level, pressure, and the presence of other chemicals is the basis for the design of a crystallization process. Moreover, a phase diagram provides useful data for designing the crystallization process for a multicomponent system.

The traditional approximation is to consider supersaturation as the driving force behind crystallization and precipitation. In fact, the true driving force is the chemical potential. In reaction crystallization, in particular, this approach is important. However, we will use the simplified concept of supersaturation in the following discussion. Supersaturation is defined as the difference between the actual and equilibrium, and is expressed as follows:

$$\Delta c = c - c^* \quad (51.1)$$

$$S = \frac{c}{c^*} \quad \text{the supersaturation ratio, } S > 1 \quad (51.2)$$

$$\sigma = \frac{c - c^*}{c^*} \quad \text{relative supersaturation} \quad (51.3)$$

Supersaturation can be brought about in various ways, some of which are listed below:

1. By the evaporation of the solvent
2. From a change in the temperature of the solution
3. By a chemical reaction
4. Through the addition of an antisolvent or supercritical CO₂, which decreases the solubility of the solute
5. With the use of high pressure up to 8000 bars

Supersaturation is the main parameter in determining the rate of crystal growth and nucleation. In order to control the properties of a crystalline product, it is essential that the supersaturation level be controlled by desupersaturation rate, i.e., by the release of supersaturation which is mainly determined by the crystal growth rate and the available crystal surface area. Furthermore, the level of supersaturation is the main parameter that determines the agglomeration, shape, and purity of the crystals.

51.1.2 METASTABILITY

In the metastable zone the concentration is higher than in saturation but below the one where spontaneous nucleation occurs. Therefore, the crystal growth is mainly occurring in the metastable zone. Figure 51.1 schematically shows the solubility and supersaturation diagram. Primary nucleation takes place when the solution does not contain any crystals. Primary nucleation can be divided into homogeneous nucleation, in the absence of other substances, and heterogeneous nucleation, in the presence of other substances. Secondary nucleation is defined as the generation of nuclei with the presence of crystals. Then the nuclei are crystal fragments or added seed crystals, which grow under supersaturated conditions. The width of the metastable zone influences the nucleation mechanism and may depend on the operating conditions such as the mixing conditions, solution composition, dust, dirt, and cooling rate. In secondary nucleation, where crystallization starts in the presence of crystals and the existing crystals catalyze nucleation, the crystal surface, shape, and size of the seeds may also be important factors for nucleation kinetics. The metastable zone is, in this case, more narrow, i.e., new nuclei form easier.

In general, in order to obtain a narrow crystal size distribution (CSD), undesired nucleation should be avoided. In batch crystallization, the use of an optimal quantity of seed crystals of an optimal size may be the way to obtain a narrow CSD. According to Kohl et al. [8], the metastable zone of organic compound systems can be quite wide. Therefore, pri-

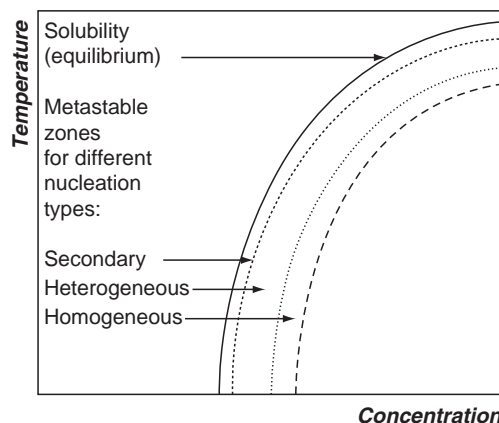


FIGURE 51.1 A schematic presentation of the metastable zone in primary and secondary nucleation. The width of the metastable zone, which is the concentration difference between saturated and operational concentration, shows the maximum degree of supersaturation where spontaneous nucleation starts.

mary crystallization occurs only at a very high level of supersaturation.

51.1.3 MEAN CRYSTAL SIZE AND CRYSTAL SIZE DISTRIBUTION

Normally, a crystalline product does not consist of single-size crystals, but consists of a wide distribution of sizes instead. The CSD or other statistical parameters, such as the mean crystal size, are used for describing the crystalline product. There are several mathematical models that can be used for describing the CSD produced by industrial crystallizers. The particle size distribution of the product can be determined by particle size analyzers.

The prediction of CSDs is still quite difficult; hence, data on crystallization kinetics are required, which means that the rate of nucleation should be known. The rate of nucleation rate can usually be obtained only to a sufficient accuracy by using empirical methods.

Generally, the “particle size” is used to denote the mean size of the crystals. However, several other terms have been used, such as the median size and the dominant size of crystals, as discussed in Allen’s book [9] where particle characterization is discussed in detail.

Particle size and the particle size distribution are important characteristics for dewatering and drying. For example, the smaller the size of the crystals, the more difficult the dewatering and drying. The effect of a wide distribution of crystal sizes is the same. Especially in precipitation, where agglomerates are formed, the porosity and roughness of the crystal surface may also affect dewatering and drying.

The population balance equation is the central equation with respect to CSD. A simple population balance is introduced by Equation 51.6. In a population balance, each crystal size is treated separately, and the growth rate in terms of an increase in the diameter of a crystal in each size class is calculated by the mass balance.

Besides agglomeration and attrition, the rates of nucleation and growth are the main kinetic parameters that determine the size distribution. These parameters, which are required in crystallizer design and simulation, can be determined on the basis of the population balance equations. Section 51.2 discusses the methods for controlling the particle size and particle size distribution.

51.1.4 CRYSTAL SHAPE

Crystalline material is usually nonspherical, except for agglomerates and larger crystals that have been

partly ground around the edges due to longer residence times under conditions where collisions occur between crystals, the agitator, and the crystallizer wall. Organic compounds, such as pharmaceutical compounds, usually have a needle-shaped crystal morphology.

The control of crystal morphology is a very complex process. This may be accomplished by the choice of an appropriate solvent system or by adding a chemical into the solution. Two methods may be used. Firstly, the experimental method can provide reliable results, but requires an excessive amount of work and extensive experience in order to be used for the testing of various solvents and additives. Secondly, computerized molecular modeling can be used in the search for suitable solvents and additives. Molecular modeling is neither very useful nor straightforward in addressing the solvent-selection problems. Molecular modeling may provide some insights, although it is still rather difficult and time-consuming to understand the whole problem of the solvent–crystal interface. There are various commercial computer applications for molecular modeling. For example, Accelrys Software Inc. [10] supplies commercial applications, named Cerius² and Materials Studio, for the modeling of the crystalline and molecular structures of organic compounds.

Programs such as C²-Polymorphs, C²-Morphology, and C²-Crystal Builder are included in Cerius². Cerius² is able to predict how impurities, tailor-made additives, the crystal growth rate, or changes in the solvent affect the structure of the forming crystal. Two prediction techniques based on the Bravais–Friedel–Donnay–Harker method and the attachment energy model are available in Cerius² that enables one to model both organic and inorganic crystals from the atomic crystal structure. It should be pointed out that the prediction of the morphology of inorganic compound systems is more complicated than that of organic compounds due to the difficulty in predicting the electrochemical phenomena in electrolyte solutions. HABIT95 is another commercial application [11]. HABIT95 predicts the external crystal structure based on the knowledge of the internal structure.

Winn and Doherty [12] have developed a new technique for predicting the shape of solution-grown organic crystals. This technique was used for predicting the crystal shape of adipic acid grown from water, biphenyl from toluene, and ibuprofen from polar and nonpolar solvents. As a general rule, the crystal shape can be described as follows: the greater the surface density of molecules on a face, the stronger the lateral interactions between the molecules and the more stable and slow the crystal growth. Furthermore, HABIT95 is based on the calculation of the

interaction energy within organic crystals solely from the knowledge of the internal crystal structure. The influence of the solution and supersaturation on crystal growth is more difficult to predict, especially for solution growth. Winn and Doherty [12] have also developed a method that requires the knowledge of readily available, pure-component properties, such as the crystal structure and internal energy of the solid, and the pure-component surface free energy of the solvent.

51.1.5 CRYSTAL PURITY

Crystallization is a unit operation where the solid product, i.e., the crystal lattice, is highly pure. More than 64% of the organic binary compound systems, where crystallization has been used as a purification or concentration method, are eutectic. In eutectic binary compound systems, only one compound in a certain solution composition forms the solid material, except at the eutectic point at which both components crystallize. As a general rule for purification by crystallization, the slower the crystal growth rate, the purer the crystal. A part of the smallest crystal fraction, i.e., fragments formed by attrition, may have a higher concentration of impurities due to the higher crystal growth rate. The reason can be explained by the “healing effect.” Tiny crystals, which are fragments of larger crystals and, therefore, of an irregular shape, tend to initially obtain a more regular morphology at a relatively high growth rate. If agglomeration takes place, the secondary crystals may contain inclusions of the mother liquor, which decreases the purity of the crystalline product. An impure product is usually caused by traces of mother liquor adhered on the crystal surface in the crystal cake. This makes filtration and crystal-cake washing so important for final purity.

A better understanding of the behavior of a solution during crystallization requires a knowledge of the chemistry of the solution that may contain numerous other solutes or solvents in addition to the crystallizing compound. Firstly, these other compounds may influence the crystal growth or nucleation rate, which are the kinetic parameters that dominate suspension crystallization. Secondly, the other compounds may greatly change the solubility of the crystallizing substance to an extent that the supersaturation level in the process changes or it becomes possible for one or more compounds to salt out the main product at a certain concentration level. Thirdly, the other compounds may lead to an impure product due to inclusions that contain the mother liquor or to adhesion onto the crystal surface. Furthermore, other compounds may influence the mass transfer phenomena

of the crystallizing substance, which leads to the need for a better understanding of multicomponent diffusion [13]. Other components may adsorb onto a certain facet of the crystal surface, thereby affecting the crystal growth rate and the final crystal shape.

In the pharmaceutical industry, in addition to chemical purity, it is essential that the product has a desired polymorphic and pseudopolymorphic form. A chemical may have several crystal structures; the chemical compositions of these polymorphs are identical. This phenomenon is called polymorphism. A different lattice usually causes changes in crystal morphology. The chemical nature may be about the same from one polymorph to another, whereas the physical properties may differ greatly. Density, heat capacity, melting point, thermal conductivity, and optical activity are examples of these properties, according to Myerson [4, pp. 36–38]. The various melting points and the difference in dissolution rates from one polymorph to another are the main factors that have to be taken into account in the pharmaceutical industry. The forming of polymorphs depends on the crystal growth conditions, which are mainly affected by temperature, pressure, impurity content, growth rate, etc. During the crystallization process, the polymorph that forms first is often crystallized under metastable conditions. This metastable polymorph often transforms into a stable polymorphic form, which is known as polymorphic transition. It is essential to avoid the formation of an undesired polymorph or the desired polymorph should be prevented from transforming into another undesired polymorphic form. This is not easy because a small change in temperature may cause a polymorphic transition, which can occur quite often in suspensions. However, some materials may change into a polymorphic form in the dry, powder form. Seeding with the desired polymorph is especially important, according to Ulrich [14], when the conditions are close to the transition point. Two polymorphs can remain stable close to or at the transition point. Mullin [1] has introduced the principle of how to measure the solubilities for metastable and stable polymorphs. He has also shown how the transformations of various polymorphs can be studied at various temperatures and pressures. The reversible transformation from one polymorph to another is called enantiotropy, whereas the irreversible transformation is known as monotropy. Pseudopolymorphism occurs when a solvent enters the crystal lattice and forms, for instance, a hydrate product in aqueous solutions and a solvate in other solvents, as presented by Ulrich [14]. The bonds forming between the solvent and the solute in the lattice are of a physical nature; they are not chemically bound.

51.2 CONTROL OF CRYSTAL SIZE DISTRIBUTION AND CRYSTAL SHAPE

This section discusses the methods for controlling the CSD and crystal shape for different batch and continuous processes.

51.2.1 CONTROL STRATEGY OF SUPERSATURATION IN BATCH PROCESSES

Several techniques, such as the programmed-cooling rate in cooling crystallization and seeding technology, have been developed for controlling the particle size in batch-cooling crystallization processes. The principle of these methods is to influence the nucleation rate.

Nucleation depends on the total surface area of the crystals. If the total crystal surface area is relatively small in relation to the degree of supersaturation, nuclei, i.e., small crystals, will be formed after the initiation of crystallization. However, if the crystal surface area is sufficiently large, the level of supersaturation will be reduced through the growth of existing crystals.

Traditionally, batch crystallization is driven by the natural cooling of the mother liquor. Natural cooling can be best understood as a process where the crystallizer is left to cool down by natural convection of heat into the surrounding air. A faster variation of essentially the same method is to have a jacketed tank, where the jacket is thermostated to the end temperature of the process as desired by the user. In both cases, the temperature initially falls rapidly, after which it then slowly approaches the wall temperature of the tank. One problem of natural cooling is incrustation, i.e., a crystal layer forms on the cooling surface. However, the major disadvantage of natural cooling is that, for some solution systems, it becomes difficult to reliably run the process due to a high level of supersaturation in the initial period. If the total crystal surface area of the seed crystals is insufficient, a large number of very small crystals will be generated. In some systems, the growth rate of small crystals is slower than that of large crystals, and small crystals grow at a slow rate for decreasing temperatures. Next, supersaturation increases again and more tiny crystals are formed; the final crystal size is small due to the large number of nuclei that are generated in the process. In order to increase the particle size, the cooling rate of the solution can be programmed as follows.

At the first stage, when nucleation starts, the cooling rate should be slower. The second stage is faster when the existing crystals grow. According to Mullin [1, p. 355], an adequate cooling profile

for a general application, when seeding is used, is shown below:

$$T = T_0 - (T_0 - T_f)(t/\tau)^3 \quad (51.4)$$

Mayrhofer and Nývlt [15] derived the following expression for the optimal temperature profile of a batch crystallizer for both seeded and unseeded processes:

$$\frac{T_0 - T}{T_0 - T_f} = \left[1 - K\left(1 - \frac{t}{\tau}\right)\right] \left(\frac{t}{\tau}\right)^3 \quad (51.5)$$

The slope of solubility as a function of temperature is assumed to be constant in this derivation.

On an industrial scale, the maximum cooling rate is often limited by the heat-transfer capacity of the equipment, which means that the theoretical cooling program cannot be followed during the final stage of the batch. A high suspension density may cause high secondary nucleation and, thus, decrease the crystal size of the product.

Industrial experience shows that the above method may produce larger crystals than by natural cooling. The obtainable size depends on the system, though an increase in crystal size due to programmed cooling can be expected. Obviously, programmed cooling should first be tested with bench-scale equipment before investing in a programmed-cooling system.

According to Mersmann [16], the supersaturation in cooling crystallization is usually small, i.e., relative supersaturation, σ , is less than 0.1 and the relative solubility expressed by the ratio of solubility (kg/m^3) and crystal density is usually higher than 0.01. As a result, primary nucleation does not take place and the nuclei are formed only as attrition fragments. If a coarse product is desired, the attrition rate should be low and the crystal growth rate at the maximum allowable level with respect to crystal purity, all of which depends mainly on the mean residence time of the slurry.

51.2.2 SEEDING

Seeding is commonly used as a control technique, especially in batch crystallization processes. The nucleation rate depends on the total surface area of the crystals. If the total crystal surface is relatively small in relation to the level of supersaturation, small crystals or crystals of undesired shapes may be formed in the early stage of crystallization. Otherwise, if the crystal surface area is sufficient, supersaturation will cause the existing crystals to grow.

Heffels and Kind [17] have reviewed the general principles of seeding in a batch and continuous solution crystallization as well as in melt crystallization. They have provided practical ideas on how to carry out seeding with various crystallization systems. For instance, in practice, 0.1 to 0.3 times the final size and 0.1 to 3% of the final mass of the seed area are good choices for the seed size and mass, respectively. Usually, seed crystals of 20 to 50 μm are appropriate for the production of final crystals of 500 to 1000 μm . However, for a final product size of less than 100 μm , a seed size of 1 to 10 μm is more appropriate. For biotechnological and pharmaceutical processes, e.g., the production of insulin, the seeds should be 1 to 2 μm in size and obtained by freeze-drying. Seed sizes below 10 μm are not recommended because it is difficult to create a mass with a narrow CSD for such small crystals. Furthermore, small crystals tend to dissolve due to the Ostwald ripening effect and the crystal growth rates of tiny crystals are lower than that of larger ones. In addition, tiny crystals may form agglomerates and stick onto other crystals. Therefore, the introduction of crystalline dust must be avoided. The seed crystals should be washed or slightly dissolved to remove adherent crystalline dust, especially when dry crystals are used as the seeds. The addition of dry seeds may cause the seed crystals to float or agglomerate if they remain floating on the surface. It has been observed that ground seed crystals grow faster than seed crystals that have a smooth surface, though the reason is not clear. Therefore, milled seeds are used in the sucrose industry, for instance. Seeding with impure seed crystals may decrease the crystal purity and the used seeds should, therefore, be sufficiently pure. With compound systems of stereo chiral isomers, optical isomers or enantiomers, for instance, the type of seed crystal usually dictates the crystalline product that is formed. In this way, seeding affects the selectivity of crystallization.

The addition of the seed slurry can be done at different stages: onto the liquid surface (the crystals tend to agglomerate if they are not dispersed); near the stirrer, where the liquid is well mixed; or by one of the feed streams in a bypass or loop in the crystallizer. The best way is to add seeds to the solution beneath the liquid level or to add them to a feed stream at the beginning of the batch or, in case of a semibatch system, continuously. In reactive batch crystallization, the nucleation rate can be decreased by the dilution of the feed streams or by the addition of fine crystals into one of the feed solutions or the reactor. The surface area and growth rate should be large enough to quickly reduce the supersaturation. The recirculation of the crystal suspension creates an

effect similar to seeding in the growth regions of the reactor. In some cases, large particles are produced by agglomeration, which is triggered by electrostatic forces. These forces can be measured with a zeta potentiometer. The simplest method to prepare the seed crystals is with the use of the final product of the previous production batch. In addition, the milling and dispersion of the seeds in inert solvents or saturated solutions are the generally used methods.

In some processes, the residual suspension droplets at the stirrer and wall are able to capture enough crystals from the previous batch, and these crystals function as seeds for the next batch. Some researchers have postulated that the growth of nuclei in secondary nucleation dominates the final CSD. However, there are reports that show that seeding does not significantly increase the crystal size. Heffels and Kind have, nevertheless, shown that seeding improves the quality of the crystalline product in many cases and has a clear advantage.

Kohl et al. [8] studied seeding in batch crystallization. The system studied was an aqueous β -cyclodextrin solution that had a wide metastable zone. They determined a critical size for the seed crystals in this system. When the size of the seeds used was larger than the critical size, secondary nucleation took place.

Furthermore, Jagadesh et al. [18,19] have studied seeded batch-cooling crystallization without temperature control for aqueous potassium alum, and potassium sulphate solutions. In their approach, the solution was cooled according to a natural profile. They succeeded in attaining a monodispersed and relatively narrow CSD by controlling the level of supersaturation in such a way that the existing seed crystals grew and no nucleation occurred. This controlling method was based on the use of small, ground, seed crystals of diameter below 50 μm . The required mass of the seeds was less than 1% of the total crystal mass in the product. Hence, the main idea was to add small-sized seeds in order for no new nuclei to be formed during the batch process. However, for some compound systems, secondary nucleation cannot always be fully avoided at usual suspension densities.

51.2.3 PRECIPITATION

Precipitation technology is not yet very advanced, and many industrial companies experience problems with small crystals that cannot be separated easily. Furthermore, for some purposes, it is necessary to produce very small crystals (e.g., pigments, paper-coating materials, and paper fillers) and, therefore, solid-liquid separation becomes inevitably difficult. In the paper industry, submicron-sized paper-coating agents may be produced in a process in connection to

the paper mill so that the suspension is transported as slurry and no filtration is required. If the filtration of small crystals is needed, the problem is the need to be handled by filtration specialists.

Precipitation is an operation known for producing small crystals that are difficult to filter and dry. Batch precipitation is usually carried out in the form of a semibatch process, i.e., one or two reactants are continuously added to the reactor. The control of the particle size distribution in a precipitation process is very complicated because of the high level of supersaturation generated by the fast reaction. According to Mersmann [16], the important factors for sparingly soluble systems in isothermal precipitation are:

- The concentration of reactants
- The rate of the chemical reaction
- The intensity of macro- and micromixing
- The dilution of solution
- Agglomeration

If nanometer-scale crystals are required, a tiny crystal size can be obtained by applying very high primary nucleation rates, which require extremely strong supersaturations. Such high levels of supersaturation can be achieved through

1. The high concentration of reactants
2. The avoidance of agglomeration (using surfactants, pH control, etc.)
3. Rapid quenching or diluting in order to halt growth (a combination of a T-mixer and a stirred vessel, Figure 51.2)

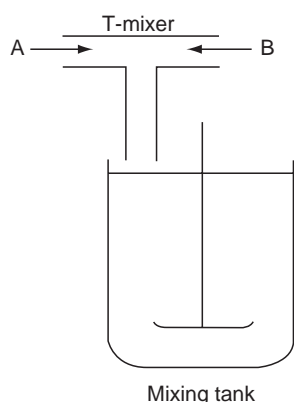


FIGURE 51.2 The operation of a T-mixer and a stirred vessel. The reactants are mixed in a T-mixer where the chemical reaction takes place. The suspension is then introduced into the vessel operating at a relatively low level of supersaturation.

4. No dilution (T-mixer) in order to produce many nuclei
5. The rapid local micromixing and poor macro-mixing of the reactants
6. Products with high concentrations and low solubilities
7. Rapid micromixing at the feed point of the reactants for a fast chemical reaction

On the other hand, if a coarse product is required, supersaturation must be kept at a low level through

1. The use of low-concentration reactants (perhaps by dilution)
2. Good macromixing in the entire precipitator
3. Seeding by recycling the slurry at high rates in order to reduce local supersaturation peaks while keeping primary nucleation at a minimum and promoting crystal growth

Kind [20] has discussed the basic factors that affect reactive crystallization. The first crystals that appear are micrometer or even nanometer in size range. Interfacial forces govern the behavior of suspended particles in these size ranges, as a result of which the fine particles are affected by colloidal interaction. Therefore, the final product particles are often secondary, formed by the aggregation of small primary particles. The final product morphology results from the action of aggregation, the rheological behavior of the suspensions, and the shear rates present in the precipitator. The primary processes are mixing on the macro, micro, and molecular scale, as well as the reaction, nucleation, and growth of the particles. The formation of primary crystals depends mainly on the mixing conditions. If mixing does not control reactive crystallization, the crystallization reaction can be best controlled by initial supersaturation. Under ideal mixing conditions, the final primary particles in the nanometer size range are rapidly formed within a few milliseconds and remain unchanged, except in the case of long-term ripening and aging. Aggregation, aging, and ripening are secondary processes. At moderate shear rates, the secondary particles can be found in the size range of 5 to 20 μm . It is relatively difficult to scale-up precipitation. Generally, the design of precipitation processes still requires experience from the behavior of each system.

51.2.4 CONTINUOUS PROCESSES

Continuous crystallizers are used for the separation of large quantities of fertilizers, salts, and sugars. The common feature of crystallization is the existence of a

driving force for crystallization, i.e., supersaturation. The method by which supersaturation is generated and the various factors affecting its local and average values have a strong influence on all the properties of the crystallized material. It may be noted that the local and average values of supersaturation affect the growth, nucleation, and aggregation rates, which in turn determine the population density and CSD. The CSD, on the other hand, determines the specific surface area and magma density, which strongly affect the growth and aggregation rates, the rate of secondary nucleation, and the solute mass balance in a feedback fashion. Irrespective of the mechanism of the generation of supersaturation, impurities exert a strong effect on the width of the metastable zone. Impurities also influence crystallization kinetics, crystal purity, crystal morphology, and the aggregation rate. Optimum supersaturation, at which the existing crystal growth is at the maximum rate permitted (i.e., the desired particle shape is obtained without mother-liquor inclusions) and at which a certain crystal number is not exceeded, must be determined experimentally.

The objectives in the operation of a crystallization process are to meet the product specification, i.e., a narrow CSD, maximum crystal purity, high yield, and an acceptable crystal morphology. Moreover, the manufacturer's requirements for economic and trouble-free operation should be met. The CSD and crystal morphology are important parameters for downstream operations of crystallization such as filtration and drying.

Any attempt to control a crystallization process to meet the above-mentioned product specifications should be directed toward generating and maintaining a mild and homogenous supersaturation in the bulk of the magma at local and average levels. This favors the growth rate and prevents spontaneous nucleation. In industrial crystallizers, this objective is achieved by improving the design of the crystallizers as well as by controlling some of the easily measurable process variables, such as temperature, pressure, solution level, and flow rates. Such an approach is not sensitive to frequently occurring disturbances such as variations in the feed composition, feed temperature, impurities, and differences in the extent to which mixing is carried out in the tank. Normally, continuous crystallization is carried out in a mixing tank. Internal or external circulation is used to reduce the level of supersaturation.

When the crystallizer is well mixed and operated continuously at steady state, with no attrition and breakage of the crystals at a size-independent crystal growth rate, the simplest population balance equation is

$$G \frac{\partial n}{\partial L} + \frac{n}{\tau} = 0 \quad (51.6)$$

The equation describes how, during the retention time, the crystals move to a higher-size class because of growth. The derivation of the above equation is simple and can be found in textbooks. The particle size distribution is obtained from the above equation and is shown below:

$$n = n_0 \exp\left(-\frac{L}{G\tau}\right) \quad (51.7)$$

The median crystal size, L_{50} , which occurs at a cumulative mass undersize of 0.5, can be evaluated by the following equation:

$$L_{50} = 3.67G\tau \quad (51.8)$$

The maximum of the mass distribution density or the dominant size, L_d , can be calculated using the following equation:

$$L_d = 3G\tau \quad (51.9)$$

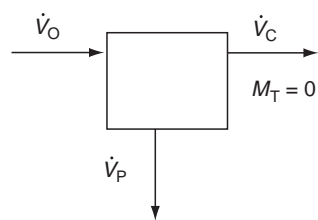
This is a very useful equation for the design of a crystallizer and can be used to obtain a rough estimation of the particle size in industrial processes.

Generally speaking, the median crystal size of a crystalline product can be increased through a decrease in the number of nuclei or an increase in the growth period of the crystals present in a crystallizer operated at an optimum level of supersaturation. Of these, the most practical way to alter the particle size distribution in a continuous crystallization process is to adjust the residence time of the crystals.

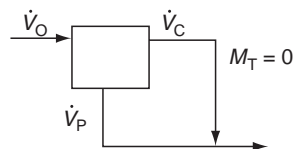
In the following discussion, we give a brief explanation of some of the principles used in controlling the crystal size, roughly following the treatment of the subject by Mersmann [3].

51.2.4.1 Clear-Liquor Advance Operation

A clear-liquor advance operation is one in which the overflow liquid is continuously removed from the tank. The overflow is not actually clear but contains small crystals that have not settled in the overflow section. This method is a simple way of controlling the CSD, because the residence times of the clear mother-liquor overflow and crystal are different. The flows are separated and the excess nuclei are removed with the overflow. The nuclei in the overflow can be used, for example, in the following crystallization unit. The larger the clear-liquor flow for a given feed flow, the longer the residence time of the crystals in the product stream.



(a)



(b)

FIGURE 51.3 Clear-liquor advance operation.

The operation of the clear-liquor advance is shown in Figure 51.3a and Figure 51.3b. The system in Figure 51.3b can also be called a double draw-off crystallizer. The liquor may be recycled separately resulting in the product stream having a high suspension density. For operation with double draw-off, as the production rate is the same as in the mixed-solution–mixed product removal (MSMPR) crystallizer for a given desupersaturation, the suspension density in the unified stream is also equal to the suspension density of the MSMPR crystallizer. In such a case, the nucleation rate of the crystals can be expressed as

$$n_0 = \frac{B_0}{G} = k_n G^{i-1} M_T^1 \quad (51.10)$$

The dominant crystal size, L_d , is proportional to the product of the crystal growth rate, G , and the residence time of the crystals

$$\frac{L_{d2}}{L_{d1}} = \left(\frac{\dot{V}_p + \dot{V}_c}{\dot{V}_p} \right)^{(i-1)/(i+3)} \quad (51.11)$$

The above expression shows that when the flow rate of the clear liquor increases, the dominant particle size also increases.

51.2.4.2 Destruction of Fines with Solute Recycling

The operation of the fines-destruction process is similar to that of clear-liquor advance; as shown in Figure 51.4, the difference is that when the clear liquor is used, the small particles return back to the product flow, thus increasing the proportion of small crystals.

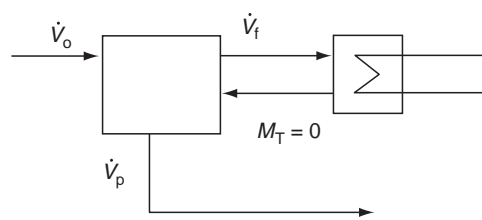


FIGURE 51.4 Fines-destruction crystallization process.

For the fines-destruction operation, very small fines can be withdrawn, and the suspension density of the overflow is very small. The number of crystals in the crystallizer decreases as a result of eliminating the small-sized particles. This operation is very useful in increasing the size of crystals in a system that has a high rate of nucleation. This method does, however, lead to a wide particle size distribution.

The advanced stream is classified at a size of L_f and all the particles with $L < L_f$ are eliminated in the dissolver. The fines are removed at a size that is negligibly small compared to that of the product-size crystals. In this case, the size-improvement ratio, L_{d2}/L_{d1} , is given by

$$\frac{L_{d2}}{L_{d1}} = \left(\frac{1}{\beta} \right)^{1/(i+3)} \quad (51.12)$$

According to this equation, size improvement is appreciable if the kinetic exponent, i , is small and the fraction β of the surviving nuclei is also small.

The efficiency of fines destruction depends on the undersaturation in the destruction system and the residence time that must exceed the dissolution time of the nuclei.

51.2.4.3 Classified Product Removal

The operating system for classified product removal is shown in Figure 51.5. The operation of classified

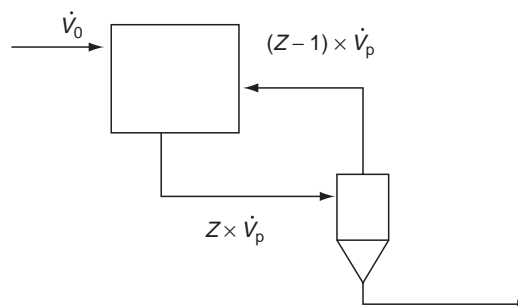


FIGURE 51.5 The classified product-removal crystallization process.

removal can be achieved by a classification device, which may be a hydrocyclone, a classifier, a wet screen, a fluidized bed (FB), or a separating centrifuge. By using a suitable classification device, only the particles of a size larger than that required can be withdrawn from the product; the crystals, the size of which is smaller than the required size, will return to the crystallizer for further growth. In this way, the CSD will become narrower, though the mean size of the crystals will decrease. The drawback of this method is that classified product removal may cause instability in the process and even force the process to be discontinued due to the excessive increase in the amount of small crystals. Therefore, in practice, classified product removal alone is not used in industrial crystallization.

In large-scale industrial crystallization processes, the most commonly used classification device is a hydrocyclone. The advantages of the hydrocyclone are its high capacity within a small equipment volume and an easily adjustable cutting size by the control of the feed-flow rate and the ratio of the up-flow to the down-flow. The fluid bed or elutriation leg is another method often used for classification in crystallization processes. With the elutriation leg, the feed or clear solution is fed from the bottom of the elutriation leg. The up-flow velocity is set based on the settling velocity of the cutting size of the particle. The crystals, the size of which is smaller than the cutting size are sent back to the crystallizer for further growth. The crystals, which are larger than the cutting size, are drawn out of the crystallizer as the product. In principle, the classifier and the separating centrifuge can both be used as classification devices. However, as it is not easy to control the actual cutting size, it is rare to use these devices as classifiers in industrial crystallization processes. These devices are mostly used in separating the solid from the solution. In the elutriation leg, described earlier, this means preventing the small crystals from entering the next separation stage. In the centrifuge, the crystals are simply separated from the crystal slurry. In some processes, the separated solution includes many small-sized crystals. If the solution is sent back to the crystallizer, the separator may be considered to be a classification device. It can be further said that the wet screen is seldom used as a classification device in industrial crystallization processes because the easy blockage of the screen will cause operation problems.

51.2.4.4 Classified Product Removal with Fines Destruction

A crystallizer equipped with the devices for product classification and fines destruction can efficiently

improve the particle size and particle size distribution. With the fines destruction, a portion of the small particles can be eliminated in such a way that the number of crystals in the crystallizer can be controlled within the required range. When this method is complemented with classified product removal, the size of the crystals in the product is controlled; on the other hand, an undesired increase in the amount of small crystals is prevented. With this method, it is possible to obtain crystals of a uniform and large size.

51.2.4.5 Role of Mixing Intensity in Continuous Suspension Crystallization

Industrial crystallization is normally carried out in a mixing tank. Many parameters are affected by the mixing conditions during crystallization. Firstly, mixing affects the temperature or concentration distribution in the mixing tank in such a way that the driving force for crystallization is related to the mixing degree in the tank. If mixing is not complete, there may be a very high local level of supersaturation, which will cause the crystallization process to become uncontrolled. In addition, it is impossible to reach the required product size, morphology, and purity. Secondly, the mixing intensity affects the mass transfer rate between the solid and the solution, and variations in crystal growth rate occur. It is well known that the secondary nucleation rate is strongly affected by the mixing intensity because of the crystal collisions with the impeller, between the crystals, and of the crystals against the wall. The effect of mixing on crystallization kinetics will directly affect the particle size distribution usually in an undesirable way.

Besides the mixing effect on the basic process of crystallization, another important effect of mixing on continuous crystallization is the residence time of the solution and the solid. As shown by Sha and Palo-saari [21] and Sha [22], as the particles of different sizes have different flow patterns in the crystallizer, the particle size distribution is different at different locations. The particle size distribution in the product line depends heavily on the mixing intensity and on the location from which the product is withdrawn.

51.2.5 SIMULATION OF CRYSTALLIZERS

The calculations required for predicting the operating conditions are complicated. It is necessary to have, or write, a computer simulation program. The simplest part of such a calculation is to elaborate the energy and mass balances of the system. The solubility and enthalpy data are also required. The calculation produces the amount of crystals produced in a specific

time unit, and the energy required is usually steam that is fed into the heat exchanger if the system is continuous. This calculation does not produce the crystal size and CSD, which would both be useful in the design of the equipment. Although, in practice, a large part of design work is based on the use of previous experience in crystallization of the same or similar materials, there is, nevertheless, the need for design methods that can predict the particle size.

One method is to solve the population balance equation (Equation 51.6) and to take into account the empirical expression for the nucleation rate (Equation 51.10), which is modified in such a way that the expression includes the impeller tip speed raised to an experimental power. In addition, the experimental value, pertinent to each chemical, is required for the power of the crystal growth rate in the nucleation rate. Besides, the effect of suspension density on the nucleation rate needs to be known. Furthermore, an industrial suspension crystallizer does not operate in the fully mixed state, so a simplified model, such as Equation 51.6, requires still another experimental coefficient that modifies the CSD and depends on the mixing conditions and the equipment type. If the necessary experimental data are available, the method enables the prediction of CSD and the production rate as dependent on the dimensions of the tank and on the operating conditions. One such method is that developed by Toyokura [23] and discussed and modified by Palosaari et al. [24]. However, this method deals with the crystallization tank in average and does not distinguish what happens at various locations in the tank. The more fundamental and potentially far more accurate simulation of the process can be obtained by the application of the computational fluid dynamics (CFD). It will be discussed in the following section.

51.2.6 ROLE OF COMPUTATIONAL FLUID DYNAMICS IN THE RESEARCH OF CRYSTALLIZATION

The accurate simulation of suspension crystallization presumes that the parameters that influence a crystallization process, such as the suspension density and supersaturation, are known for each location in the tank. One method of obtaining this information is the application of CFD, which is the basic tool for the study of fluid dynamics and has been applied in many areas. Recently, CFD techniques have been developed to a level that allows complex flow situations to be modeled to a reasonable degree of detail and accuracy. The application of CFD in chemical engineering processes has been reported in some fields [25–27]. Many methods have been developed for the

simulation of stirred tanks [28–30]. These studies offer a good background for the study of crystallization.

Garside [31] provided an extensive review of crystallization from solution in which he emphasized the pivotal role of fluid mechanics in the kinetics and resulting CSD. As summarized by Rielly and Marquis [32], crystallizer fluid dynamics act at a variety of scales and, to a certain extent, have some local influence on each of these kinetic steps, such as the growth and reaction rates, aggregation, breakage, supersaturation, nucleation, and residence-time distribution of the particles. On a macroscale in the order of the crystallizer dimensions, the fluid mechanics of the mean flow affect the spatial distribution of the fluid and solid phases, imposing a degree of backmixing that determines the solid and liquid residence-time distributions. Mesoscale fluid mechanics in the order of the impeller blade width affect the local liquid-concentration distributions of the feed reagents, interaction of the crystals with the impeller (crystal abrasion on impact with the impeller leads to secondary nucleation), and heat-transfer rates from crystallizer internals. At an even smaller scale in the order of the size of turbulent eddies or individual crystals, microscale fluid mechanics affect the mass transfer rates, the rates of turbulent collisions between the particles, and the micromixing of chemical reagents. The fluid mechanics of the suspension flow at these various scales affect the precipitation rates, distribution of supersaturation, nucleation rates, growth, as well as the numerous kinetic processes by which crystals of a given size are born and die. Even processes such as nucleation, which are more directly related to local supersaturation, are indirectly related to suspension fluid mechanics; supersaturation is determined by local micromixing and the reaction and mass transfer rates.

As described above, many phenomena of crystallization are related to the fluid dynamics of the crystallizer and can be studied with the use of CFD. The basic transport phenomena in the crystallization process, which include momentum, heat and mass transfer, can be easily solved using CFD. The mass transfer between the particle and the solution is the main process that describes crystal growth. Zöller et al. [33] have studied single crystal growth under the flow of solution. The nucleation rate was considered directly proportional to the energy of the moving particle, the movement of which can be obtained from CFD. In this way, the nucleation rate can be studied. Ten Cate et al. [34] suggested a model for the study of individual crystals in the flow, through which the collision frequency and energy can be obtained. The collision data obtained in this way can be used in the models that predict the rate of secondary

nucleation and attrition. The nucleation study using CFD is still at an initial stage; researchers are still looking for the best model for simulation of the nucleation of crystallization, although not much work has been published as yet. Expressions describing the mass transfer rate can be used together with the cooling and chemical reaction rates in order to obtain the concentration distribution in the solution. Then the local supersaturation in the crystallizer can be calculated with improved accuracy for the nucleation and growth rate. Falk and Schaer [35] present a probability density function (PDF) method for the calculation of evolution of the particle size distribution in precipitation reactors. This method contains a micromixing model based on a Lagrangian frame where chemical reactions are treated without modeling and which requires minimal computational resources. The micromixing model considered the concentrates on the interaction and exchange between the fluid and the mean-sized crystals and takes into account nucleation, growth, and aggregation. With this method, it is possible to obtain, for any point in the flow, the reactant concentration and supersaturation fields and the particle size distribution field, using simple moment methods. The agglomeration phenomena can also be described in the CFD environment. Hollander et al. [36] used a lattice Boltzmann scheme to simulate the turbulent flow field and a Monte Carlo algorithm to solve the particle size distribution and studied the dependence of the agglomeration rate constant on the shear rate and on turbulent transport. They concluded that the turbulent transport of particles is an important factor in the overall agglomeration rate.

Apart from the crystallization kinetics, the mixing state of the suspension in a crystallizer has a strong effect on the crystallization process. For the precipitation system, the particle size distribution in the reactor depends on the kinetics of the crystallization process, although it is also strongly influenced by mixing of the different chemical species, including solid particles. This kind of study has been carried out by many researchers [35,37–41]. The different fluid dynamics between the solid and the solution and between the different-sized crystals is the crucial information for the description of the crystallization process, especially in the case of a continuous crystallizer [21] where the concentration fields of different-sized particles are studied using multiphase flow. Here, the effect of the particle size distribution at each point in the flow on the crystallization process is studied. Studies, which have been carried out on crystallization using CFD such as those mentioned earlier, will bring new insights to the understanding of the crystallization processes.

Another objective in the study of the application of CFD in crystallization is to simulate the particle size distribution in crystallization. In order to solve this problem, the simulation should take into account the population balance. The internal coordinates of the population balance make it difficult to utilize it in the CFD environment. In addition, different-sized particles have different hydrodynamics, which causes further complications. Wei and Garside [42] used the assumption of MSMPR and the moments of population balance to avoid the above difficulties in the simulation of precipitation. In the CFX commercial application, the MUSIC model offers a method for solving the population balance equation in CFD and defines the flow velocity of different-sized particles on the basis of the ratio of the velocity to a reference-sized particle. The velocity of this reference-sized particle is then simulated using the multiphase-flow model. In this way, the problem of the different hydrodynamics of different-sized particles is solved. However, the accuracy of this model is not sufficient for it to be directly used for bubble-sized distribution only. More work is required before this model can be applied to crystallization.

For the simulation of all the crystallization processes, a detailed approach, as well as certain geometry-independent kinetic models, should be able to take the geometry of the crystallizer into account. Kramer et al. [43–44] propose the division of the crystallizer into a number of well-defined regions in which supersaturation, the rate of energy dissipation, the solid concentration, and CSD are more or less uniformly distributed. With such an approach, the kinetic parameters for these models can be obtained from laboratory-scale experiments or can be estimated from simulation technology, whereas industrial-scale model simulations are able to predict the performance of a full-scale crystallization process. In order to simulate continuous crystallization in an imperfectly mixed suspension, Sha et al. [45] assumed different-sized particles to be distributed differently in the crystallizer. They proposed a population model that takes into account the relationship of particle size distributions, the product-removal location, and different points inside the crystallizer. The particle size distribution in the product can be obtained using kinetic models that can be obtained from experiments or using simulation technology. With this method, it is possible to study the effect of the geometry of the crystallizer, the mixing intensity, and the location of product removal on a continuous crystallization process.

Another task of applying CFD to crystallization is to study the geometry of the crystallizer. It can offer detailed information on the influence of the shape of

the crystallizer and impeller in producing the required mixing conditions of crystallization. Furthermore, the scale-up of crystallization will be a useful task for the study in the CFD environment. It is well known that the scale of the crystallizer affects the crystallization. This is mainly caused by the hydrodynamic differences for different scales of crystallizer. It is a very important subject with respect to the design of the crystallizer. Experimental studies on this subject require large financial resources. CFD simulation, on the other hand, supplies a possibility to investigate crystallization processes for different scales of crystallizer. Zauner and Jones [46] present a model in which the micromixing and mesomixing are combined with a population balance to study the effect of the mixing on the particle size distribution in the precipitation process. Satisfied results were obtained. Wei and Garside [42] studied the power-input effect on the precipitation process for different scales of precipitator with the moment transport population balance equation model. Because the basic model of simulation of crystallization with the tool of CFD is still under development, the scale-up subject cannot produce good results yet.

51.3 INDUSTRIAL CRYSTALLIZERS

Nowadays, a large number of industrial crystallizers are in use. In this section, we will discuss only a few of the basic and most presentative constructions in order to complement the principles explained earlier.

In cooling crystallization, supersaturation is obtained by undercooling the solution that is supersaturated by the evaporation of the solvent in evaporative crystallization. The relatively large average sizes of the end product can be obtained through both cooling and evaporative crystallization.

The criteria for the selection of an evaporative or cooling crystallizer depend on the slope of the solubility curve. If the slope of the solubility curve of a fairly soluble substance, i.e. ($\Delta c^*/\Delta t$), is relatively

low, evaporative crystallization is more preferable; but for a high-solubility slope, such as that of hydrated salts, cooling crystallization is a better alternative. In some compound systems, due to the high solubility of the solute, the final solute concentration of the residue remains too high, even at low temperatures. Evaporative crystallization should be employed in order to obtain as low a final solute concentration in the process solution as possible. This is the case when the main purpose of the crystallization process is to utilize solutes from the process water in order to obtain low-concentration waste solutions.

The operational costs of evaporative crystallization in countries where the cost of energy is low are lower than those of cooling crystallization.

Commercial crystallization processes are usually run in jacketed and baffled mixing tanks equipped with a draft tube, i.e., a draft-tube baffled (DTB) crystallizer. There are also forced-circulation (FC) processes where the solution is pumped through an external heat exchanger. In cooling crystallization, the pumped solution is cooled in the heat exchanger and heated in evaporative crystallization. An evaporative crystallizer is equipped with a vacuum system, which contains either a vacuum pump or an air ejector. Usually, the evaporative crystallization process consists of several stages. Table 51.1 shows the approximate operational conditions of FC, DTB, and FB-type crystallizers. A DTB, FC, and FB crystallizer, which is also an Oslo crystallizer, are schematically shown in Figure 51.7 through Figure 51.9. This equipment will be described in more detail in Section 51.3.2.

Commercial suspension crystallizers are manufactured, among others, by Swenson and GEA Messo GmbH. Wirges [47] cites the following companies as suppliers of solution crystallization processes: CT Umwelttechnik, HPD Evatherm AG, Chematur Eco-planning Oy, Ebner CO KG, Balcke-Dürr GmbH, and Gea Kestner SAS.

TABLE 51.1
The Approximate Operation Conditions of Crystallizers When $\rho^*/\rho_c > 0.01$.

Type	Suspension Density (kg/m ³)	Volume Ratio of Crystals to Suspension (m ³ /m ³)	Residence Time (h)	Specific Power Input (W/(kg Suspension))	$\Delta\rho/\rho_c$	L_{50} (mm)
Forced-circulation (FC)	200–300	0.1–0.15	1–2	0.2–0.5	10^{-4} – 10^{-2}	0.2–0.5
Draft-tube baffled (DTB)	200–400	0.1–0.2	3–4	0.1–0.5	10^{-4} – 10^{-2}	0.5–1.2
Fluidized-bed (FB)	400–600	0.2–0.3	2–4	0.01–0.5	10^{-4} – 10^{-2}	1–5(–10)

Note: ρ^* is the density of the saturated solution and ρ_c is the density of the crystal, i.e., solid density.

Source: Mersmann, A., Ed., *Crystallization Technology Handbook*, Marcel Dekker, New York, 1995, p. 219.

51.3.1 BATCH CRYSTALLIZATION

Usually, batch crystallization is used when a relatively low production capacity is required, e.g., below 50 t of product per day. When batch crystallization is equipped with the proper temperature control and seeding system, the crystallization conditions can be adjusted in such a way that the residence times of the crystals, of various sizes, can be kept about the same. Therefore, the CSD can be narrower in batch crystallization than in continuous crystallization, which is one of the significant differences between batch and continuous crystallization without fine removal or a classification method for the product. In practice, industrial continuous crystallization processes contain fines-removal or classification units, such as hydrocyclones, in order to produce crystals of a narrow CSD.

Jacketed tanks with propeller agitation are widely used for batch crystallization. The agitator is selected by the use of the same principles as in ordinary liquid blending when high pumping rate and low turbulence are required. This condition minimizes excessive secondary nucleation. For cooling crystallization, the programmed-cooling method, presented in Section 51.2.2, produces larger crystals than that by natural cooling. For natural cooling, the temperature first falls rapidly and then more slowly. In some cases, a batch process is equipped with a fines-destruction unit in order to increase the average crystal size or enable a more reliable downstream process, i.e., filtration, where operation conditions may be disturbed by small crystals. In fines destruction, a side flow

from the tank, after a crystal classifier, is directed to pass through a heat exchanger where the smallest crystals dissolve before the flow returns to the tank. Seeding, as was explained in Section 51.2.4, is commonly used.

Figure 51.6 shows a schematic figure of a jacketed, batchwise, multipurpose reactor with baffles, which is used in batch crystallization. For fine chemicals, the tank is usually glass-lined. Various kinds of impellers and propellers can be used as stirrers. From the perspective of the mixing basis and homogenous supersaturation conditions, it should be pointed out here that in this kind of mixing tank, equipped with a single impeller, good mixing takes place only close to the bottom, especially with large-sized crystals. Figure 51.6b shows the control system for programmed cooling, as discussed in Section 51.2.1.

51.3.2 CONTINUOUS CRYSTALLIZATION

Figure 51.7 through Figure 51.9 show some of the most common types of continuous crystallizers for large-scale production. They usually operate under vacuum conditions in order to reduce the evaporation temperature, and heat energy is provided by a heat exchanger. Of these, the FC crystallizer is shown in Figure 51.7. Although it is both reliable and easy to use, it does not have any devices that would allow the CSD to be influenced. It is basically an evaporator that is used as a crystallizer. Figure 51.8 shows a DTB crystallizer, which has become very popular over the years. Its useful features include the draft tube, which

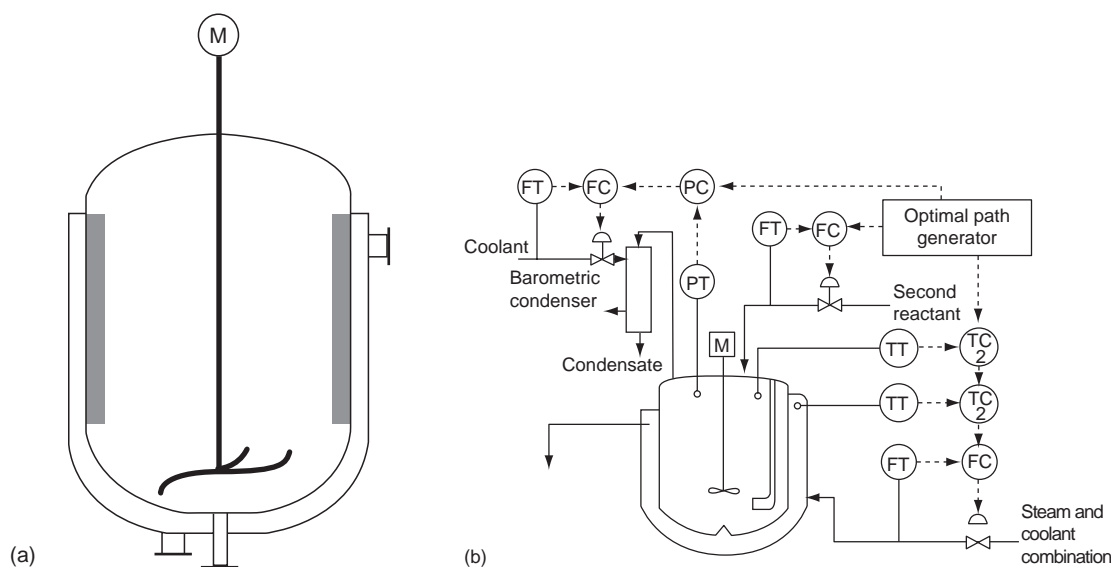


FIGURE 51.6 A schematic figure of a simple batch crystallizer (a) and one with a programmed-cooling system (b). (Reprinted from Mersmann, A., Ed., *Crystallization Technology Handbook*, Marcel Dekker, New York, 1995. With permission.)

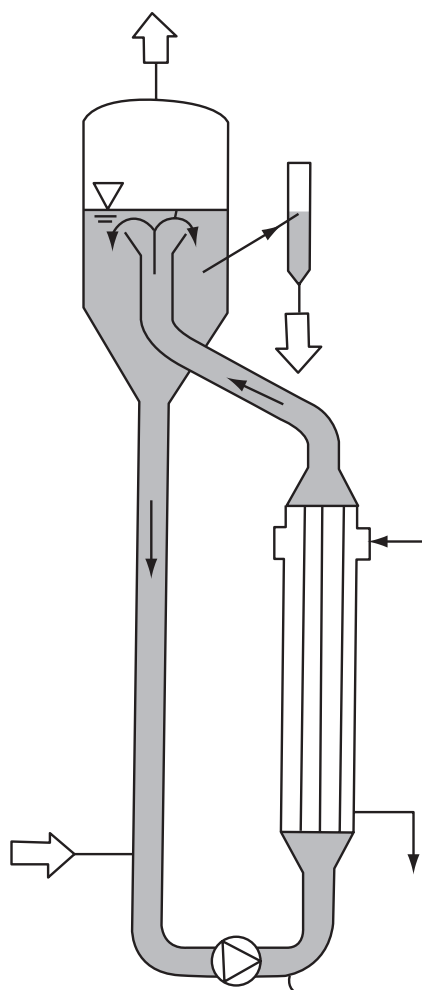


FIGURE 51.7 A forced-circulation (FC) crystallizer. (Reprinted from Mersmann, A., Ed., *Crystallization Technology Handbook*, Marcel Dekker, New York, 1995. With permission.)

improves circulation of the crystal magma, and a settling zone, which selects the small crystals to the circulation flow for partial destruction in the heat exchanger. In this way, the heat exchanger also functions as a fines dissolver, which helps to increase the mean crystal size in the product. The process of fines destruction was discussed earlier in [Section 51.2.5](#).

In the FB crystallizer ([Figure 51.9](#)), the heated mother liquor first goes to the evaporator section where supersaturation takes place. The supersaturated solution then directly enters into the fluid bed of product crystals where the desupersaturation occurs and the crystals grow. This equipment is known for its capacity to produce large crystals of a reasonably narrow CSD. Nevertheless, FB crystallizers are not as commonly used as would be expected on the basis of the advantages of the large product size. The disadvantage of FB crystallization is, for some

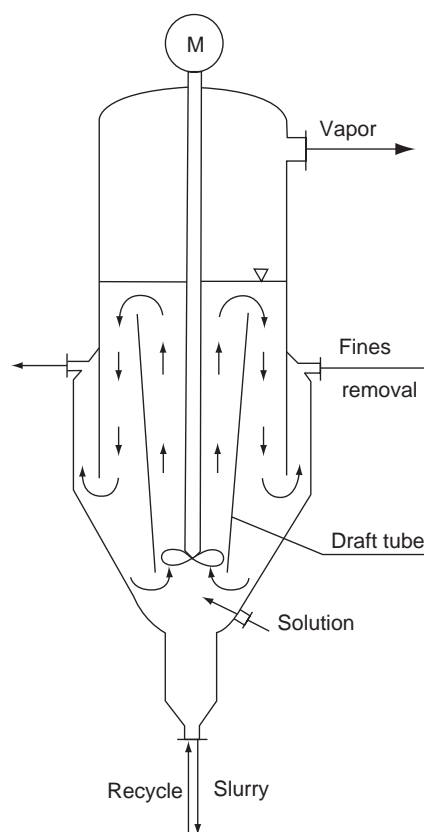


FIGURE 51.8 A draft-tube baffled (DTB) crystallizer. (Reprinted from Mersmann, A., Ed., *Crystallization Technology Handbook*, Marcel Dekker, New York, 1995. With permission.)

chemicals, the unreliability of its operation owing to the fact that unusually high crystal densities may develop in the fluid bed in the bottom end of the crystallizer tank because of which the slurry-withdrawal line tends to get blocked. Operational instability has been reported for some chemicals. When correctly operated with some chemicals, however, the FB crystallizer can produce large crystals of a narrow CSD.

[Table 51.2](#) shows some examples of operating data of large-scale crystallizers. Only some characteristic information of the more comprehensive data given by Mersmann [3] is presented here.

51.3.3 THE SELECTION OF A CRYSTALLIZER

To a certain extent, industrial crystallization is still more an art than science, which can be seen in the selection of a crystallizer. Often existing crystallization methods that have been previously proven successful for certain materials are used for these same materials in new plants, instead of new untried crystallizers. If exotic and unusual crystallization

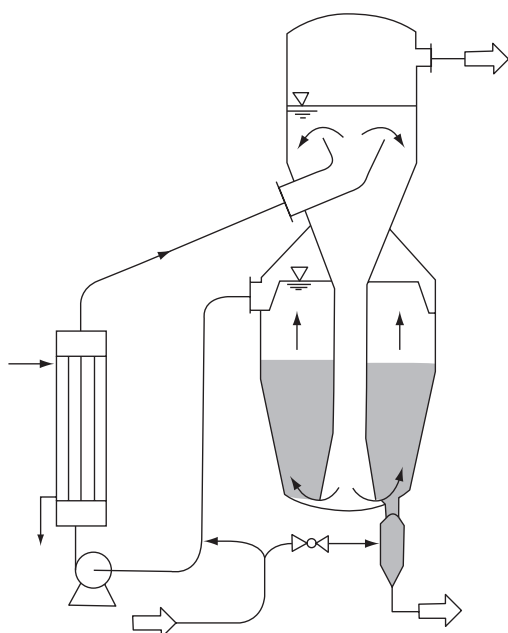


FIGURE 51.9 A fluidized-bed (FB) crystallizer (an “Oslo crystallizer”). (Reprinted from Mersmann, A., Ed., *Crystallization Technology Handbook*, Marcel Dekker, New York, 1995. With permission.)

processes are not taken into account, the thumb rule is that an agitated tank shown in Figure 51.6 is used in batch processes for the production of fine chemicals, whereas one of the types shown in Figure 51.7 through Figure 51.9 is normally selected for continuous processes.

51.4 FILTRATION OF CRYSTAL SUSPENSIONS

Suspension crystallization processes are the most common in industry. After crystallization, the crystalline product is usually separated from the solution

by filtration. The separation of crystals from the mother liquor has an important influence on the whole process. The reader is referred to some basic handbooks on filtration, such as *Solid-Liquid Separation* as edited by Svarovsky [48] and those by Wakeman and Tarleton [49], and Rushton et al. [50].

For easy separation of crystals from a solution, it is important that the crystals are sufficiently large. This can be shown by the Kozeny–Carman equation derived for laminar flow through an incompressible bed of particles, which describes the pressure drop in the bed. The pressure drop relates inversely to the square of the quadratic particle size and directly to the dynamic viscosity of the fluid. The influence of the particle size and solution viscosity on the pressure required in filtration can be estimated as follows. If the viscosity increases tenfold, the pressure drop also increases tenfold. If, on the other hand, the particle diameter drops to 10% of its original value, the pressure drop increases by 100-fold.

However, there are cases when the mother liquor has a relatively high viscosity due to high concentration of dissolved solute, the filtration of crystals from the mother liquor may require higher pressures as the driving force. This occurs quite often in the sugar industry where solutions have a high dynamic viscosity up to 100 MPa (cP). In such cases, the pressure drop in the bed, as compared with water, is 100-fold.

For the above reasons, the aim of suspension crystallization is usually to produce sufficiently large crystals. However, for example, in some applications, such as in the pharmaceutical industry and the production of pigments, paper fillers, and paper-coating substances, crystals of rather small sizes are required. For this reason, the selection of the filter medium plays a key role in order to obtain a sufficient filtration capacity and as low a mother-liquor volume in the crystal cake as possible. Besides cake washing, the fouling of the filter medium is a factor that affects the filtration of small particles. The most commonly used filter media are made of synthetic fibres, such as nylon, polyester, polypropylene, polyethylene, or fluorocarbon [50]. In some fine particle applications, it has been observed that, by adjusting the pH level, it is possible to control the formation of agglomerates or aggregates on the basis of changes in the surface charge at the particle surface surrounded by a double layer. Prior to filtration, the pH level is adjusted in such a way that the solution attains its isoelectric point and the repulsive forces are at their minimum. After filtration, the pH level is adjusted in order for secondary particles to decompose. Flocculation agents and filter aids, such as diatomite and perlite, are commonly used to facilitate the filtration of

TABLE 51.2
Data of Continuous Industrial Suspension Crystallizers

	Citric Acid	KCl	Na ₂ SO ₄
Production rate (kg/h)	560	7200	10100
Residence time (s)	8380	8100	4040
Volume (m ³)	12	1050	2800
Medium size (mm)	0.41	1.1	0.30
Suspension density (kg/m ³)	132	15	151
Crystallizer type	FC	DTB	FC

Source: Mersmann, A., Ed., *Crystallization Technology Handbook*, Marcel Dekker, New York, 1995, p. 614.

fine particles. This subject is discussed in detail in *Solid-Liquid Separation* [48].

The obtained CSD affects the required filtration time. Furthermore, it influences on the final moisture content of the filtered cake, which means with higher moisture content higher energy in thermal drying is required. The width of CSD and the crystal shape are important for the filterability of the crystalline product. If the product contains a lot of tiny crystals, filtration becomes more difficult due to increased specific surface of the crystal cake. Disk-shaped crystals decrease the permeability of the cake due to their high tortuosity and low void ratio. Furthermore, the elasticity of crystals, which is typical for organic compounds, may cause an increase in the compressibility of the crystal cake, thus decreasing the permeability of solid cake at higher filtration pressures.

Häkkinen et al. [51] investigated the filterability of sulfathiazole crystals obtained by crystallization from one-solvent and two-solvent solutions. Their results showed that the choice of the mother-liquor composition affected the obtained resistance of crystal cake. Lowering the cake resistance in a system usually decreases the moisture content of the cake that in turn facilitates the drying process. Organic compounds may also form amorphous solid particles that usually lead to low permeability of the solid bed due to the compression of the filter cake.

Crystal purity depends mainly on filtration efficiency and crystal washing. For instance, when the mother liquor contains high levels of impurities, the level of purity after filtration is affected by the moisture content of the crystal cake, i.e., the amount of mother liquor left in the crystal cake. The washing of the crystals may require high volumes of washing solution. The use of large volumes of washing solution does not dissolve sparingly soluble crystals, i.e., washing does not decrease the yield. Contrary to precipitates, when compounds that have a high level of solubility are used, washing may dissolve the crystals. If the impurity has a relatively low melting point, it can be partially melted by heating the crystal cake and then separating the impurity from the crystal cake by the force of gravity, an airflow, or with the use of a pressure chamber. This technique is called sweating.

One of the most commonly used solid-liquid separation methods in crystallization processes is centrifugal filtration, such as continuous pusher and batchwise peeler centrifuges shown schematically in Figure 51.10. A manufacturer of centrifuges used in crystallization processes is KMPT AG [52]. In addition, Nutsche filters, frame pressure filters, and belt filters have also been used. Most of these filters have a possibility of cake washing which is important for the final purity as discussed earlier.

It may be difficult to choose the right filtration equipment for a crystallization process. The selection of filtration equipment involves balancing the process specifications and objectives with the capabilities and characteristics of the various equipment alternatives. Wakeman and Tarleton [49] provide guidelines for the selection of filtration equipment. Important process-related factors are the slurry character, production throughput, process conditions, performance requirements, and permitted construction materials. The important equipment-related factors are the type of cycle (batch or continuous), driving force, production rates of the largest and smallest units, separation sharpness, washing capability, dependability, feasible construction materials, and cost. No absolute selection techniques are available for obtaining the best choice, as so many factors are involved. Nevertheless, there are some general guidelines for the engineer who faces the selection of filtration equipment. Furthermore, a commercial software package, p^C -SELECT, based on expert systems is available for filter selection, developed by Wakeman and Tarleton [49]. The relative performance characteristics of different filtration equipment under vacuum and pressure are shown in Table 51.3 with the solid concentrations of feed higher than 10 vol%. The crystal volume related to the total suspension volume in industrial crystallizers is frequently 20 vol%. In vacuum filtration, the maximum pressure difference over the cake is below 1 bar, whereas in pressure filtration the maximum driving force of conventional filters can be up to tens of bars. The indices are given for the dryness of the solid product, the effectiveness of solid washing, the quality of the liquid product, and the tendency of the equipment to cause crystal breakage. In addition, Table 51.3 shows the basic feed properties that the equipment can generally handle. Once an initial selection of equipment has been made, the list of the equipment can be sensibly ranked using Table 51.3. Table 51.3 shows that centrifugal filtration results in the highest product dryness. On the other hand, average particle sizes are relatively large and the centrifuging may increase crystal attrition. Vacuum and pressure leaf filters give less-dry cakes than centrifuges, but leaf filters are gentler on the crystalline product and cake washing is more efficient.

51.5 CRYSTAL DRYING

Under otherwise similar drying conditions, the drying rate depends on the surface area of the wet crystals and the air used in drying. This means that small crystals dry quickly in airflow, such as that produced by a pneumatic dryer. Instead, the drying of a solid bed of

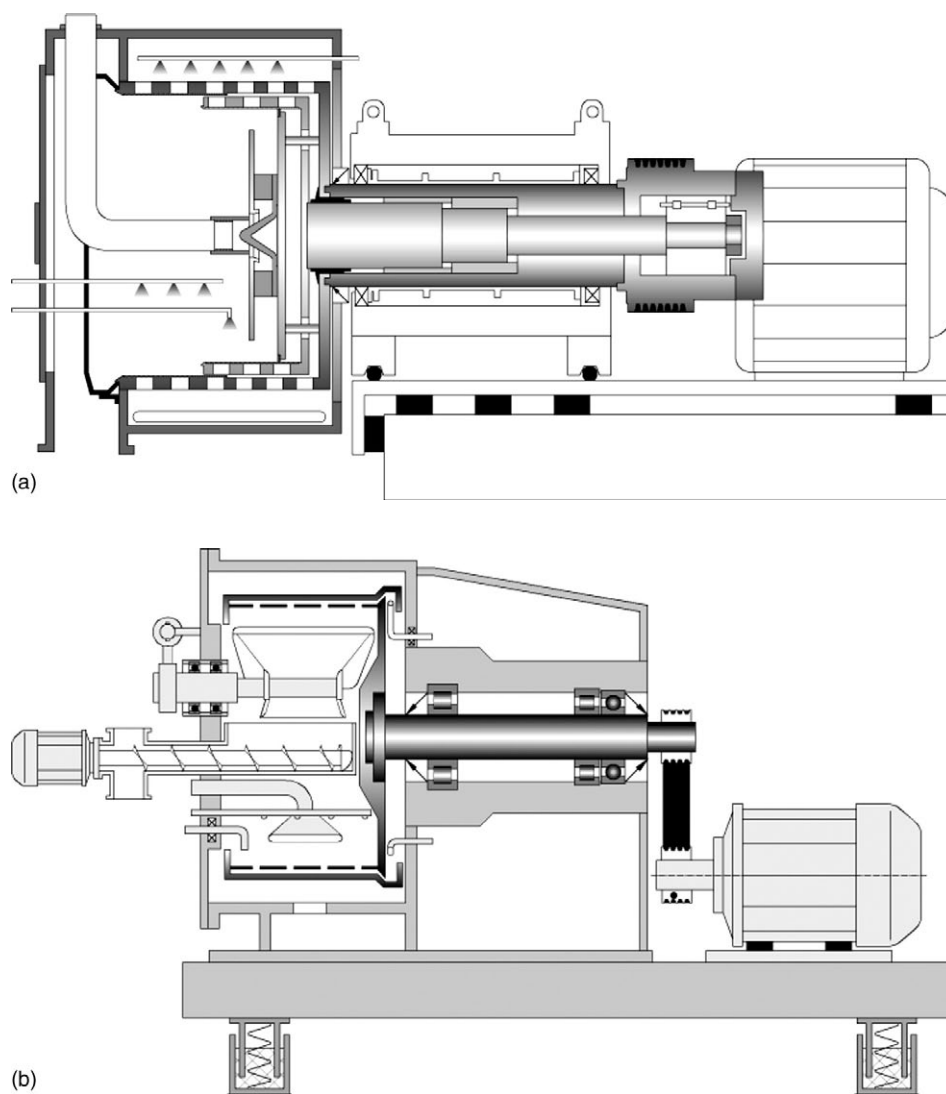


FIGURE 51.10 A pusher centrifuge (a) and a peeler centrifuge (b). (Reprinted from KMPT AG, Internet <http://www.kmpt.com>. With permission.)

crystals tends to be less satisfactory because of the long drying time required and due to agglomeration caused by the contact of the wet crystals. On the other hand, a bed of small crystals cannot be filtered easily but, instead, tends to block the filter. From the viewpoint of drying, it would be best that the crystals be large, within the range of about 1 mm or above. After the filtration stage, the amount of mother liquor in the crystals is low. The large crystal size also improves purity because the same thickness of the attached mother liquor on the surface, which contains impurities, results in a lower level of impurities in large crystals. If the mother liquor remains on the surface of the crystals, it solidifies, with the impurities that it contains, on the surface of the crystal. It should

be mentioned here that crystal sizes above approximately 1 mm tend to be harmful. For crystals larger than 1 mm, it may be difficult to maintain the steady state in a continuous process, due to the decreased overall crystal surface required for releasing supersaturation. Furthermore, large crystals may break in the centrifuge.

The aim of the earlier discussion was to explain how crystals of a desired size could be produced. Furthermore, the CSD should be as narrow as possible for easy drying. In principle, the drying of crystals can be carried out in the same way as that of any particulate material. However, there are some cases when the crystalline structure itself poses problems in drying. We will briefly discuss these cases.

TABLE 51.3
Relative Performance Characteristics of Different Filtration Equipments

Types of Equipment	Performance Indices				Feed Solid Properties	
	Solid Product Dryness	Washing	Liquid Product Quality	Crystal Breakage	Particle Size (μm)	Solids in Feed (vol%)
<i>Filtering centrifuges</i>						
Basket (pendulum)	9	6	5	6	10–1,000	2–10
Basket (peeler)	9	6	5	5	2–1,000	2–30
Cone screen (slip discharge)	7	5	4	4	80–10,000	10–40
Cone screen (vibrating, oscillating, or tumbling)	8	5	4	3	100–10,000	10–40
Cone screen (worm screen)	9	5	4	4	60–5,000	10–40
Pusher (single-stage)	9	7	4	4	40–7,000	4–40
Pusher (multistage)	9	8	4	4	40–7,000	4–40
Baffle	9	5	5	4	100–7,000	4–40
<i>Vacuum filters</i>						
Single leaf (tilting pan)	7	9	7	8	20–80,000	3–40
Multielement leaf	5	5	7	8	1–100	3–30
Horizontal belt or rotary tilting pan	7	9	7	8	20–80,000	3–40
Rotary table	7	8	7	8	20–80,000	3–40
Rotary drum (bottom fed)						
-Knife discharge	6	7	7	8	1–200	3–30
-Roller discharge	6	7	7	8	1–50	3–30
-String discharge	6	7	7	8	1–70	3–30
Rotary drum (top fed)	5	2	7	8	1–600	3–30
Rotary drum (internal fed)	5	—	7	8	10–600	3–30
Rotary disk (cloth covered)	4	—	6	8	1–700	3–30
Rotary disk (ceramic)	4	—	9	8	1–700	3–30
<i>Pressure filters and Presses</i>						
Single leaf (pressure Nutsche)	6	8	8	8	1–200	0.005–30
Filter press	6	8	8	8	1–100	0.005–30
<i>Variable volume pressure filters and presses</i>						
Diaphragm filter press	8	8	8	7	1–200	0.1–25
Tube press	8	4	7	7	1–200	0.1–25
Expression press	6	—	6	5	1–200	5–50
<i>Continuous pressure filters</i>						
Belt press	8	7	7	7	1–200	0.1–25
Vertical filter press	6	6	7	7	1–100	0.01–30
Vertical diaphragm filter press	8	8	8	7	1–200	0.1–25
Tower press	6	—	7	7	1–300	0.1–25
Rotary pressure drum	6	6	7	7	1–100	3–30
Rotary pressure disk (cloth covered)	5	—	6	8	1–100	3–30
Rotary pressure disk (ceramic)	7	—	9	8	1–100	3–30

Allocation 0 to 9 with larger numbers indicating better performance, A “—” performance indicates either equipment is not effective or not suitable for a particular duty.

Source: Wakeman, R.J. and Tarleton, E.S., *Filtration, Equipment Selection, Modelling and Process Simulation*, Elsevier Science, Kidlington, U.K., 1999.

Most crystals are so soft that the corners of the crystalline particles tend to get rounded if collisions occur between the crystals during drying and, as a result, the quality of the product suffers. In addition, dust may be a problem. For example, a traditional rotary

dryer is not suitable for most crystals. Surprisingly, both fluid-bed dryers and pneumatic dryers are relatively gentle, perhaps, because of the shorter residence time.

Then, there is the problem of crystal water. These are often salt hydrates, i.e., inorganic crystals with

different numbers of water molecules attached to each molecule of the basic molecule. Drying may remove crystal water, which leads to quality problems in the product. Furthermore, crystallization at high temperatures may cause the agglomeration and solidification of the product during storage.

51.6 NEW ADVANCES

To conclude the present review, we will mention some trends in the research and development into the technology that may be of importance in the present context in the future. Due to constraints of space, we have to omit most of the important works presently being carried out on various aspects of crystallization, and will only mention those works that may lead to a particle size distribution and particle shapes which facilitate the filtration and drying stages in the process.

The traditional study of suspension crystallization has been carried out using the MSMR crystallization model. It has been found that uniform mixing in a commercial-size crystallizer, as required by the MSMR model, is impossible to achieve. Therefore, the understanding of industrial crystallization is hampered by the use of the MSMR model. Also, it is difficult to experimentally study the effects of mixing on crystallization, as described earlier in [Section 51.2.5](#). Therefore, the CFD presents the means for local simulation in the tank. Furthermore, CFD simulation enables the tank to be designed so that the shape and the positioning of the impellers and the liquid velocity create the optimal level of supersaturation and mass transfer rate in all locations. This is likely to result in a narrowing of the particle size distribution.

Then, there is work with the objective of altering the particle shape, usually from a plate- or needlelike shape into a more rounded shape such as that of a cube or octahedron. The starting point here is the concept of polymorphism discussed earlier in [Section 51.1.5](#). A material may have different crystal structures, and it is useful to find ways of generating the ideal structure. One way to do this is to use different solvents, as mentioned by Davey and Garside [6]. They also discuss the new and promising field of tailor-made additives, which alter the crystal shape during the growth process. In addition, Finnie et al. [53] reported that the level of supersaturation selected different polymorphs for growth, i.e., the level of supersaturation can be used to control the crystal shape. This idea was taken further by Ma and Braatz [54] who showed

that different cooling profiles in batch crystallization, discussed earlier in [Section 51.2.1](#), produced different crystal shapes. Although the above are early results, they, however, show that there may be ways to control the crystallization processes in order to achieve the most desirable product shapes.

ACKNOWLEDGMENT

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NOMENCLATURE

B	nucleation rate per unit volume, $1/(s\ m^3)$
c	concentration, mol/m^3
G	crystal growth rate, m/s
K	parameter, $K = (1 + 4N_s/B\tau)^{-1}$, ranging from 0 (no nucleation) to 1 (unseeded batch), dimensionless
k_n	kinetic empirical constant, different units interchangeably
L	crystal size, m
M_T	suspension density, kg/m^3
n	population density, number of crystals in a unit volume at each size class, $1/m^4$
N_s	number of seed crystals per unit volume of solvent, $1/m^3$
S	supersaturation ratio, dimensionless
t	time, s
\dot{V}	flow rate, m^3/s

Greek Symbols

β	fraction of nuclei that survive in the fines-destruction system, dimensionless
σ	relative supersaturation, dimensionless
τ	overall batch time, residence time, s

Subscripts

*	equilibrium, saturated
0	beginning, feed
1	without fines destruction
2	with fines destruction
50	cumulative mass undersize of 0.5
c	clear-liquor advance
d	dominant
f	final, fines
p	product

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52 Frying of Foods

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52.1 INTRODUCTION

Frying is defined as a process of cooking and drying through contact with hot oil. It is intended to make food more palatable and tasteful, but at the same time makes food safer and provides a preservative effect that results from thermal destruction of microorganisms and enzymes, and a reduction in water activity at the surface or throughout the food. The shelf life of fried products is mostly determined by the moisture content after frying. Products that retain a moist

interior should be consumed shortly after preparation, or can be stored for a relatively short time under chilling conditions or for a longer time under freezing conditions. Most of these foods—with the exception of par-fried goods—are not produced on a commercial scale for distribution to retail stores, but are important in catering applications. Foods that are dried throughout during frying have a shelf life up to several months, which is mostly limited by quality deterioration of the absorbed oil and development of a rancid odor and flavor. Storage stability of these

products may be increased by using packaging materials with adequate barrier properties.

The term “industrial frying” is used to describe high-volume deep-fat frying conducted usually as a primary business. Most industrial fryers are continuous and highly automated with capacities up to some tons per hour. Large batch frying systems are also included in the industrial frying classification. Fried foods have a unique flavor—texture combination that makes them highly acceptable by the consumers. Therefore, a wide range of products has been developed and new products are continuously introduced in the market. The main categories of fried products are presented in Table 52.1.

Potato chips, extruded snacks, nuts, and nut mixtures are dried throughout during frying, therefore, they have a low moisture and can be stored for a long time at ambient temperature. Potato chips (crisps) are the original savory snack food and the leading one in the United States and Canada [48,52]. In addition to the traditional thin, salted slices of potatoes, chips of different sizes, shapes, and flavors are currently produced in several countries [58,69]. Also, fabricated chips are prepared from dough containing mainly dried potatoes, gelatinized starch, and gluten [33]. Moreover, corn and tortilla chips are becoming popular in many parts of the world [48,69].

Extruded fried snack foods consist another major produce of the food industry [69]. As extruder technology has developed, the sizes and shapes of extruded products have become almost limitless. Having a milled cereal, e.g., corn, wheat, oats and rice, potatoes, or legumes as starting material, the extruded snacks are flavored with various ingredients to provide many common flavors, e.g., pork, cheese, and pizza, and new ones are also continuously introduced in the market [1,69]. Nuts and nut-based snacks are very popular all over the world. Nut mixtures containing rice, legumes, vegetables, and many

other ingredients, as well as various flavors are used in the fried products.

Doughnuts and other fried bakery goods like pies probably exceed in tonnage, in the United States, any other type of sweet dough product [41]. They have intermediate or high moisture content and can be preserved for a short period before consumption. Doughnuts are usually prepared from yeast-leavened dough, where honey or flavor ingredients may be added. The familiar doughnut shape predominates, but other shapes, e.g., twists, sticks, honey buns, and rolls are also produced through sheeting and cutting or extrusion processes. After frying they may be coated with powdered sugar, glazes, icings, and toppings or filled with jellies, fruits, and jam. Fried pies present special problems during frying due to the filling they contain. The crust must be thoroughly cooked to a medium brown color and the filling must be cooked to an optimal point before significant steam evolution occurs in the interior of the pie that may lead to bursting.

Par-fried (or prefried) french fries is one of the most important products of potato processing industries [33]. They have high moisture content and are stored under freezing until the finish frying. The industrial process includes a water-blanching step, prior to frying, in order to obtain a more uniform color and improved texture of the fried product, to reduce fat absorption by gelatinizing the surface layer of starch and to reduce frying time. The use of additives during blanching can further improve color and texture [33]. Coating with hydrocolloids may also be used to reduce oil absorption and to improve texture [34]. Other par-fried foods include, e.g., fish sticks and pieces of chicken. Most of them are also stored under freezing until the finish frying, but storage under chilling and consumption in a few days may be also practiced. Some of these products are covered with some batter before prefrying. This coating results in the formation

TABLE 52.1
The Main Classes of Fried Products

Class	Products	Fat Content (%)
(Potato) Chips	Potato chips of various shapes and flavors, tortilla chips, and corn chips	25–40
Extruded snacks	Snacks of various shapes, flavors and texture, prepared with cereal, potato, or legume flour	20–40
Nuts	Nuts or nut mixtures with legumes and flavorings	5–6 ^a
Bakery products	Doughnuts and fried pies	10–25
Par-fried frozen food	French fries, fish, chicken, and pancakes	5–7
Food-service products	Chicken, fish, meat, and potatoes	Broad range

^aExchanged with natural oil.

of a crust with the desirable color and texture [2,56]. Par-frying aims to decrease the necessary frying time at home or restaurant, while at the same time reduces the load of microorganisms in food, offering a preservative effect.

The foods mentioned so far are the main products of industrial frying operations. Several other foods like chicken, fish, meat, and potatoes are fried in large quantities in fast food, catering enterprises, and restaurants. The fried-chicken-on-the-bone is a typical example of a product with annual sales of few billion dollars [60].

52.2 THE FRYING PROCESS

When food is placed in hot oil, a complex heat-and-mass-transfer phenomenon takes place. Heat is transferred by convection from the surrounding oil to the surface of the food and by conduction within the solid food. Water is vaporized and escapes from the surface of the food to the oil, while oil penetrates into the food. Some food components may also diffuse and dissolve into the oil. As water vapor moves away, the surface begins to dry and crust is formed. Additionally several physicochemical changes occur in food, e.g., starch gelatinization and protein denaturation. Also physicochemical changes take place in oil, e.g., oxidation reactions, hydrolytic cleavage, change of color, and viscosity. Dissolvment of food components into the oil may affect the rate of these changes.

Convective heat transfer at the surface of the food is complicated due to vigorous movement of vapor bubbles escaping from the food that cause a considerable turbulence in the oil. Additionally vapor bubbles are entrapped under the bottom surface of the food being fried and prevent efficient heat transfer to that side of the food. Another problem is the accumulation of some food items at the top of the oil bath where oil foam is formed that presents a significantly different heat transfer coefficient.

As frying proceeds the amount of water vapor bubbles escaping from food decreases as a result of decrease in the remaining moisture in the material. The crust that is initially formed at the surface of the material moves towards inside and the temperature of the dried regions rises above the boiling point of the water. The moist core of the material cannot exceed the boiling point of the liquid phase. Since the liquid phase is water with some solutes, its boiling point is slightly elevated above the boiling point of water. Based on experimental observations Farkas [66] suggested that the frying process is composed of four distinct stages:

1. *Initial heating*—It begins with immersion of the food in the frying oil and lasts until the temperature at the surface of the food becomes equivalent to the elevated boiling point of the liquid. The duration of this stage does not exceed a few seconds. During this short period the mode of heat transfer between the oil and the food is natural convection and no vaporization of water occurs from the surface of the food.
2. *Surface boiling*—Water vaporization signals the beginning of this stage. The bubbles formed cause turbulence in the oil surrounding the food and, therefore, the mode of heat transfer to the food surface changes from natural to forced convection. Crust starts to form at the surface of the food.
3. *Falling rate*—During this stage, that resembles the falling-rate period observed in food dehydration processes, more internal moisture leaves the food, the internal core temperature rises to the boiling point of the liquid, and the crust layer increases in thickness. After a sufficient time and more removal of moisture, the vapor transfer at the surface decreases.
4. *Bubble end point*—If the frying process is extended to a considerably long time, the rate of moisture removal diminishes due to effective decrease of moisture content, and no more bubbles escape from the surface of the product. This final stage is referred to as the bubble end point.

Oil penetrates into the food at the dried region that has a porous structure, consisting of different-sized capillaries. As more water leaves the food and the crust thickness increases, more oil is absorbed in the food. At the end of frying, when the food material is removed from the oil bath, an oil layer remains at the surface of the food and oil from this layer is further absorbed into the pores of the material due to capillary forces.

Foods that retain a moist interior are not fried to the bubble endpoint stage, but it is essential for food safety that the thermal center has received adequate heat to destroy microorganisms. This is particularly important for foods that are able to support pathogenic bacteria, like meat and fish products.

There are two main methods of commercial frying: (1) the deep-fat frying and (2) the swallow frying. In *deep-fat frying* or *immersion frying*, the food is covered by the oil and all surfaces receive a similar heat treatment to produce a uniform color and appearance (Figure 52.1). Heat is transferred by convection to the surface of the food material and by conduction to the interior of the food. Frying follows

the stages mentioned above and convective heat transfer coefficient changes with bubble formation because of turbulence. Deep-fat frying is suitable for foods of all shapes, but irregularly shaped food tends to entrain a greater volume of oil when it is removed from the fryer. The deep-fat frying is the method used almost exclusively by the food industry, to produce snacks, potato chips, and several fried goods.

Shallow frying or contact frying is most suited to foods that have a large surface/volume ratio. The food is placed in the frying pan that contains a thin layer of oil and heat is transferred to the bottom surface of the food material mostly by conduction from the hot surface of the pan through the thin layer of oil (Figure 52.1). The thickness of the oil layer varies owing to irregularities in the surface of the food. This, together with the escape of water vapor bubbles from the hot surface of the food, causes temperature variations and produces the characteristic irregular browning of shallow-fried foods. Shallow frying has high surface heat transfer coefficients, but not stable across the entire surface of the food.

52.2.1 HEAT TRANSFER

Heat transfer has been studied in deep-fat frying of various foodstuffs. All the developed models consider that only conductive heat transfer occurs inside the product. The liquid water is considered to move from the inside of the product being fried to the evaporation zone, leaving the product surface as a vapor. Therefore, the rate of water loss is accounted for, in heat transfer models. The oil penetrates into the food mainly during the cooling period after frying, as proved by many investigators, therefore, heat transfer due to oil migration in the product is considered negligible.

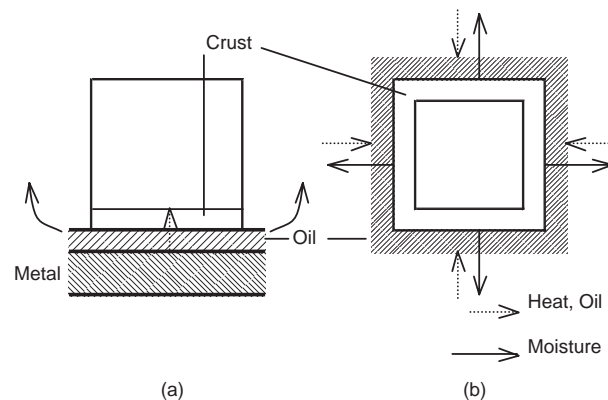


FIGURE 52.1 Heat and mass transfer in: (a) shallow frying and (b) deep-fat frying.

A model for heat transfer in chips, which are dried throughout frying, was developed by Moreira et al. [48]. Since thickness of the chips was one order of magnitude smaller than the other dimensions, infinite slab geometry was considered, with negligible shrinkage. The food was considered homogeneous during the frying process and a unique differential equation was used to describe heat transfer through the chip with a constant thermal diffusivity

$$\frac{\partial \Theta}{\partial t} = \alpha \left(\frac{\partial^2 \Theta}{\partial x^2} \right) \quad (52.1)$$

and with the boundary condition

$$-k \frac{\partial \Theta}{\partial x} \Big|_{x=x_0} = h(\Theta_{x=x_0} - T) - h_{fg} D \rho \frac{\partial M_w}{\partial x} \Big|_{x=x_0} \quad (52.2)$$

where Θ is the tortilla chip temperature ($^{\circ}\text{C}$), t is the time (s), α is the thermal diffusivity ($\text{m}^2 \text{s}^{-1}$), x is the variable distance across the thickness of the chip (m), k is the thermal conductivity of the chip ($\text{W m}^{-1} ^{\circ}\text{C}^{-1}$), h is the convective heat transfer coefficient ($\text{W m}^{-2} ^{\circ}\text{C}^{-1}$), h_{fg} is the latent heat of vaporization, (kJ kg^{-1}), D is the mass diffusivity ($\text{m}^2 \text{s}^{-1}$), ρ is the density (kg m^{-3}), and M_w is the moisture content (% dry basis).

Farkas et al. [19,20] studied heat transfer in larger items of food, which retain a moist core after frying, e.g., french fries. They considered the food as a semi-infinite slab and the movement of the crust/core interface as a moving boundary problem. The properties of the crust were considered to be uniquely different from those of the core and separate mathematical expressions were developed to describe heat transfer in each region. The crust region was assumed to contain a negligible amount of water and the conductive heat transfer coefficients of each region were considered constant.

In core region

$$k^{\text{II}} \frac{\partial^2 \Theta}{\partial x^2} + N_{\beta x} c_{p\beta} \frac{\partial \Theta}{\partial x} = (\epsilon_{\beta} \rho_{\beta} c_{p\beta} + \epsilon_{\sigma} \rho_{\sigma} c_{p\sigma}) \frac{\partial \Theta}{\partial t} \quad (52.3)$$

In crust region

$$(\epsilon_{\gamma} \rho_{\gamma} c_{p\gamma} + \epsilon_{\sigma} \rho_{\sigma} c_{p\sigma}) \frac{\partial \Theta}{\partial t} = k^{\text{I}} \frac{\partial^2 \Theta}{\partial x^2} + N_{\gamma x} c_{p\gamma} \frac{\partial \Theta}{\partial x} \quad (52.4)$$

where k^{I} , k^{II} are the thermal conductivities in crust and core regions, respectively ($\text{W m}^{-1} ^{\circ}\text{C}^{-1}$), Θ is the temperature ($^{\circ}\text{C}$), N_{ix} is the flux of species i in x -direction ($\text{kg m}^{-2} \text{s}^{-1}$), c_{pi} is the specific heat of species i ($\text{J kg}^{-1} ^{\circ}\text{C}^{-1}$), ϵ_i is the volume fraction of species i ($\text{m}^3 \text{m}^{-3}$), ρ_i is the density of species i ($\text{kg of } i \text{ m}^{-3}$), and subscripts β , γ , and σ refer to liquid water, water vapor, and solid, respectively.

Several initial and boundary conditions were considered, as well as, different values for the convective heat transfer coefficient for free convection ($250 \text{ W m}^{-2} \text{ K}^{-1}$) and convection when boiling exists ($500 \text{ W m}^{-2} \text{ K}^{-1}$). Experimental results were in good agreement with values of temperature predicted by the model. The crust temperature raised quickly at the boiling point of water, remained for a short period until the water present vaporized and raised subsequently much higher than the boiling point of water, depending on the oil temperature. The crust/core interface remained at a constant temperature of 102°C after the crust was formed. The temperature profile in the core region remained unaffected by the oil temperature because of the presence of the boundary at the crust/core interface. The same results in potato strip frying were observed by Pravisani and Calvelo [57], who found the boundary temperature equal to 103°C .

Vijayan and Singh [70] to predict heat transfer during frying of frozen foods developed a similar model. The model involves two moving boundaries, one between frozen and unfrozen regions and the other between crust and core.

For products that do not form a crust during frying, e.g., sausages, a model considering the food material as homogeneous and isothermal throughout frying was proposed by Dincer [15], and a new parameter defined as frying coefficient was inserted to facilitate and accurately determine thermal diffusivity of the product, which was considered constant.

$$\frac{\partial^2(\Theta - T)}{\partial r^2} + \frac{1}{r} \frac{\partial(\Theta - T)}{\partial r} = \frac{1}{\alpha} \frac{\partial(\Theta - T)}{\partial t} \quad (52.5)$$

where Θ is the food temperature ($^\circ\text{C}$), T , is the oil temperature ($^\circ\text{C}$), $\alpha = \frac{FR^2}{\mu_1^2}$ is the thermal diffusivity ($\text{m}^2 \text{ s}^{-1}$), F is the frying coefficient (s^{-1}), R is the radius (m), and μ is the root of the transcendental characteristic equation.

The sausages were considered as infinite cylinders and thermal diffusivity was calculated equal to $3.846 \times 10^{-7} \text{ m}^2 \text{ s}^{-1}$. Constant thermal diffusivity and homogeneity of the food were also the simplifying assumptions necessary to develop a model to predict internal temperature profile in chicken pieces fried under pressure [60]. For the chicken pieces a three-dimensional heat transfer was considered and the predicted results were closer to experimental ones obtained at the center of the pieces, than those obtained near the boundaries. Better simulation was also obtained for more regularly shaped pieces.

In the models described above, a constant convective heat transfer coefficient was assumed during the hole frying process, or two independent values were considered: (1) one for the initial frying stage

and (2) another for the following stages where bubbling occurs. It is generally accepted that the convective heat transfer coefficient depends on the type of shortening used for frying, its viscosity, and the turbulence caused by water vapor bubbles released during the process. In the stage of initial heating, when no vaporization of water occurs surface heat transfer is accomplished by natural convection. Using a metal transducer, simulating the food being deep fried Califano and Calvelo [9] found that the convective heat transfer coefficient ranged between 150 and $165 \text{ W m}^{-2} \text{ K}^{-1}$ for temperatures of 50 to 100°C . Following the same method Miller et al. [46] determined the convective heat transfer coefficients for canola oil, palm oil, corn oil, and soybean oil at 170 , 180 , and 190°C and found that they ranged between 250 and $280 \text{ W m}^{-2} \text{ K}^{-1}$. Determination of the convective heat transfer coefficient during frying was attempted by Sahin et al. [64], who assumed a constant value during the whole process. They used sunflower oil at different frying temperatures and found that the convective heat transfer coefficient increased from $90 \text{ W m}^{-2} \text{ K}^{-1}$ at 150°C to $200 \text{ W m}^{-2} \text{ K}^{-1}$ at 190°C . The dependence of heat transfer coefficient on the water loss rate of potato during frying was studied by Costa et al. [13]. They found that the convective heat transfer coefficient increased, following the increase of water loss, until complete drying of the potato surface and then decreased until the end of frying. The maximum values were up to two times greater than the values obtained in the absence of frying, and ranged (for sunflower oil) between 443 and $750 \text{ W m}^{-2} \text{ K}^{-1}$, depending on the oil temperature and product geometry. Higher values (30% on an average) were observed for potato chips, compared with french fries. They also commented that when the water loss rate is very high, the bubbles near the product surface may hinder the heat transfer. Hubart and Farkas [31] presented values starting from $300 \text{ W m}^{-2} \text{ K}^{-1}$ before the beginning of bubbling, increasing to $1100 \text{ W m}^{-2} \text{ K}^{-1}$ and then decreasing below $200 \text{ W m}^{-2} \text{ K}^{-1}$ after 900 s . Fellows [21] reported values of 250 to $300 \text{ W m}^{-2} \text{ K}^{-1}$ before bubbling and 800 to $1000 \text{ W m}^{-2} \text{ K}^{-1}$ during vigorous boiling. He also reported surface heat transfer coefficients of 200 to $450 \text{ W m}^{-2} \text{ K}^{-1}$ during shallow frying.

Viscosity is another factor that affects the heat transfer coefficient between the oil and the food [46]. Viscosity increases with frying time at a very slow rate, therefore, the oil has to be used for several hours before any major difference is observed. The fresh oil has a higher convective heat transfer coefficient than the used oil. Thus, the value for fresh soybean oil at 188°C was found equal to $282 \text{ W m}^{-2} \text{ K}^{-1}$ compared with $261 \text{ W m}^{-2} \text{ K}^{-1}$ for the used oil [45], while similar

values, i.e., $285 \text{ W m}^{-2} \text{ K}^{-1}$ for fresh oil compared with $273 \text{ W m}^{-2} \text{ K}^{-1}$ for used one were obtained for soybean oil at 190°C [48]. It should be noted, however that the increase of viscosity is accompanied by decrease in surface tension, which facilitates the contact between the oil and the food and increases the rate of heat transfer, according to Blumenthal [5].

Another parameter considered as constant in most models is thermal conductivity. In fact thermal conductivity decreases with frying time, as water is removed from the product and oil, which has a lower thermal conductivity than water, is absorbed. Thermal conductivity is also affected by other changes that may occur in the food, e.g., starch gelatinization. Results for tortilla chips fried in soybean oil at 190°C showed that the thermal conductivity decreased from 0.23 to $0.09 \text{ W m}^{-1} \text{ K}^{-1}$ [47]. Most of this change occurred during the first 10 to 15 s of frying when most of the water evaporated from the chips and almost all free starch granules gelatinized. Similarly, Sahin et al. [64] found a decrease in thermal conductivity with frying time, which was greater at the beginning of frying and leveled off towards the end. They also found that, for the same frying time, thermal conductivity decreased as frying temperature increased.

Thermal diffusivity is also expected to change during frying, depending up on the change of specific heat and density of the product being fried, in addition to the change in thermal conductivity. For example, for tortilla chips the change was not significant as the decrease in thermal conductivity was balanced by a decrease in specific heat and bulk density [47].

52.2.2 MASS TRANSFER

Dehydration in hot oil at temperatures between 160 and 180°C is characterized by very high drying rates. This rapid drying is critical for ensuring favorable structural and textural properties of the final product. Dehydration in oil inevitably leads to a substantial uptake of oil on or into the potato chips. In the early stages, surface water is lost by a combination of water droplets leaving the food surface and evaporation of boiling water. Water loss is not uniform through the entire food slice. It occurs first on the edges and gradually evolves to the core. Crust formation is a characteristic of fried foods and is closely linked to oil penetration. A dry surface crust is formed and a 103°C isotherm evolves towards the center. It is a complex process in which factors such as temperature of the oil, food weight/frying-oil volume ratio, surface area/volume ratio, length of frying time, oil source, food characteristics, and industrial method used, are involved. Oil adsorption into the voids created by the release of vapor occurs mostly when

the slice is removed from the frying medium as a result of a pressure drop, caused by the cooling of the product. Moreira et al. [50] measured the oil content on the surface and at the core of tortilla chips during frying and cooling, and found that only 20% of the total oil contained in the product was absorbed inside the chip during frying, while it amounted to 64% during cooling. Oil absorption is affected by the porosity of the product. Porosity increases during frying and longer frying times resulted in more uniform pore size distribution [35]. For longer shelf life the moisture content of chips and other products that are dried throughout during frying should be within 2 to 3%, and from the economic point of view, their oil content should be as low as possible. Nutrition studies on fried products are related to oil uptake.

Moreira et al. [49] observed that moisture loss and oil absorption rates were faster during the first 15 s of frying of tortilla chips and became fairly constant as frying continued. Oil absorption as a function of frying time was described by a first-order exponential equation

$$O(t) = O_e[1 - \exp(-kt)] \quad (52.6)$$

where O_e is the final oil content and k is the rate constant, with values varying from 0.223 to 0.330 s^{-1} . The increase of frying temperature from 150 to 190°C increased the moisture loss rate; the oil absorption rate was not affected during the first 15 s, but the final oil content was higher for tortilla chips fried at higher temperature [49,50]. On the contrary, according to Guillaumin [28] a temperature variation between 150 and 180°C had no significant effect on oil absorption.

Baumann and Escher [4] developed a correlation for the quantitative description of oil uptake between frying time t (s), oil temperature T , initial dry matter content DM_r , wet basis (kg/kg), and final moisture content W , wet basis (kg/kg), while the thickness of the potato slices b (mm), was considered to be inversely related to the oil uptake, assuming that the oil is deposited primarily on the surface of the slice. It is expressed as

$$F = C_1 + C_2 t + C_3 T + C_4 DM_r + C_5 W + C_6 \frac{1}{b} \quad (52.7)$$

where F is the fat content of chips, wet basis (kg/kg), and C_1 (kg/kg), C_2 (kg/(kg s)), C_3 (kg/(kg $^\circ\text{C}$)), C_4 (kg/kg), C_5 (kg/kg), and C_6 (kg mm/kg) are constants.

Reddy and Das [61] presented a model for the prediction of moisture M' , and oil O content (wt %),

of chips that include time of frying t (s), oil temperature T (°C), and thickness of slice b (mm), as parameters. It is given by the regression equations

$$M' = 192.42 - 0.426807t - 0.795T + 9.958b \quad (52.8)$$

$$O = -54.98 + 0.21156t + 0.398T - 4.904b \quad (52.9)$$

Rice and Gamble [62] proposed a similar model that can be expressed as

$$\hat{M}^2 = 19.2t + 40T - 6036 \quad (52.10)$$

where \hat{M} is the moisture loss (% initial weight), t is the frying time (s), and T is the oil temperature (°C).

Costa and Oliveira [14], assuming that the potato slice consists of two compartments, proposed the following relationship for the kinetics of water loss during potato frying

$$\frac{m}{m_o} = \alpha e^{-K_c t} (1 + K_c t) + (1 - \alpha) e^{-K_e t} (1 + K_e t) \quad (52.11)$$

where m is the moisture content at time t , m_o is the initial moisture content, K_c and K_e are the rate constants of the mass lumped capacity of the core and the edge, respectively, and α is the fraction of the total water contained in the core of the raw material.

Kozempel et al. [36], proposed the use of Fick's law (for an infinite slab) for the estimation of the moisture content of fried french fries, which depends on the diffusion of water vapor in the material

$$\frac{M}{M_o} = \frac{8}{\pi^2} \times \left[e^{(-D\pi^2\tau/L^2)} + \frac{1}{9} e^{-9(-D\pi^2\tau/L^2)} + \frac{1}{25} e^{-25(-D\pi^2\tau/L^2)} + \dots \right] \quad (52.12)$$

where M is the potato moisture, τ is the residence time, M_o is the initial moisture, L is the nominal thickness of cut potato, and $D = D_o e^{(-E/RT)}$ is the diffusion coefficient, which has to be calculated as a function of temperature due to rapid temperature change that the potato pieces undergo in the oil bath. It should be noted that only the first term is significant.

They also develop the following equation for the estimation of oil absorption, assuming that it depends on the moisture and texture of the potato pieces,

$$\frac{dS}{dt} = [k_o e^{(k_1 TX)} + k_2(M)] \cdot e^{(-E/RT)} \quad (52.13)$$

where S is the concentration of oil (wt oil/wt potato), t is the time (h), TX is the texture (N), k_o is a

frequency factor (g oil/g potato solids, min), k_1 coefficient (N^{-1}), k_2 coefficient (g oil–g potato/g potato solids–g water, min) and T is the temperature of frying oil (K).

Gamble et al. [26] came to the conclusion that oil content is closely correlated to the square root of the frying time. To a reasonable approximation the oil content was twice the square root of the frying time in seconds. The oil content was not directly related to frying temperature, but was more closely related to the remaining moisture present. An overall correlation of all data independent of frying temperature produced a direct relation between oil and moisture content (if not mechanistically linked), which can be expressed as

$$O = 37.8 - 0.483M' \quad (52.14)$$

and stands in the temperature range from 145 to 185°C.

Moreover, they proposed the following equations to correlate moisture and oil content with the frying time t (s), and square root frying time:

$$M' = 62.9 - 0.189t \quad (52.15)$$

$$O = 7.87 - 0.0888t \quad (52.16)$$

$$M' = 100 - 5.34\sqrt{t} \quad (52.17)$$

$$O = -9.70 - 2.53\sqrt{t} \quad (52.18)$$

Krokida et al. [37] presented the following kinetic model for water loss during frying of french fries:

$$\frac{X - X_e}{X_o - X_e} = \exp(-K_X t) \quad (52.19)$$

where X is the potato moisture content (kg/kg db), in frying time t (min), X_o is the initial moisture content (kg/kg db), $K_X = K_{oX} \left[\frac{T}{170} \right]^{K_{XT}} \left[\frac{b}{10} \right]^{K_{Xd}} \left[\frac{1}{2} + \frac{C}{100} \right]^{K_{XC}}$ is the rate constant of moisture content kinetics (min^{-1}), b is the sample size (mm), C is the concentration of hydrogenated oil in total oil (%), $X_e = X_{oe} \left[\frac{T}{170} \right]^{X_T} \left[\frac{b}{10} \right]^{X_d} \left[\frac{1}{2} + \frac{C}{100} \right]^{X_C}$ is the moisture content in equilibrium (kg/kg db), and K_{oX} , K_{XT} , K_{Xd} , K_{XC} , X_{oe} , X_T , X_b , X_C are parameters.

For the oil uptake they proposed the kinetic model,

$$Y = Y_e [1 - \exp(-K_Y t)] \quad (52.20)$$

where Y is the oil content of potato strips (kg/kg db), $Y_e = Y_{oe} \left[\frac{T}{170} \right]^{Y_T} \left[\frac{b}{10} \right]^{Y_d} \left[\frac{1}{2} + \frac{C}{100} \right]^{Y_C}$ is the oil content at infinite process time (kg/kg db), $K_Y = K_{oY} \left[\frac{T}{170} \right]^{K_{YT}} \left[\frac{b}{10} \right]^{K_{Yd}} \left[\frac{1}{2} + \frac{C}{100} \right]^{K_{YC}}$ is the rate constant of oil uptake (min^{-1}), and Y_{oe} , Y_T , Y_b , Y_C , K_{oY} , K_{YT} , K_{Yb} , K_{YC} are parameters.

Oil content of tortilla chips was higher for higher initial moisture content [50].

The oil quality, especially its viscosity and surface tension, affects the oil absorbed in the product. Viscosity increases and surface tension decreases with extended oil use due to the formation of polymers and polar compounds, respectively, through degradation reactions. Consequently, absorption is greater if the oil is extensively used [28]. However, Moreira et al. [50] found that higher viscosity and lower surface tension caused an increased oil adherence on the surface of the product, but not an increased final oil content.

Krokida et al. [38] studied the osmotic pretreatment and came to the conclusion that it affects all the quality attributes for the french fries. Both the mass transfer phenomena (e.g., water loss and oil uptake) that take place during frying get less intense due to the osmotic pretreatment before frying. Color darkness takes place during osmotic dehydration and browning reactions during frying are promoted resulting in more dark and red colored fried product. Salt treated samples have the most acceptable color. The structural properties are also affected by osmotic pretreatment, which can be an effective method to produce low-fat french fries.

52.3 PROPERTIES OF OILS RELATED TO FRYING

52.3.1 THERMAL PROPERTIES

52.3.1.1 Specific Heat

Oils, fats, and their derivatives such as fatty acids at temperatures just above their melting points have specific heats of about $1.9 \text{ kJ kg}^{-1} \text{ K}^{-1}$ or somewhat higher. Published data on the specific heats of fatty acids or triglycerides indicate that, in the liquid state, the specific heat increases with fatty acid chain length and decreases as fat becomes more unsaturated. In the solid state there is little change in specific heat with chain length and a progressive increase with increase of the degree of unsaturation. In both solid and liquid states there is a progressive increase in specific heat as temperature increases [24]. For example, for simple and hydrogenated cottonseed oils the specific heat ($\text{kJ kg}^{-1} \text{ K}^{-1}$) as a function of temperature in degree Celsius was found as: simple oil (iodine value, 108.3):

$c_p = 1.934 + 0.00255T$ for T : 15 to 60°C
partially hydrogenated oil (iodine value, 59.5):

$c_p = 1.989 + 0.00230T$ for T : 40 to 70°C
highly hydrogenated oil (iodine value, 0.85):

$c_p = 1.917 + 0.00406T$ for T : 60 to 80°C

Similar equations have been published by Formo [24] for simple and hydrogenated peanut oil.

According to values tabulated by Rahman [59] specific heat of cottonseed oil varied from $2.176 \text{ kJ kg}^{-1} \text{ K}^{-1}$ at 80°C to $2.489 \text{ kJ kg}^{-1} \text{ K}^{-1}$ at 220°C , and that of soybean oil from $2.063 \text{ kJ kg}^{-1} \text{ K}^{-1}$ at 20°C to $2.502 \text{ kJ kg}^{-1} \text{ K}^{-1}$ at 220°C . The results obtained by Kasprzycka-Guttman and Odzeniak (cited by Rahman [59]) for various oils, in the range of 70 to 140°C , are presented in Table 52.2. Choi and Okos (cited by Rahman [59]) proposed a general equation for specific heat of fats as a function of temperature, T , from -40 to 150°C :

$$c_p = 1.9842 + 1.4733 \times 10^{-3}T - 4.8008 \times 10^{-6}T^2$$

Figure 52.2 presents the change in specific heat of pure triglycerides and natural oils according to data published by Formo [24] and Rahman [59].

52.3.1.2 Thermal Conductivity

Thermal conductivity data for oils, fats, and derivatives are very limited. They are poor conductors of heat and the thermal conductivity values decrease with temperature. For example, thermal conductivity of olive oil is $0.1675 \text{ W m}^{-1} \text{ K}^{-1}$ at 19°C and $0.1611 \text{ W m}^{-1} \text{ K}^{-1}$ at 71°C [24]. Slightly higher values, i.e., $0.1887 \text{ W m}^{-1} \text{ K}^{-1}$ at 5.6°C and $0.1627 \text{ W m}^{-1} \text{ K}^{-1}$ at 100°C , were presented by Polley et al. (cited by Rahman [59]). The effect of temperature increase on thermal conductivity of commercial fatty acids is given in Table 52.3 [24]. Choi and Okos (cited by Rahman [59]) proposed the following equation for thermal conductivity of fats as a function of temperature in degree Celsius, T , from -40 to 150°C :

$$k = -0.181 - 2.76 \times 10^{-3}T - 1.77 \times 10^{-7}T^2$$

TABLE 52.2
Specific Heat Correlation of Edible Oils
as a Function of Temperature

Edible Oil	Equation ^a
Olive	$\text{Coi} = 1.9518T + 0.0006T^2 + 6.2992T^3$
Sunflower	$\text{Coi} = 1.5951T + 0.0097T^2 - 0.0000T^3$
Soybean	$\text{Coi} = 2.1339T - 0.0034T^2 + 0.0000T^3$
Rape	$\text{Coi} = 1.9339T + 0.0005T^2 + 0.0000T^3$
Linen	$\text{Coi} = 1.9460T + 0.0022T^2 - 4.3697T^3$
Castor	$\text{Coi} = 2.0676T + 0.0035T^2 - 0.0000T^3$
Lard	$\text{Coi} = 2.0849T - 0.0043T^2 + 0.0000T^3$

^a T in K.

Source: From Rahman, S. 1995. *Food Properties Handbook*, CRC Press, Boca Raton, FL, pp. 327, 238–264. With permission.

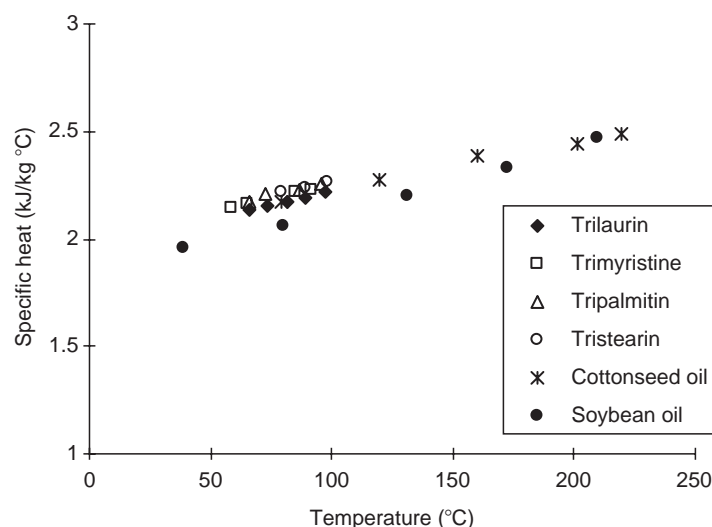


FIGURE 52.2 Specific heat of common pure triglycerides and natural oils.

52.3.2 PHYSICAL AND CHEMICAL PROPERTIES

Fresh oils consist mainly (up to 98%) of triglycerides. Characteristics permitting oil use for deep fat frying are high vapor pressure and resistance to decomposition. Decomposition of oils during frying proceeds through thermal, oxidative, and hydrolytic reactions. The first reactions result in the formation of various oxidized and polymerized products, while the latter in free fatty acids, mono- and di-glycerides. All these substances cause an alteration of not only the

physical properties of the oils, but also the sensorial properties and nutritional value.

52.3.2.1 Vapor Pressure

Vapor pressure, related to boiling point, is one of the most important characteristics for the application of oils to frying processes. Triglycerides of long-chain fatty acids have extremely low vapor pressure, which increases as the fatty acid chain length decreases. Vegetable oils, consisting mainly of triglycerides with long-chain fatty acids, have consequently very low vapor pressure, e.g., 0.001 and 0.05 mm Hg at 254 and 308°C, respectively, for soybean and olive oil [24].

Monoglycerides have considerably higher vapor pressures, and fatty acids are even more volatile. Therefore, these products of hydrolytic cleavage are a source of smoke arising from used frying oil. The vapor pressures or boiling points of some saturated fatty acids are given in Table 52.4 [24]. The logarithm of vapor pressure of fatty acids and typical synthetic triglycerides is a linear function of the reciprocal of the absolute temperature [24]. The heat of vaporization of fatty materials has been calculated from vapor pressure data by means of the Clausius-Clapeyron equation and amounts to 238 kJ kg⁻¹ for oleic and stearic acids at atmospheric pressure and 209 kJ kg⁻¹ for soybean oil at 0.001 to 0.05 mm Hg [24].

52.3.2.2 Smoke, Fire, and Flash Points

The smoke, fire, and flash points of a fatty material are measures of its thermal stability when heated in

TABLE 52.3
Thermal Conductivities of Commercial Fatty Acids

Acid	Temperature (°C)	Thermal Conductivity (W/m °C)
Lauric	72.5	0.192
	90	0.185
	106	0.174
	148	0.138
Oleic	72.5	0.188
	90	0.178
	106	0.155
	148	0.115
Palmitic	72.5	0.172
	90	0.157
	106	0.138
	148	0.102
Stearic	72.5	0.160
	90	0.146
	106	0.132
	148	0.096

TABLE 52.4
Boiling Points of Saturated Fatty Acids

Pressure (mm)	Boiling Point (°C)						
	Caproic	Caprylic	Capric	Lauric	Myristic	Palmitic	Stearic
1	61.7	87.5	110.3	130.2	149.2	167.4	183.6
2	71.9	97.9	121.1	141.8	161.1	179.0	195.9
4	82.8	109.1	132.7	154.1	173.9	192.2	209.2
8	94.6	121.3	145.5	167.4	187.6	206.1	224.1
16	107.3	134.6	159.4	181.8	202.4	221.5	240.0
32	120.8	149.2	174.6	197.4	218.3	238.4	257.1
64	136.0	165.3	191.3	214.6	236.3	257.1	276.8
128	152.5	183.3	209.8	234.3	257.3	278.7	299.7
256	171.5	203.0	230.6	256.6	281.5	303.6	324.8
512	192.5	225.6	254.9	282.5	309.0	332.6 ^a	355.2 ^a
760	205.8	239.7	270.0	298.9	326.2 ^a	351.5 ^a	376.1 ^a

^aValues obtained by extrapolation.

contact with air. The smoke point is the temperature at which smoking is first detected in a laboratory apparatus protected from drafts. The flash point is the temperature at which the volatile products are evolved at a rate sufficient to be ignited, but not to support combustion. The fire point is the temperature at which the volatile products support continued combustion.

Fatty acids are much more volatile than glycerides; therefore, smoke, flash, and fire points of oils depend principally on the content of free fatty acids, and decrease if hydrolytic degradation is extended during frying. The smoke point of cottonseed, corn, or peanut oils, e.g., decreases from about 232°C at a free fatty acid content of 0.01% to 93°C at a free fatty acid content of 100%. The unsaturation of oil has hardly any effect on its smoke, flash, and fire points.

52.3.2.3 Viscosity

The viscosity of an oil is strongly affected by the degradation during frying, increasing as a result of the formation of dimers and polymers through thermal and oxidative degradation [5,11,55]. Keeping the oil at frying temperature caused an increase of viscosity with time, which was even higher when the oil was aerated [12]. The increase observed during continuous heating of the oil was higher than that during frying at the same temperature (27% compared with 11% after 90 h at 185°C [11]. Experiments of potato frying with corn oil showed that after 36 h of frying, when smoking and foaming of the oil indicated that it was no longer suitable for frying, its viscosity had increased by 28.6% [27].

52.3.2.4 Surface Tension

Fresh oils consist mainly of triglycerides and are immiscible with water, which is the main constituent of all foods. The surface tension between oil and food reduces by the action of polar compounds that are formed through degradation reactions and leaching of food materials into the oil during the frying process. As a result, oil and food contact increases and heat transfer becomes more effective. Blumenthal [5] observed that the contact time of fresh oil at the food surface is only about 10% of the food immersion time. As the oil continues to be used, the contact time increases to about 20%, then optimally to about 50%. This causes normal absorption of oil into food and proper cooking of the exterior and the interior. If the oil degrades further, more surfactants are formed, causing an increased contact time (80% or higher). As a result, an excessive oil pickup by the food and an increased rate of heat transfer to the surface of the food is observed, while conduction of heat to the interior cannot be sped up by changes in the oil. Thus, excessive darkening and drying at the surface occur.

Sodium oleate, a water-activated surfactant, was isolated from the polar fraction of frying fat [6,27]. Other water-activated surfactants are phospholipids and inorganic salts. Lipid-activated surfactants include low-polar thermal polymers and high-polar oxidized compounds. The mono- and di-glycerides formed through hydrolysis also cause a rapid drop of the interfacial tension between oil and water [23]. The oil-air surface tension was found to decrease by 2.20% after 36 h of potato frying [27]. At high

surfactant concentration, the reduction of the interfacial tension causes the formation of persistent vapor bubbles at the air/oil interface, or in other words, foaming.

52.3.2.5 Color

Carotenoid and chlorophyll pigments are responsible for the color of the fresh oils. Usually color is one of the properties most frequently determined at the purchase of the oil in the frying industry [53]. The color of frying oil darkens with use and eventually affects the color of the fried product. Perhaps, the color change is the most noticeable of all changes occurring during frying. Food introduces various components to the oil during frying. Many of these components contribute to color formation by reacting with oil or its degradation products. Maillard browning reactions are believed to be the source of some of the color development [32,54]. Since Maillard reactions are greatly accelerated by increasing temperature, dark coloration at the surface of the fried food and in the frying oil can readily occur. Furthermore, aldehyde groups of triglyceride molecules, derived from lipid degradation react with amino groups to form dark coloring compounds. However, a darkening is observed even when the oil is simply heated at frying temperature, without the addition of any food. In fact darkening during heating is more intense than that developed during frying [11,25].

52.3.2.6 Alteration of Frying Oil

During the frying process the lipid material is exposed to high temperature in the presence of air and moisture. Under these conditions a wide variety of reactions takes place and the main groups of alteration compounds generated are summarized in Table 52.5 [11,25,7,16].

The rate of the oxidation reactions is higher for the fatty acids that contain more double bonds. However, even saturated fatty acids undergo oxidation. The relative reaction rates of oleic (C_{18:1}), linoleic (C_{18:2}), and linolenic acid (C_{18:3}) with oxygen are 1, 10, 25, respectively [18]. Therefore, the inherent stability to oxidation, expressing the relative reaction rate with oxygen of oil depends on the fatty acid content of its triglycerides and can be calculated as [18]:

$$\text{Inherent stability} = (\% \text{ oleic acid} \times 1 + \% \text{ linoleic acid} \times 10 + \% \text{ linolenic acid} \times 25)/100$$

The primary oxidation products are hydroperoxides, which decompose quickly at frying temperature and cannot be detected. The decomposition products include several volatiles. The volatile compounds increase at the beginning of frying until they reach a plateau, as a result of simultaneous evaporation, decomposition, and formation [11,55]. Other secondary products formed are nonvolatile polar compounds, e.g., dimers and oligomers. The content of polar compounds increases linearly with frying time [29,30,43,65] and may be considered as the most representative index of oil degradation during frying [25,63,71]. Several countries have established 25 to 27% total polar content as the regulatory limit for use of the frying oil [22]. The polar content increase is accompanied with interfacial tension decrease, as polar materials act as surfactants and viscosity increase, since dimers and oligomers are formed. Thus, the decrease in surface tension was highly correlated with the increase in viscosity [27], and a higher fat absorption in frying food was observed as the viscosity of the frying oil increased [55]. The viscosity increase is not only due to the polymers formed through secondary reactions of hydroperoxides, but

TABLE 52.5
Main Groups of Compounds Formed during Deep-Fat Frying

Type of Reaction	Causative Agent	Resulting Compounds
Oxidation	Oxygen	Oxidized monomeric triglycerides Dimeric and oligomeric triglycerides Volatile compounds (e.g., aldehydes, ketones, alcohols, esters, hydrocarbons, and aromatic compounds)
Thermal degradation	Temperature	Cyclic monomeric triglycerides Nonpolar dimeric and oligomeric triglycerides
Hydrolysis	Moisture	Fatty acids Monoglycerides Diglycerides

also due to polymers formed through thermal alterations.

The polyunsaturated fatty acids upon oxidation tend to form conjugated structures. Thus, the conjugated dienes content increases with frying time, when oils rich in linoleic acid are used [25,30,71] and conjugated trienes are also formed when linolenic acid is present.

The rate of all oxidation reactions increases with temperature. However, the increase from 155 to 195°C was moderate [30]. The elimination of air, on the other hand, seems to affect greatly the degradation reactions [55].

Products formed through hydrolytic cleavage, i.e., free fatty acids, mono- and di-glycerides increase at the beginning of frying and remain almost constant afterwards [30]. Normally 0.5 to 0.8% free fatty acids content is encountered in frying oils [55]. Higher contents may cause extensive smoking of the oil, but usually the simultaneous formation and evaporation of the acids leads to a plateau.

52.4 INDUSTRIAL FRYING SYSTEMS

52.4.1 EQUIPMENT

The equipment used in industrial frying consists mainly of continuous, highly automated deep-fat fryers. Large batch frying systems are also included in the industrial frying classification. The first continuous fryer was developed in 1929 for the production of potato chips. A wide variety of industrial fryers followed, suitable to produce every type of fried food from savory snacks, doughnuts, and vegetables to breaded meat and seafood. The more widely used are the traditional fryers for chips and several snacks, but pressure fryers, vacuum fryers, shaped product fryers, and a number of others are also commercially available in a variety of sizes and configurations. They have production capacities from a few hundred kilograms per hour for some snacks to tens of tons per hour for par-fried french fries.

Continuous deep-fat frying systems consist of five independent sets of equipment: (1) an oil tank where the food to be fried is submerged; (2) a conveying means for moving the food into, through, and out of the oil; (3) a heating unit; (4) an exhaust system to remove the vapors emerging from the oil tank; and (5) an oil recirculation system, which pumps the oil through filters (and external heaters if used) and replenishes it from a bulk supply to maintain a steady volume in the tank. Banks [3] designated six zones or fryer areas in an industrial fryer used for snacks production, where product handling and processing conditions are controlled to regulate frying:

1. *The entry zone* where the food is rapidly heated and is covered in a few seconds with small steam bubbles, more or less uniform in size and distribution. Bubbles keep the food pieces separated and, along with the flow of oil, prevent clump formation. Mechanical pressure or even pressure due to excessive oil flow that would force the food pieces into close contact must be avoided in the entry zone. The food must move freely with the aid of a smooth oil flow. At the end of the entry zone a “breaker bar” or any other appropriate device may be used for the separation of any clumps that have been formed. Starch, in starch containing foods like savory snacks and potato products, begins to gelatinize in the entry zone, first on the surface and later through to the interior.
2. *The case-hardening zone* where the bubble pattern on the food surface changes, with some areas emitting larger and more frequent steam bubbles and others smaller and at a slower rate. The larger bubble sites mark surface points where steam channels have erupted due to steam pressure generated in the interior of the product. The steam pressure increase may cause intercellular separation and formation of channels that break through to the surface or formation of blisters in the interior of the product. The smaller bubble sites reflect the dehydration of cells close to and on the surface. An outer layer of a crust begins to form as a result of this dehydration and food begins to case harden. When case hardening is complete food pieces have less tendency to stick together, but they are still pliable and malleable.
3. *The shape-firming zone* where the larger bubble sites on the food surface continue to emit steam, while the smaller bubble points further diminish. The thickness of the crust layer increases and cells below the crust dehydrate and become fixed. Crust thickness and subsurface structure depend greatly on frying temperature. Higher temperatures result in a thinner crust layer. The original subsurface cellular structure, interlaced with some voids, is generally maintained, although cell walls are distorted, in nominal frying conditions. When a temperature profile, typical of kettle frying, is used, the original cellular structure is almost completely destroyed and a porous, cavernous structure is formed.
4. *The cooking or moisture reduction zone* where the food having a nonsticky surface and a relatively firm shape continues to lose moisture through steam emission and cooking. The

temperature that was maintained slightly over the boiling point of water until now begins to rise, initiating the final phase of frying. Usually there is a need to keep the products under the surface of the oil in this zone by a submerger belt.

5. *The finish frying zone* which starts as the product approaches the end of the submerger belt and continuous until the product is removed from the fryer by the carry out conveyor. Finish frying is very important for the quality of the product and must be carefully controlled. If the product is removed from the fryer before finish frying is completed it would have poor flavor, light color, poor textural characteristics, and low oil content. It is the final dehydration and temperature rise that support chemical reactions generating much of the flavor and color of the products and create the desired texture.
6. *The takeout zone* where the product is moved out of the oil by a conveyor. The product is still hot in that zone and continues to cook. Its surface is covered with oil and has a wet appearance. As the product cools, the residual water vapor condenses, creating a pressure difference that pulls the oil from the surface into the pores and leaves the surface almost dry. Handling of the product at the takeout zone influences its moisture, texture, flavor, color, and especially oil content. To increase oil uptake, additional oil can be sprayed on the hot product, to be absorbed as the product cools. On the contrary, if low oil content is desired the product has to be maintained at temperature as it leaves the oil bath and the surface oil can be removed by a stream of hot gas. Some of the early commercial processes utilized a stream of hot air to reduce surface oil, but caused rapid oxidation and decrease of product shelf life. To avoid oxidation, units developed later use dry steam in conjunction with an oxygen-free chamber [3]. Centrifugal force is also used in some operations to reduce oil content of the products.

52.4.1.1 Oil Tank

The oil tank is usually constructed of stainless steel and may have a rectangular shape or a V-shaped bottom. The V-shaped bottom facilitates settling and removal of sediment. Tubes or electrical elements may be placed in the tank, if an internal heating system is used. The thermal insulation of the tank is a very important feature since it improves fryer efficiency and lessens the risk of burns to those making

casual contact. Some metals are strong catalysts of oil oxidation. Their catalytic activity varies following the order: copper > copper alloys > iron > zinc > stainless steel > magnesium. Copper and copper alloys are detrimental for oil deterioration; therefore, all metal surfaces that come in contact with the oil should not contain any of these metals.

The size of a fryer varies depending on the finished product output, which may range from 100 to 5000 kg/h for potato chips and other snacks, or 14,000 kg/h for par-fried french fries [3]. The proper sizing of the fryer is very important. The best quality products are produced when fryers operate at full capacity and on a continuous basis. Operating the fryer outside of its designed capacity can cause both unacceptable product quality and equipment failures, with risks increasing in proportion to the deviation from design capacity. Operating the fryer below the rated capacity results in unnecessary quantities of frying oil being kept heated and reduction of oil turnover, which increases oxidative deterioration. Therefore, extended operation below the rated capacity affects negatively the frying oil quality and the flavor, nutritive quality, and shelf life of the fried product. Operating above the rated capacity changes the frying temperature profile and may result in undercooked products with a bland unacceptable flavor. Also, it forces extensive heat input, which can lead to early equipment failure [3,10].

Apart from the conventional frying process, some frying systems have a short frying step followed by a finish-cook step in an oven with steam or hot air. In impingement ovens, high-velocity hot air is forced against the food at 90° angle [32]. In a reverse situation, microwaves or steam have been used for the initial cook, followed by a finish-frying step. One of the main goals in this procedure is to reduce oil absorption in the food.

52.4.1.2 Conveying System

The food to be fried is carried into and through the oil tank by an appropriate conveyor. Food can also be fed by slow-moving paddles and it either sinks to the submerged conveyor or, if it floats, is held below the surface by a second conveyor. The conveyor speed controls the frying time. The fried food is removed from the fryer by an inclined conveyor and excess oil is drained back into the tank.

Matz [41] has classified the following conveying mechanisms: (1) spacer bar conveying, in which the frying piece floats between transverse bars that push it over the surface at a rate adjusted to give it the desired cook by the time it emerges from the end of the tank; (2) drop plate conveying, in which the product rests on shelves moving below the fat surface until

the pieces develop enough buoyancy to rise to the surface where they contact an upper conveying means; (3) hold-down conveying, in which a mesh band moving just underneath the surface of the oil contacts the upper surface of the buoyant product and carries it along; (4) restrained conveying, in which products are carried between two horizontal mesh belts; (5) conveyors for nonfloating products, traveling well below the surface of the oil and utilizing baskets, belts, or other holding devices having no covers; (6) compartmentalized conveying, in which dough pieces are held in small molds that may help to shape the product as it cooks (e.g., fabricated chips); and (7) conveying by the current of the oil that is being created by pumps. Some other types also have been devised to transport foods requiring special treatment. An example is the rotating drum conveyor described by Bullock (cited by Matz [41]), in which the drum has internal shelves. During part of the drum's rotation, the food is immersed into hot oil.

Some products, e.g., doughnuts, must be turned over a little more than half way through the frying process in order to get a symmetrical shape and a uniform crust color. For these products conveyors that automatically turn the dough pieces at the proper time are used.

The food may be transferred out of the fryer by an inclined conveyor or some more complex mechanism. Usually foods have to be cooled before packaging in order to become firmer and to slow moisture migration from the crumb to the crust for those products that retain a moist interior.

Improved product quality is closely related to appropriate gentle handling through the fryer. Contemporary fryers often feature interchangeable inlet sections produced in a cartridge design, which promote simple and rapid product changeover and easy height adjustment to support various types of transfers. Developments in hold-down conveyors permit an easier vertical adjustment to provide the needed gap to ensure both product control and minimal damage from the belting. To minimize belting damage conveyors with Teflon slats that present a nearly smooth, nonstick surface are now available [68].

52.4.1.3 Heating Units

Fryers are heated by electricity, gas, fuel oil, or steam. Dual-fuel systems convertible from fuel oil to gas are available. A direct heating system for heating the oil in the tank may be applied, or an external one by means of a heat exchanger.

The direct heating systems include bottom-fired gas strip burners, tubular heating systems, or electrical heating elements installed inside the tank. In

bottom heating systems a problem may arise from the accumulation of sediments on the bottom of the vat and formation of an insulating layer between the heat source and the oil. Moreover solids are overheated and deteriorate quickly affecting the frying oil quality. This situation can be overcome to a certain extent by constructing the tank with a V-shaped bottom, so that the solid material accumulates on the lowest part of the vat, below the level at which heat is being applied.

The tubular heating systems transfer heat by passing burning gases, thermal oil, or steam through tubes that run through the vat from side to side. One advantage of tubular systems, as compared with direct bottom heating is that there is a cooler zone below the heating tubes, where solids settle away from the high heat and turbulence so that they do not interfere with heat transfer and do not deteriorate very fast. The V-shaped bottom facilitates settling and removal of solid sediment in this case, as well. Another advantage is the large heating surface that enables a sufficient heat transfer rate, at a lower temperature difference between the surface of the tubes and the surrounding oil. As a result of overheating of the oil, generation of burnt deposits on the heating surfaces, and quality deterioration are diminished. Disadvantages are the greater difficulty of cleaning tubes and the need for a larger quantity of frying oil for the same output.

Electrical heating systems are convenient, but they are used mostly in batch fryers, or in small continuous systems, due to higher energy cost. Electricity is used to heat resistance elements inside small stainless steel tubes assembled in grids. These grids are fixed in the fryer, just above the bottom of the tank and can be raised out of the tank to make cleaning easier. Electrically heated fryers have the advantage of requirement of smaller amount of oil than the tubular ones. Also it is very easy to create different temperature zones inside the fryer by the appropriate position of the heating elements.

Microwave heating, as an additional source to conventional heating, has been claimed to offer advantages to the cooking of some products, e.g., doughnuts [68]. Microwave radiation is applied when the doughnuts are fried on their first side and result in a faster increase of the temperature inside the doughnut and consequently in a faster expansion. The total volume of the product is larger and fat absorption may be reduced up to 25%. A patent of a short-wave infrared heated conveyor, simulating contact frying, was proposed for pizzas and pancakes by Dagerskog et al. (cited by Matz [41]).

External heat exchangers are of direct or indirect heating type. Direct heating is accomplished by gas, oil, or propane burning. Indirect systems use chlorinated

hydrocarbon heat transfer fluids or steam. The external heat exchangers allow the fryer tank to be shallower and, therefore, a smaller amount of oil is needed in the tank, but a considerable amount is needed to fill the heat exchanger and the piping to and from it. A major advantage of the external heat exchangers is the more uniform temperature of the oil in the fryer, as compared with heating systems installed in the tank. The initial cost of external heating is higher [41], but a longer equipment life, an improved product quality, a higher safety, and environmental protection are achieved according to Swackhamer [68]. For fryers with multiple temperature zones, oil can be pumped from an external heat exchanger through separate pipes at a different flow. The flow into each separate zone is controlled individually to maintain the preset temperature of the zone.

Heat and oil recovery systems are used to reduce energy and oil costs. The exhaust vapor escaping from the vat is used to preheat incoming food or oil. If direct contact of this vapor with the incoming food is applied some of the oil carried by exhaust vapor is absorbed in the food reducing the amount of oil that is required in the fryer. Alternatively a heat exchanger may be mounted in the exhaust hood to recover heat and use it to preheat incoming food or oil, or to heat process water. Oil recovery systems are also used to remove the oil entrained in exhaust vapor and to return it to the oil tank. In addition, pollution control systems prevent smoke and other volatile products of oil degradation from being discharged to the atmosphere, by feeding the exhaust air into the burner used to heat the oil. By appropriate controlling, thermal efficiencies of 80 to 90% can be achieved [21].

52.4.1.4 Exhaust System

A large amount of vapor is generated by the fryer and contains droplets of oil, volatile products formed through oil degradation, and gases given off by the food. Therefore, exhaust systems are installed in industrial fryers to collect and dispose the hot gases in a safe and sanitary manner. The fryer may be completely enclosed in the exhaust system, or an overhead hood may be used. The exhaust gases are not emitted directly to the environment. Several oil and heat recovery systems are used, as mentioned above.

Oil droplets entrained in the exhaust vapor are usually separated by cyclones. Water spray is mixed with the hot gases as they enter the centrifugal blower, where violent turbulence disperses the water and throws the oil and water droplets to the blower walls by centrifugal force. Separation is completed in a tangential entry centrifugal separator that follows

the blowers. Volatile organic compounds are directed to the burner used to heat the oil, as already mentioned, or to a combustion chamber if such a burner is not used.

52.4.1.5 Recirculation of the Oil and Oil Turnover

Industrial fryers are equipped with pumps for continuous recirculation of the oil through external heaters and filters. Fresh oil (makeup oil) is added automatically to replace the oil absorbed by the food and maintain the desired level in the fryer.

Oil turnover is usually expressed as the number of hours required for the fresh oil added during frying to equal the total quantity of oil in the frying system, and is calculated as:

$$\text{Oil turnover} = \frac{\text{total weight of oil in the frying system, kg}}{(\% \text{ oil in product})(\text{fryer output, kg/h})}$$

Industrial fryers commonly have oil turnover rates ranging from 5 to 10 h, with some fryers ranging a few hours higher [3]. After one turnover, one half of the original oil remains in the frying system and is further halved with each subsequent turnover. Usually equilibrium in the oil analytical values is achieved after four turnovers and reflects blending of the fryer and fresh oil at a constant rate. A low value of oil turnover indicates a reduced stress of the oil during processing and is an important consideration in fryer selection together with the type of oil, the fried product, and the operating conditions. Some frying systems operate successfully with an oil turnover rate of 12 h, while others, involving different products and fryers, require a turnover rate <7 h [3].

52.4.1.6 Filtration Systems

Screens and filters are used to remove solid particles and to extend the life of the frying oil. Screens, located ahead of the oil circulation pumps, to remove large particulates and to protect heat exchanger tube bundles from becoming clogged, have been widely used in industrial frying systems. These basic crumb-removal screens do not remove small particles from the oil. The particles that sink to the bottom may be carried out of the tank by a screw conveyor or special augers [41,51]. The discharged material is pumped together with surface oil to filters. The particles that remain in suspension are effectively removed by more complex multistage filters [68].

Passive filtration with possible addition of an inert filter aid such as diatomaceous earth has been a common practice in frying industry for several years

[8,74]. Passive filtration may be defined as the operation or removal of particulates with an inert filter medium. However, within the last 15 y demands for better quality along with the augmenting use of vegetable oils, have created the need for filter absorbents, which aid in the removal of degradation products. This type of filtration is designated as active filtration. In processing of foods that absorb a significant amount of oil, e.g., potato chips or doughnuts, the oil turnover value is low enough, and the oil quality never deteriorates to the point where the oil must be discarded. Passive filtration is sufficient for this type of operation. On the contrary for high values of turnover, degradation products may accumulate and active filtration may help in slowing the rate of oil decomposition.

Many natural and synthetic absorbents have been promoted for use in active filtration of frying oils [72]. Processors concern free fatty acids as one of the main degradation products that they want to keep at a low level and evaluate absorbents relevantly. Natural absorbents in combination with diatomaceous earth were judged effective to control free fatty acids, off-flavors, smoke, and foaming, without negatively affecting the flavor of the oil [8]. Nevertheless, an increased concern of producers is observed about the level of total polar compounds, oxidized compounds and polymers in fried oil and consequently in fried food. The different absorbents have different selectivity towards each group of degradation products. Alumina absorbs free fatty acids, while it is almost inert for other polar compounds [73]. Silica and active silica are effective in absorbing free fatty acids and mono- and di-glycerides, and less effective for other polar compounds [42,72]. Activated carbon is effective in removing saturated and unsaturated carbonyl compounds [42]. Synthetic magnesium silicate absorbs more polymeric substances and other polar compounds than silica absorbents [73]. Mixtures of activated carbon and activated silica were effective in removing cholesterol oxidation products too [74]. Also, mixtures of commercial absorbents, based on calcium silicate, magnesium silicate, and porous rhyolite removed free fatty acids and polar compounds and improved the stability of the frying oil [40].

Filtration can be distinguished in two types: (1) absolute and (2) depth. In absolute filtration a steel or plastic screen, cartridge, or paper filter removes particles from the oil. These filters have certain porosity and most of them retain particles down to 5 μm , when used for passive filtration [72]. If diatomaceous earth or other absorbents are used, they are mixed with the oil in the fryer at the end of the day and/or shift and the suspension is pumped through the filter

and circulated through the filter and back into the fryer until a filter bed is established and the fryer is free of fines and particulates. Filter media suppliers recommend a dose between 0.1 and 2.0%, by weight on the frying medium [8].

In depth filtration a thick layer of porous material with irregular channels is used. A depth filter can be a pad made of thick nonwoven fabric, or a cake of filter aid type. Depth filters may retain particles down to about 1 μm [72] and do not blind as easily as absolute filters because they have a greater filtering area. Diatomaceous earth, or perlite is used as inert powders in depth filters. Active absorbents are also incorporated in filter pads, e.g., bonded on refined cellulose fibers [74].

52.4.2 PROCESS TEMPERATURE AND TIME

The necessary frying time for the product to obtain surface color, crispness, and organoleptic quality depends greatly on oil temperature. It seems that there is no optimum time-temperature combination for each product and depends on the size and configuration of the product and its ingredients.

For potato chips the use of two different frying temperatures provides an example of how temperature can affect product attributes. If potato slices are fried at 185°C, a frying cycle of approximately 3 min is required and a delicate texture and a thin crust are obtained. If a temperature of 167°C is used, the frying cycle is extended to about 7 min and the chips have a harder texture and a thicker and firmer crust. In general the lower the temperature and the longer the process time, the slower the cooking rate of the product, the firmer the crust and texture, and the higher the oil absorbed by the product. Products, e.g., fabricated chips or various snacks, made from dough with low moisture content (i.e., 40%) are often fried at higher temperature (200 to 205°C) for a very short time amounting to a few seconds. Under these conditions a small quantity of oil is absorbed in the products and an additional amount is sprayed on them at the exit of the fryer, if higher oil content is desired. Tortilla chips, which also have moisture content around 40% before frying, are fried at approximately 190°C for 60 s and the oil content of the product amounts to about 20% [35].

For products that expand during frying, e.g., doughnuts, the frying temperature is critical for maximum expansion. Temperatures in the range of 190 to 202°C have been recommended for yeast-leavened doughnuts [41]. Higher temperatures within this range may encourage greater expansion, but excessive temperatures firm up the skin too fast so that maximum expansion cannot occur. Frying at temperatures

that are too low allows too much fat to penetrate the crust before the interior is cooked. Optimum frying times in the above-mentioned temperature range are 110 to 120 s for round doughnuts weighing 1.5 oz.

Other products, e.g., pies, may present bursting problems before they are sufficiently cooked, if not fried properly. Some times temperature combinations have been proposed for various sizes of pies ranging from 180°C for 225 s for smaller pies to 195°C for 255 s for larger ones [41].

Par-fried french fries are usually subjected to a fast frying for approximately 30 s at 175 to 185°C. A partial drying of the raw material may be applied after blanching, to adjust the solid content and to add to the crispness of the final product.

A temperature variation along the fryer has been increasingly recognized as offering certain advantages for some products. A higher temperature may be applied at the entry zone of the fryer. Another such zone may be needed in the area where pieces are turned over, in doughnuts fryers.

52.4.3 FRYER OPERATION AND CONTROL

The highest quality and most consistent products are produced when fryers are operated at full capacity on a continuous basis, stopping production only for schedule maintenance and cleaning at the end of each week. However for business and staffing reasons many business use a daily schedule, therefore, appropriate oil and fryer management programs must be carefully developed and utilized to produce a competitive product.

At fryer start-up, preheating of the oil at 100°C is usually required and maintaining at that temperature until moisture is completely removed. Heating to process temperature and starting production must follow immediately in order to minimize oil degradation. Any interruption in production, when the oil is circulated at frying temperature, creates a high-risk period for the oil and finished product. According to Banks [3] in industrial operations, increased consumer complaints are correlated with product made following interruptions in production exceeding 20 min. Each interruption should be managed as a potential problem. When it is apparent that an interruption will last longer than a few minutes, a partial fryer shutdown should be initiated by stopping the addition of makeup oil and heating, while continuing oil circulation. After the last product is removed from the fryer, the conveyor system should also be stopped to minimize oil aeration. The same steps are followed at the end of production. The proper time to stop heating is determined so that the oil is at the minimum frying temperature as the last product leaves the fryer. Oil

circulation is not stopped until the temperature reaches the allowed limit, which is usually defined by the manufacturer. If production is restarted the following day, the oil is left in the fryer; otherwise it is better to be transferred to a storage tank in order to minimize oxidative degradation.

Contemporary industrial fryers are computer controlled. In the absence of computer control a set of systematic operating procedures is developed. Energy requirements and operating parameters are determined for the full range of variations in raw material type and feed rate, and translated into fryer settings. During production, operating parameters can be adjusted to the predetermined settings as changes in raw material occur.

52.4.4 FRYING MEDIUM

The products used for frying range from nonhydrogenated refined fats and oils to hydrogenated products specifically designed for frying. The selection of an oil or fat is influenced by many factors as the inherent stability to oxidation, the fried product and especially its flavor and eating characteristics, the cost, the fryer design, and the historical usage.

Quite often the most important factor, apart from cost, is the inherent stability to oxidation since it affects the quality and shelf life of the finished product. The inherent stability of a frying oil decreases with increasing content of polyunsaturated fatty acids and can be calculated as described in Section 3.2.6. This is the reason why cottonseed oil is considered the first vegetable oil for deep frying and why several fats, e.g., palm kernel, are used in mixtures specially prepared for frying. Olive oil also has a good inherent oxidative stability, due to high content in oleic acid, which is monounsaturated and low content in polyunsaturated fatty acids, but it may have a smoking problem due to its nontriglyceride constituents [17,18].

Pourable frying oils are popular because of their convenience in handling. They range from clear to opaque fluids at room temperature. Refined, bleached, and deodorized (RBD) vegetable oils, e.g., soybean, cottonseed, corn, and canola may be used, but they are not stable to oxidation. Therefore, partially hydrogenated vegetable oils are preferred and they are frequently used in blends that may contain a small amount of solid fat. Also new varieties of oils with low content of linolenic acid and high content of oleic acid are promoted for use in frying [44,10]. Clear liquid frying oils are also made by fractionation of hydrogenated oils. Cottonseed oil, hydrogenated soybean or canola oil, and palm olein are mostly used in liquid frying oil formulations.

A wide variety of plastic or solid frying fats, e.g., RBD palm, palm stearine, coconut, palm kernel, lard, and tallow are also used in frying. They are mainly blended with RBD oils, hydrogenated oils, or with each other to prepare heavy-duty shortenings. Lauric acid oils (e.g., palm kernel and coconut) can be blended only with each other and not with fats and oils containing longer fatty acids, because in the last case an early onset of excessive foaming is observed [17,18]. Although solid fats and hydrogenated oils generally present a higher oxidative stability their use may be limited by the sensorial characteristics they impart in fried foods. A shortening with high solid content and elevated melting point creates a waxy surface of the product and an unpleasant mouthfeel. On the other hand, a greasy surface may be created with a shortening of very low solid fat content. Another cause that has limited or even excluded animal fats from frying blends is their content in saturated fat and cholesterol, which were linked to increased risks of coronary disease. Consequently consumers concern has forced the industry and food-service processors to switch to all vegetable blends containing partially hydrogenated oils [8]. Erickson [17,18] presents the properties of some oils and blends that are suitable as frying media in different applications.

Apart from the properties of the bulk oil, its frying performance is evaluated by frying tests and analytical measurements when the frying system has reached steady-state conditions. The fresh oil placed in the fryer is free of degradation products and may contain some antioxidant additives, therefore, analytical measurements conducted before the system reaches equilibrium lead to misleading results. In most industrial frying operations, the status of the oil approaches equilibrium after four oil turnovers. The analytical values recorded usually include color, flavor, viscosity, and free fatty acids. If analytical laboratory facilities are available measurements more representative of the oxidative status of the oil are executed, e.g., polar compounds content, anisidine value, polymers, oxidized triglycerides, and conjugated dienes. Especially the polar compounds content determination is practiced by many industries because an upper limit is allowed by the legislation of several countries [22].

The storage of the bulk oil is another very important factor, which affects its quality. The storage tank must be sized properly to assure rapid oil turnover and frequent fresh oil supplies, e.g., on a one- or two-week basis. Peroxide value (PV) determination can be conveniently used to monitor storage oil quality. PV on receipt should be below 1.0 meq/kg, preferably below 0.5. By avoiding temperature increase, aer-

ation, and contamination, PV should be maintained below 1.0 and 2.0 meq/kg at the end of first and second week, respectively. Longer storage times may require additional protection by nitrogen to depress oxidation.

52.5 PRODUCT QUALITY AND SHELF LIFE

The sensorial attributes of a fried product, i.e., flavor, color, and crispness, depend on the raw material, the frying medium, and the frying conditions. The frying medium is absorbed in the products and, therefore, is an important ingredient of some of them, especially chips and other snacks. Consequently, the nutritional quality of the fried product depends on the quality of the fried oil. The compounds produced through the oil deterioration occurring during frying, are uniformly distributed in the oil remaining in the fryer and in the oil absorbed by the food [28]. If the oil has deteriorated extensively a product of unacceptable quality will be produced.

During storage, the quality of the fried snacks mainly depends on the deterioration of the absorbed oil, which continues under storage. This deterioration is accompanied by the formation of off-odors, and products that may cause some health problems. The oils undergo oxidative and hydrolytic alteration under storage, similar to the ones occurring during frying, although the reaction rates are much slower. Hydrolytic alterations are caused by lipase enzymes in the presence of moisture. Therefore, preservation of foods from hydrolytic rancidity can be achieved by the control of processing conditions, in order to inactivate enzymes, and to maintain low moisture during storage. Oxidative alterations during storage are the most important of the two mechanisms, with respect to food acceptability. These alterations are caused by oxygen attack and are accelerated by temperature increase and presence of light. Preservation from oxidative rancidity can be achieved by storage at lower temperature, elimination of oxygen and light, or addition of antioxidants. Several synthetic and natural antioxidants have been added either to the frying oil or to the final fried product. It should be reminded that the oils that contain higher amounts of unsaturated fatty acids deteriorate faster and, therefore, need a higher protection against oxidation.

Another important quality defect of fried snacks is the reduction of crispness during storage due to absorption of moisture. In addition, moisture gain accelerates rancidity. Data concerning the effect of water activity (a_w) on oxidation rate are presented by Labusa [39]. As a_w increases above the monolayer, the reaction is slowed up to the value where intermediate moisture

foods begin (i.e., approximately 0.4). A further moisture increase tends to increase the rate of oxidation. Data on water activity that results in loss of crispness to an extent where the chips are perceived as not acceptable are not available, but Labusa comments that it probably occurs at an a_w of 0.4 to 0.5.

Potato chips are fried to a final moisture level of 3% or less, which accounts to an a_w of about 0.2. Tortilla chips, corn chips, and other snacks have a final moisture content of <3%. From a study on potato chips a moisture content of 3.57% was the point at which a taste panel rejected the chips because of their texture deterioration [39]. Moisture absorption and the consequent deteriorative changes can be prevented by the use of packaging materials with appropriate moisture barrier properties. In this case the shelf life of the products can be extended to several months. If the packaging material is not transmitted by light and if the air inside the packaging is replaced by an inert gas, e.g., nitrogen, a considerably longer shelf life can be obtained.

Fried nuts are subjected to deterioration, similar to potato chips. Their shelf life can be extended by the use of appropriate packaging too. Long-term storage up to 1 to 2 y can be achieved by vacuum or inert gas packaging, while it can be extended to 3 y by storage under refrigeration. Of particular concern with nuts, especially peanuts, is the potential presence of aflatoxins. Aflatoxins are carcinogenic compounds, which are produced by certain fungi. If they are present in the raw material, they are hardly destroyed by roasting and frying; therefore, inspection of the raw material is obligatory. Prevention of the formation of these toxic compounds during storage relies on the prevention of the development of fungi, which can be achieved by the elimination of moisture migration to the product.

NOMENCLATURE

b	thickness of slice, mm
C	concentration of hydrogenated oil in total oil (%)
C_1	constant in Equation 52.7, kg/kg
C_2	constant in Equation 52.7, kg/(kg s)
C_3	constant in Equation 52.7, kg/(kg °C)
C_4	constant in Equation 52.7, kg/kg
C_5	constant in Equation 52.7, kg/kg
C_6	constant in Equation 52.7, kg mm/kg
c_p	specific heat ($\text{kJ kg}^{-1} \text{K}^{-1}$)
D	diffusion coefficient, m^2/min
D	mass diffusivity, m^2/s
D_o	frequency factor, m^2/min
DM_r	initial dry matter content of potato, kg/kg wb

E/R	activation energy/gas constant, K
F	fat content, kg/kg wb
h	convective heat transfer coefficient, $\text{W/m}^2 \text{°C}$
h_{fg}	latent heat of vaporization, kJ/kg
K_c	rate constant of the mass lumped capacity of the core, s^{-1}
K_e	rate constant of the mass lumped capacity of the edge, s^{-1}
K_{oX}	parameter in Equation 52.19
K_{oY}	parameter in Equation 52.20
K_X	rate constant in Equation 52.19 min^{-1}
K_Y	rate constant of oil uptake min^{-1}
K_{Xb}	parameter in Equation 52.19
K_{XC}	parameter in Equation 52.19
K_{XT}	parameter in Equation 52.19
K_{Yb}	parameter in Equation 52.20
K_{YC}	parameter in Equation 52.20
K_{YT}	parameter in Equation 52.20
k	thermal conductivity, $\text{W/m}^2 \text{K}$
k_o	frequency factor, g potato/g potato solids, min
k_1	coefficient in Equation 52.13, N^{-1}
k_2	coefficient in Equation 52.13 (g oil–g potato/g potato solids–g water, min)
L	potato strip thickness, m
M	moisture, g water/g potato
m	moisture, gr
M'	moisture content, wt %
\bar{M}	moisture loss (% initial weight)
M_o	initial moisture, gr
M_w	moisture content (% dry basis)
m_o	initial moisture content, gr
N_{ix}	flux of species i in x -direction ($\text{kg m}^{-2} \text{s}^{-1}$)
O	oil content, wt %
O_e	oil content at equilibrium, wt %
S	concentration of oil, g oil/g potato solids
T	temperature of frying oil, °C
t	frying time, s (min in Equation 52.19 and Equation 52.20)
TX	texture, N
W	final moisture content, kg/kg wb
X	moisture content, kg/kg db
x	variable distance across the thickness of the chip, m
X_b	parameter in Equation 52.19
X_c	parameter in Equation 52.19
X_e	moisture content in equilibrium, kg/kg db
X_o	initial moisture content, kg/kg db
X_{oe}	parameter in Equation 52.19
X_T	parameter in Equation 52.19
Y	oil content, kg/kg db
Y_b	parameter in Equation 52.20
Y_C	parameter in Equation 52.20
Y_e	oil content at infinite process time, kg/kg db
Y_{oe}	parameter in Equation 52.20
Y_T	parameter in Equation 52.20

Greek Symbols

α	fraction of the total water contained in the core in the raw material in Equation 52.11
α	thermal diffusivity, m^2/s
ε_i	volume fraction of species i ($\text{m}_i^3 \text{ m}_t^{-3}$)
θ	tortilla chip temperature, $^{\circ}\text{C}$
ρ	density, kg/m^3
τ	residence time, min

Subscripts

β	liquid water
γ	water vapor
σ	solid

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53 Cost-Estimation Methods for Drying

Zbigniew T. Sztabert and Tadeusz Kudra

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53.1 INTRODUCTION

Reliable estimates of the fixed capital investment (FCI) of a drying system as well as the system's fixed and operating costs are essential in the preliminary equipment selection or when comparing different competing systems. In general, estimation of the costs involved in a drying process follows the well-known methods used widely in the process industries [1–9]. This chapter provides the empirical information on shortcut methods that may be used to quicken the predictions of FCI and operating costs on the basis of the previously prepared process balances and one-parameter equipment sizing. The method described here is useful for preparing the rough cost estimates applicable to many conventional drying systems. However, extending this information to unique or newest methods of drying must be regarded with caution. When capital cost estimates of a definitive investment are required to obtain budget authorization, competitive quotes from vendors have to be requested, as these quotes are the most reliable estimates.

A dryer itself is only one component of the overall drying system. The system may comprise preprocessing equipment (extruders, grinders, mixers, and blenders), feeders, product-discharge devices, and postprocessing equipment such as gas cleaners (cyclones, filters, and scrubbers) and solvent-recovery systems (condensers and separators). In some cases, heat-recovery systems (exhaust-gas recycle, run-around coil, heat pump, or heat wheel) may also be used. Thus, drying systems for the same end product can be different. Therefore, the total fixed costs as well as the operating costs should be calculated for the complete drying system.

Cost components of drying are usually divided into the following two groups:

1. Fixed costs (fixed charges), which are incurred for a long period of time, tend to be unaffected by fluctuations in the level of production activity. They include: depreciation of equipment and buildings, interest charges for the investment capital, plant protection, insurance, fixed part of taxes and rents, fixed part of maintenance costs, and executive salaries (administration, overhead).
2. Variable costs (operating costs) tend to vary with the production level. They include: cost of raw materials, cost of product degradation, costs of energy and utilities, direct labor (operating, works transport, supervision, and laboratory control), interest on working capital, royalties, variable part of maintenance costs, and miscellaneous direct costs.

The estimation process of drying costs is schematically shown in [Figure 53.1](#). The following is a detailed presentation of the method for preliminary estimates preparation, together with a worked example of cost estimates.

53.2 FREE-ON-BOARD COST AND INSTALLED COST OF THE EQUIPMENT

Equipment-price data concern the free-on-board (FOB) cost and the installed cost of the equipment. The installed cost of a drying system covers the cost

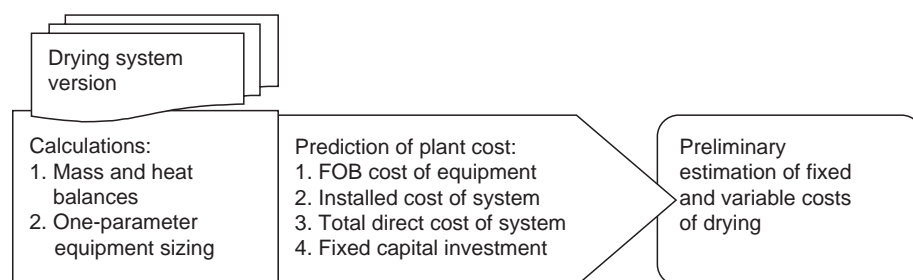


FIGURE 53.1 The steps of preliminary estimation of drying cost.

of the purchased equipment, cost of equipment installation, and costs of piping, wiring, and instrumentation. These costs can be estimated on the basis of vendors' data or, with less accuracy, on the basis of known costs for the equipment or system built previously, assuming a great degree of similarity.

Data for equipment costs obtained from technical literature are usually based on the technical and economic conditions of the past. Among the various methods accounting for price changes due to inflation and inevitable fluctuations in economic conditions, the method of the *Chemical Engineering Plant Cost Index* (based on the 1957/1959 cost level and abbreviated as CE Index) and the method of the Marshall and Swift All-Industry Equipment Cost Index (formerly known as the Marshall and Stevens Index) are the most commonly used. This last inflation index (formulated in the US market in the year 1926 with the set value of 100, and abbreviated as M&S Index) is recommended for updating the costs of high-level process equipments [4,10]. The updated M&S Index as well as the CE Index are listed regularly in *Chemical Engineering*, a periodical published by McGraw-Hill, New York [11], and are valid for cost updating in US\$ and for the US market. A part of such listing for the equipment-cost indices over the period 1989–2004 is given in Table 53.1. The following formula using M&S Index is widely accepted for equipment-cost updating:

$$\text{Updated cost} = \text{Original cost} \times \frac{\text{M\&S Index at updated time}}{\text{M\&S Index at the time of original cost}} \quad (53.1)$$

The M&S Index and the CE Index can also be used for world-price comparison, if the changes in costs according to the local economics, type of industry, and kind of equipment (besides time) are accounted for [12].

For scale-up or scale-down of the drying system, the cost of the equipment of the same type but of

different capacity can be predicted approximately from the following empirical correlation:

$$\text{Predicted cost} = \text{Original cost} \times [\text{desired capacity}/\text{original capacity}]^n \quad (53.2)$$

where desired capacity is the capacity of the equipment for which the cost is to be estimated and original capacity is the capacity of the same-type equipment for which the cost is known (original cost). Values of the exponent n for the typical equipment used in drying systems are listed in Table 53.2 [10]. If not given here or in the reference literature (e.g., [4,13]), $n = 0.6$ can be taken for rough estimation. The exponent n may vary with the design details. For example, in the case of cloth dust collectors, n ranges from 0.70 to 0.86. The cost estimated according to Equation 53.2 must be corrected not only for the inflation index according to Equation 53.1 but also for the

TABLE 53.1
Marshall and Swift All-Industry Equipment Cost Index and Chemical Engineering Plant Cost Index (1989–2004)

Year	M&S Index	CE Index	Year	M&S Index	CE Index
1989	895.1	355.4	1997	1056.8	386.5
1990	915.1	357.6	1998	1061.9	389.5
1991	930.6	361.3	1999	1068.3	390.6
1992	943.1	358.2	2000	1089.0	394.1
1993	964.2	359.2	2001	1093.9	394.3
1994	993.4	361.1	2002	1104.2	395.6
1995	1027.5	381.1	2003	1123.6	402.0
1996	1039.2	381.7	2004	1178.5	444.2

Source: Marshall and Swift Equipment Cost Index. *Chem. Eng.*, last pages of every issue (excerpted by special permission from *Chemical Engineering* (2005). Copyright 2005, by Access Intelligence, New York, NY 10038).

TABLE 53.2
Typical Exponents Relating Equipment Cost to Capacity

Equipment	Capacity Range	Unit	Exponent
Centrifugal fan	0.4–40	m ³ /s	0.44
Cyclone	0.01–4	m ³ /s	0.61
Multiple cyclone	4–94	m ³ /s	0.66
Dust collector (cloth)	1–470	m ³ /s	0.78
Electrostatic precipitator	23–470	m ³ /s	0.74
Gravity spray scrubber	0.7–189	m ³ /s	0.66
Venturi scrubber	3.3–38	m ³ /s	0.46
Impingement scrubber	1–23	m ³ /s	0.78
Centrifugal scrubber	3.3–47	m ³ /s	0.73
Ribbon blender ^a	0.2–7	m ³	0.54
Cage-type disintegrator	0.3–6	kg/s	0.68
Roll crusher	3–66	kg/s	0.97
Screen (vibrating)	0.2–6	m ²	0.66
Bucket elevator ^a	3–50	m	0.52
Belt conveyor ^a	5–19	m ²	0.50
Screw conveyor ^b	0.4–0.8	m ²	0.53
Pan dryer	1–19	m ²	0.50
Vacuum-shelf dryer	1.4–92	m ²	0.54
Tunnel dryer	2–10	m ²	0.50
Roto-Louvre rotary dryer	4–93	m ²	0.62

^aCarbon steel.

^bStainless steel, conveyor length \times diameter.

Source: Sztabert, Z.T., in A.S. Mujumdar, Ed., *Drying of Solids*, Sarita Prakashan, New Delhi, 1990, pp. 136–153.

local economic situation when installation will be erected in a country other than the United States [14]. Plavsic [12] has given a number of useful references [15–20] and provided a simplified “direct factoring” method for converting the US data to numbers that are applicable elsewhere. However, further discussion of this aspect is beyond the scope of this chapter.

Very approximate cost of carbon steel dryers can be calculated as [4]

$$C = (\text{M\&S Index}) \left[30 + 36.74 \frac{W}{(17.77 + T)} \right] \quad (53.3)$$

where C is the installed cost (US\$ in the year of the M&S Index) covering the dryer and auxiliaries, including foundations and erection, but no secondary dust collectors or building; W is the evaporation capacity, kg moisture/h; and T is the inlet-air temperature ($^{\circ}\text{C}$). This relation, however, does not reflect the material properties, so the calculated prices are underestimated. The graphs according to Equation 53.3 are presented in Figure 53.2 for the M&S Index equal to 1100.

Aside from the general formula, numerous data on dryer cost dependence on evaporation capacity exist in literature for the given dryers and definite operating parameters [13,21,22]. For example, cost of a spray dryer with a pressure nozzle atomizer, residence time of 16 s, inlet- and outlet-air temperature of 538 and 121 $^{\circ}\text{C}$, respectively, can be calculated as [21]:

$$\text{Price} = 13793.8 \times E^{0.435} (\text{first-quarter-1995 US\$}) \quad (53.4)$$

where E is the evaporative load ($181 < E < 38,780$ kg/h) and the price includes the 304-stainless steel dryer body, the access platform, support steel, burner, air-heater shell, feeding system, and instrumentation. Insulation, refractory material, fan, and solids-recovery system are not included.

The above method of prices presentation is restricted to a narrow range of technology. Taking into account that dryer cost depends mainly on its size and the material of construction, drying indices such as dryer throughput, evaporation rate, or unit heat consumption should not be used as parameters for cost estimation, mainly because the same indices may be obtained for dryers of different size by manipulating the operating conditions. This situation changes when a contract relating to technology and equipment is considered—then a dependence of dryer cost on evaporation capacity can be a valuable criterion for dryer selection.

The best characteristic parameters for preliminary cost estimation of most dryers are:

- The volume of the drying chamber (e.g., for spray or rotary dryers)
- The surface area of the supporting grid, shelves, belt conveyors etc.
- The heat-transfer area in case of contact dryers

Usually, the estimated fabrication cost of a dryer can be calculated from the following correlation:

$$P = (\text{M\&S Index}) \times A \times Q^c \quad (53.5)$$

(US\$ in the year of M&S Index)

where Q is the capacity of a dryer defined either by its effective volume (V) or by its characteristic surface area (S), depending on the dryer type. Values of the coefficient A and exponent c for some common dryers, determined from the available cost data, are listed in Table 53.3.

The fabrication cost of a drying unit besides its size depends also on the material of construction; it is the lowest for carbon steel. If the process requires a

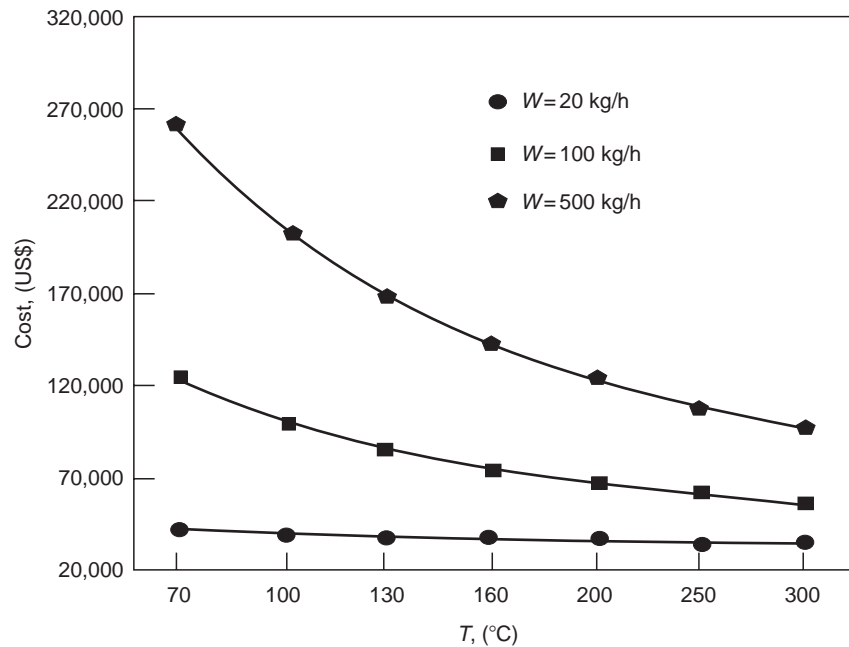


FIGURE 53.2 The dependence of dryer cost on evaporation capacity in kg moisture/h and inlet-air temperature in °C, according to Equation 53.3 (M&S Index = 1100).

TABLE 53.3
Parameters A and c in Equation 53.5 for Selected Dryers

Dryer Type	Capacity Range	Parameter		Construction	Equipment Included
		A	c		
Cabinet (atmospheric)	$16 < V < 160$	7	0.50	Aluminum	Heater, instrumentation
Cabinet (low vacuum)	$5 < V < 30$	18	0.62	Aluminum	Heater, instrumentation, vacuum pump
Vacuum with unheated agitator	$0.02 < V < 9$	66	0.30	Carbon steel	
Vacuum with unheated agitator	$0.02 < V < 9$	103	0.42	304-stainless steel	
Vacuum double-cone tumbler	$0.1 < V < 14$	80	0.70	304-stainless steel	
Indirect rotary (steam tube)	$3 < V < 300$	55	0.58	Carbon steel ^a	Instrumentation
Drum (single or double)	$1 < S^b < 40$	21	0.50	Cast iron	Drive, fan
Belt conveyor (through-flow)	$18 < S^c < 50$	21	0.59	304-stainless steel	Drive, fan, steam heater
Direct rotary, flue-gas heated	$20 < S < 300$	17	0.68	Carbon steel ^d	Drive, burner, heat exchanger, fan, dust collector, instrumentation
Direct rotary, hot-air heated	$20 < S < 300$	14	0.66	Carbon steel ^d	Drive, burner, heat exchanger, fan, dust collector, instrumentation
Fluid bed (batch)	$0.003 < V^e < 0.66$	141	0.47	304-stainless steel	Heater, fan, filter, instrumentation
Spray	$1 < V < 300$	47	0.52	304-stainless steel	Heater, atomizer, fan, filter, instrumentation
Vibrated bed	$1 < S^f < 12$	9	0.70	Carbon steel ^g	

^aFor 304-stainless steel, cost is multiplied by 1.7.

^bDrum peripheral surface area.

^cDeck surface area.

^dFor 304-stainless steel, cost is multiplied by 1.5.

^eBed volume.

^fSurface area of vibrated grid.

^gFor 304-stainless steel, cost is multiplied by 1.64.

Note: Dryer capacity is defined by dryer volume V in m^3 or surface area S in m^2 .

more chemical-resistant material, a cladding may be applied in some cases. In certain drying operations, optimum design of the equipment as well as product-quality requirements calls for the use of not only more durable but also more expensive materials. In general, drying equipment built of 304-stainless steel typically costs from 1.5 to 1.7 times more than that made of carbon steel and 2 to 3 times more if built of titanium. The cost depends also on the auxiliary equipment, and this is a reason for some uncertainty in cost prediction from literature sources without vendors' data. Figure 53.3 presents, as an example, the purchase cost for several dryers in relation to their capacity that is updated to the year 2000 with the M&S Index equal to 1089.

Once the FOB costs of all equipment comprising the drying system are established, the installed cost and other costs of the plant, as well as fixed costs, can be estimated using the factor method of cost estimation. Usually, the following components of installed cost are considered:

1. FOB cost (vendors' prices or estimated prices)
2. Assembly, piping, wiring, instrumentation using vendors' data, or the following factors:
 - Fully equipped compact system—25 to 30% of FOB cost
 - Fully equipped system, delivered in many parts—40 to 45% of FOB cost

- Equipment delivered without piping and instrumentation—71 to 99% of FOB cost
3. Freight (function of transport method, weight, and distance, or 7% of [component 1 + component 2])

The total direct cost of the plant expressed in terms of the “battery limits” is much higher than the installed cost of the equipment because of additional costs such as

- Buildings, laboratories, warehouses
 1. If a dryer system is installed inside a building, the cost of the building is factored in as 20 to 35% of the installed cost.
 2. If a dryer system is outside of the building, the cost of the building is factored in as 15% of the installed cost.
- Utility-supply facilities, site development, other direct expenses (20 to 40% of the installed cost).

To roughly estimate the total direct cost of the drying installation, it is recommended to multiply the equipment FOB cost by a factor of 2.25 if the equipment is made of carbon steel or by 2.75 if stainless-steel equipment is used [22]. These figures include the cost of piping, instrumentation, electrical connections, insulation, building space, and engineering when applied to

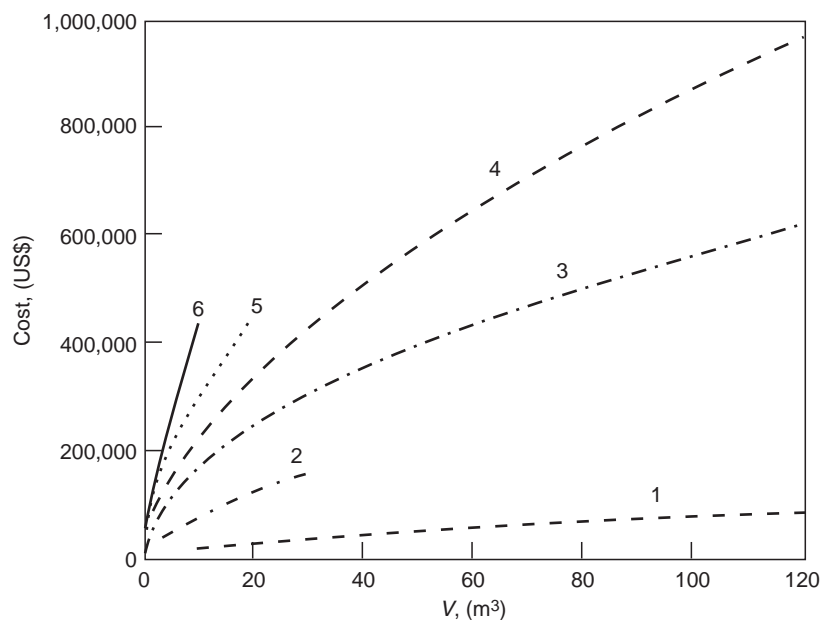


FIGURE 53.3 FOB costs (US\$, in the year 2000) vs. dryer volume in m^3 of selected dryers according to Equation 53.5: 1, atmospheric cabinet dryer (aluminum); 2, vacuum cabinet dryer (aluminum); 3, spray dryer fully equipped (304-stainless steel); 4, steam tube rotary dryer (carbon steel) with instrumentation; 5, vacuum dryer with unheated agitator (304-stainless steel); 6, double-cone tumbler dryer (304-stainless steel).

spray, flash, rotary, or fluid-bed continuous dryers. In specific cases, however, different factors may have to be used. For instance, the costs of instrumentation and piping may be included in the purchase price of a fully equipped dryer [23].

Besides the total direct cost, the FCI of a drying plant contains

- Contractor's fee (10 to 27% of direct plant cost)
- Insurance, customs, taxes, land, and other owner's costs (as per local regulations or 7% of direct plant cost)
- Contingencies (10 to 30% of direct plant cost)
- Procurement, supervisory, administration, and other owner expenses related to the plant (5 to 15% of direct plant cost)

Overall, the FCI of a drying plant that is sensitive to the dryer type and the product processed may be much higher, even by 3 times of the equipment cost. A number of dryers that form multistage, combined, or hybrid drying systems are often assembled from the components purchased from different suppliers. Therefore, the cost of the entire installation is not directly related to the dryer capacity but depends greatly on the cost of the ancillary equipment. When other operations such as conveying, grinding, or screening must be performed simultaneously with drying, the investment and operating costs can be reduced significantly by allowing for a partial write-off of the dryer cost for these secondary functions [13].

53.3 FIXED COST

Fixed costs (fixed charges) of a drying process are incurred for a long period of time and tend to be unaffected by fluctuations in the production level. The factor method of fixed-cost estimating is based on some simplifications such as:

- Fixed cost of drying is a function of installed cost of a drying system and FCI.
- Variable part of maintenance cost is given together with fixed part as a percentage of FCI.

The factors depend on the current economic conditions that should be analyzed up front, especially interest charges for the investment. Also the inflation index for local currency should be taken into account. In developed countries, the estimation of an annual fixed cost for drying can be made on the basis of the following factors:

- Depreciation of equipment—7 to 9% of installed cost

- Depreciation of buildings, utility facilities, site improvement—5% of these costs
- Interest charges on the investment—3 to 5% of FCI
- Fixed taxes and rent—2% of FCI
- Plant protection, insurance—1% of FCI
- Maintenance cost (fixed and variable):
 1. Complex system, explosive or toxic raw materials, significant number of rotating or vibrating parts—10% of FCI
 2. Simple system—5% of FCI
- Executive salaries and other fixed costs 6% of FCI

53.4 VARIABLE COST

The variable cost of drying is the last item in the cost-estimation process and it is predicted on the basis of heat and mass balances, shortcut factors, or, if available, vendors' data, unit cost of raw materials, labor, and utilities.

The cost of raw material depends on the applied technology and required characteristics of the end product. In any case, this cost should take into account the transportation costs to the considered site. When one compares several drying methods for the same product, it may be convenient to judge only costs of the lost and degraded material.

The cost of labor depends significantly on the mode of a drying process: batch or continuous. The labor cost of batch processes depends on the quantity of handled material, the type of the equipment, the method of material transportation, the method of feeding, and the schedule of plant operation. Rough estimation can be made considering the following rule: $2 \text{ man-hour}/(\text{m}^3 \text{ of dried material}) + 1/3 \text{ man/dryer/shift}$. The unit cost of labor should reflect the local economic conditions and the cost should include the insurance, taxes, and social charges. The labor cost of continuous drying with an automatic process control is almost independent of the production rate because supervision of control panels and supervision of the feeding and discharge systems do not depend on the throughput of the drying system. Overall, with the increasing automation, the direct labor cost decreases, but they are somewhat offset by the increase in labor cost to maintain the control devices and higher initial capital costs. Rough estimation can be made considering the following rule: $1/2 \text{ man/dryer/shift} + \text{operating labor of the other equipment in addition to the dryer}$.

The utilities costs are usually estimated on the basis of mass and heat balances, on the equipment characteristics, and on the utilities' prices. The prices

of utilities have to be corrected for local conditions and historical trends. The main constraint at this point is the cost of energy because of unstable prices of fuels, which may significantly vary both in the long and short terms. The historical trends are shown in Figure 53.4, where one can see the well-known effects of the Iranian Revolution and the Iran–Iraq War during 1979–1980 on the crude oil and secondary fuel prices. Besides fuel-price fluctuation caused by historical events, there is also a seasonal fluctuation influenced by supply and demand. For example, price of heating oil given by NYMEX increased from 32 ¢/L in September 2004 to 40 ¢/L in October and November 2004, and then decreased to 34 ¢/L in January 2005 [24].

Fuel-prices data and estimation for the years 2001–2006 in the US market are presented in Table 53.4 [25]. Trends of electricity prices indicate slower increase than the increase of natural gas and oil prices [26]. In long terms, the most stable prices are characteristically for electricity and coal though the cost of coal delivery and managing may be significant. Average costs of coal receipts at electric-utility plants in the United States by some states in dollars (Spring, 2000) per short ton are presented in Figure 53.5. More information on fuel prices can be found on the

following websites: <http://www.coaltrade.com>, <http://www.eia.doe.gov>, <http://www.energyshop.com>, <http://www.oilenergy.com>, <http://www.wtrg.com/prices.htm>.

The problem of energy price volatility can be resolved by dividing the utility price into separate terms, each of them being a product of a certain coefficient and the price parameter [27]. The first term reflects the conventional inflation rate while the second one represents the contribution from the fuel price:

$$P_{\text{utility}} = A \times (\text{CE Index}) + B \times P_{\text{fuel}} \quad (53.6)$$

The utility cost coefficients A and B are characteristic of the considered utility and can be given either by a single number, as in the case of purchased electricity ($A = 0.00013$ and $B = 0.01$), or by a correlation, as in the case of compressed air ($A = 0.00005 q^{-0.3} \ln p$ and $B = 0.0009 \ln p$, where q and p are the total air-plant capacity and the delivered air pressure, respectively). Tabulated data on the utility cost coefficients are given by Ulrich [27,28]. The CE Index in Equation 53.6 is the Plant Cost Index reported monthly in *Chemical Engineering*.

The variable maintenance cost, according to the previous assumption, is estimated together with a fixed part in relation to FCI.

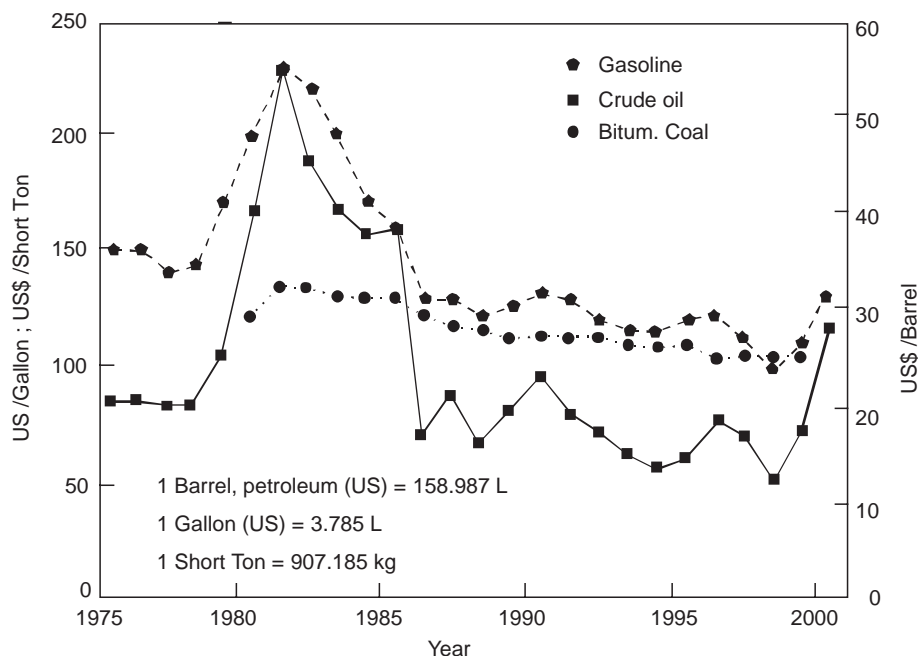


FIGURE 53.4 Average US first-purchase price of crude oil (1996 US\$/barrel)—the right scale, gasoline price (1995 US\$/gallon)—the left scale, and bituminous coal in 1996 US\$/short ton—the left scale. (From US Energy Information Agency (<http://www.eia.doe.gov>).)

TABLE 53.4
Fuel Prices for the Years 2001–2006 (United States)

Year	2001	2002	2003	2004	2005	2006
Crude oil spot prices (FOB) (\$/barrel)	21.2 (March)	22.5 (March)	26.5 (March)	31.1 (March)	47.6 (March)	
Crude oil West Texas intermediate (\$/barrel)			31.12 (year)	41.44 (year)	58.77 (year prediction)	63.46 (year prediction)
No. 2 heating oil, wholesale (\$/gallon)	0.76 (year)	0.69 (year)	0.88 (year)	1.13 (year)	1.70 (year prediction)	1.76 (year prediction)
Natural gas, City Gate (\$/thousand cubic feet)	5.72 (year)	4.12 (year)	5.8 (year)	6.6 (year)	7.18 (June)	
Natural gas, average Wellhead (\$/thousand cubic feet)	4.01 (year)	2.95 (year)	4.89 (year)	5.50 (year)	7.81 (year prediction)	7.64 (year prediction)
Natural gas residential average price (\$/thousand cubic feet)	9.63 (year)	7.89 (year)	9.51 (year)	10.74 (year)	13.03 (year prediction)	15.33 (year prediction)
Electricity nominal industrial average retail prices (¢/kWh)	5.04 (year)	4.88 (year)	5.13 (year)	5.11 (year)	5.71 (June)	
Electricity residential average prices (¢/kWh)	8.62 (year)	8.46 (year)	8.70 (year)	8.92 (year)	9.22 (year prediction)	9.37 (year prediction)
Imported coal average prices (\$/short ton)	33.72 (year)	35.51 (year)	31.45 (year)	37.52 (year)	46.49 (June)	
Coal (11,700 Btu, 0.8 lb SO ₂ /mm Btu) spot price (\$/t)		18 (September)	17 (September)	30 (September)	37 (September)	
Coal for electric power sector (\$/million Btu)	1.23 (year)	1.25 (year)	1.27 (year)	1.35 (year)	1.55 (year prediction)	1.62 (year prediction)

Source: <http://www.eia.doe.gov/>.

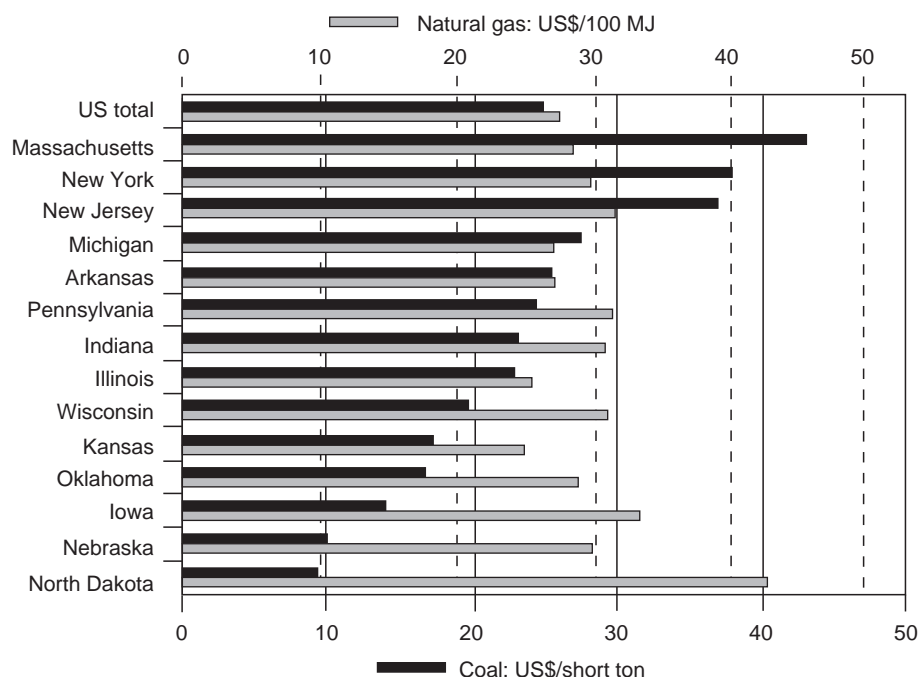


FIGURE 53.5 Average delivered cost of fossil fuel receipts at US Electric Utility Plants by state; coal: April–June 2000, US\$/short ton; gas: 1999, US\$/100 MJ. (From Federal Energy Regulatory Commission, FERC Form 423, “Monthly Report of Cost and Quality of Fuels for Electric Plants.”)

Servicing the working capital depends on the credit value and its type, as well as on the interest rate. It is reasonable to assume the credit to be equal to the variable cost of a drying process. In developed countries the interest charges on the working capital can be estimated as 4 to 7% of the annual credit, but it is strongly recommended to consider local conditions.

Royalties usually do not occur; but when a profit is calculated on the basis of new technology, royalties should be taken into consideration as a cost item.

When comparing the competing drying technologies, it is recommended to relate the capital investment (FCI) and the annual drying costs to the annual production rate. Usually, comparable estimates are made on optimistic and pessimistic assumptions of the level of costs (level of factors). With progress in computer-aided design, a number of software programs have been custom-developed in order to provide the engineers with an efficient tool for calculations of approximate cost [29,30]. Although such software programs are necessarily provided without warranty of any kind, the reliability of the estimate of $\pm 30\%$ covers the majority of the cases tested. An implementation of a process simulator with a cost evaluator is useful for the analysis of variants of a new plant, especially for

- Process calculation and equipment sizing
- Estimating fixed investment cost
- Performing a scheduling calculation, including direct and indirect labor costs
- Providing a detail evaluation of production cost
- Evaluating project profitability

53.5 DRYING COST ESTIMATE: A WORKED EXAMPLE

The following example of the cost-estimate procedure made in the year 2005, for example, illustrates the method of a preliminary cost evaluation for a drying process.

Problem description: Based on the previously calculated balance and one-parameter dryer sizing, the “factored estimate” for a cost of the laundry detergent formulating and drying in a spray dryer is to be done. The system will be located in Pennsylvania in the year 2006. The required capacity of the dryer is 4.9 t of the product per hour (production period: 7,200 h/y; 35,280 t of dry product per annum) of which the moisture content should be about 7% w/w. The feed moisture content is 32.5% w/w, the feed temperature is 80°C, and the price of raw materials in the slurry in the year 2003 was US\$1.48/kg (dry weight).

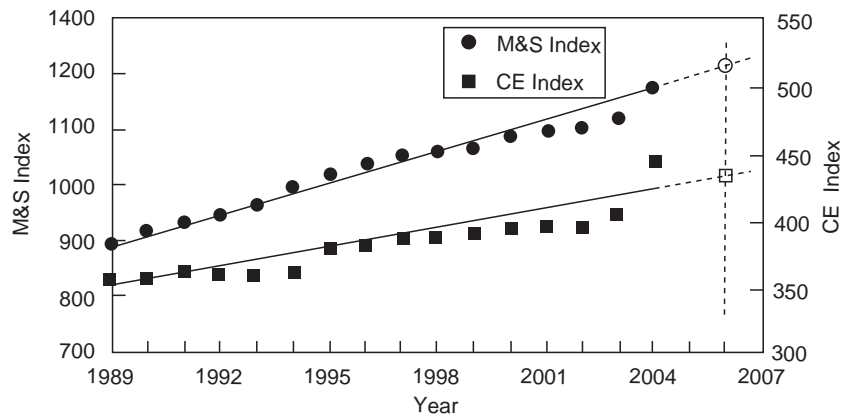


FIGURE 53.6 Trends in the M&S Index (left scale) and the CE Index (right scale) in the years 1989–2004 with extrapolation to the year 2006.

The task is to predict the cost of drying in a spray system in the year 2006.

Solution: Based on the data presented in Table 53.1, the M&S Index and the CE Index in the year 2006 are predicted by linear extrapolation, as shown in Figure 53.6. The values of the indices are:

$$\text{M\&S Index}_{2006} = 1221.7$$

$$\text{CE Index}_{2006} = 432.1$$

From previous experience, it is known that the laundry detergent can effectively be formulated and dried in a countercurrent spray dryer using combustion gases at 240 and 120°C at the dryer inlet and

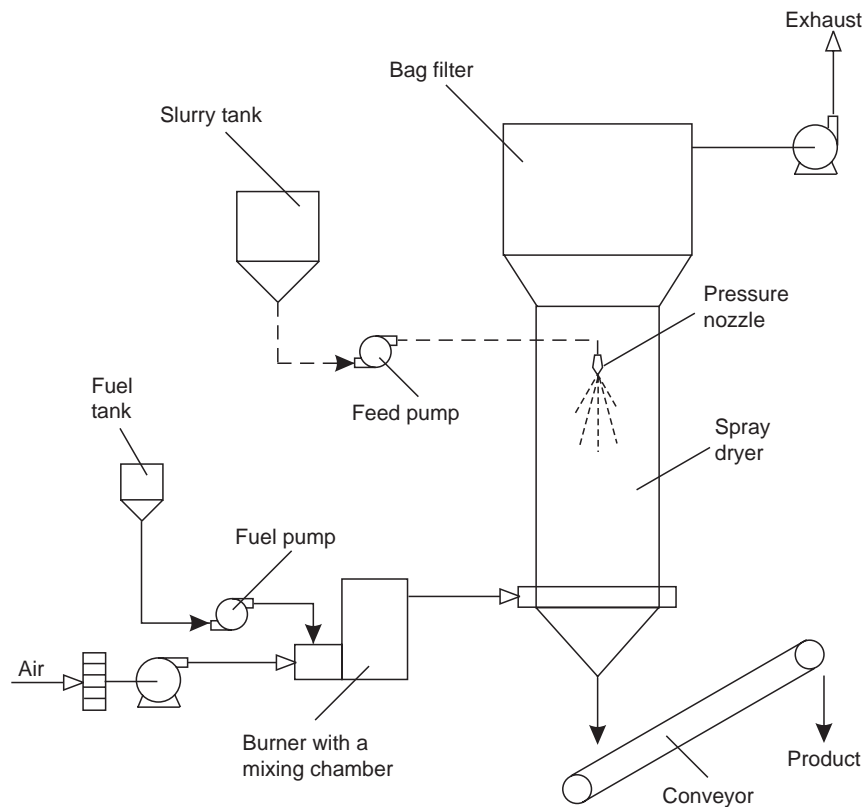


FIGURE 53.7 Flow sheet of the system for drying of the laundry detergent slurry.

outlet, respectively. The volumetric evaporation rate under these conditions is 3.5 kg H₂O/m³ h.

For the preliminary cost estimate, an assumption is made that ambient air is heated by the combustion of fuel oil; the flue gases are then mixed with fresh air to ensure the required gas temperature. Taking into account the specifics of the drying technology, moisture content of the feed, as well as fines that are likely to be entrained with the drying gas, the drying plant should consist of a slurry feeding system, an oil burner with a secondary air-mixing chamber, a dryer with pressure nozzles, and an exhaust-gas cleaning system (Figure 53.7). As product cooling, mixing of supplements, and packaging are downstream processes, they are outside of the battery limits of the considered project. The basic material of construction should be stainless steel.

The following values are known from the heat and mass balances:

The mass flow rate of the flue gas–air mixture entering a dryer: 44,000 kg/h

The quantity of fuel oil fed to the burner: 300 L/h (including heat losses)

The power of electrical motors: 190 kW

According to the specified values of the dryer capacity and the volumetric evaporation rate (3.5 kg H₂O/m³ h), the dryer volume should be 570 m³. Hence, the dryer FOB cost in the year 2006 can be calculated from Equation 53.5:

$$P = 1221.7 \times 47 \times 570^{0.52} = 1,554,000$$

(in the year 2006, US\$)

According to the data presented in Table 53.3, the above value includes the cost of a dryer, a fan, a filter, a heater, and the instrumentation. The cost of the remaining equipment calculated on the basis of the past cost data is US\$201,000 (in the year 2006). Then the cost of the entire drying system is to be US\$1,757,000. The installed cost of the system should contain the components presented in Table 53.5. As a result of calculations, the fixed cost for drying of 35,280 t/y of the laundry detergent is predicted for the year 2006 at the level of US\$1,449,780 per annum, resulting with US\$41.09/t of dried product.

TABLE 53.5
Application of the Factor Method for Predicting Fixed Capital Investment and Fixed Cost of Drying (Data for the Presented Example)

It.	Cost	Factor Range and Cost Basis	Estimate (US\$)
A	FOB cost of equipment	Vendors' data or actualized previous data	1,757,000
B	Assembly, piping, wiring, instrumentation (1,757,000 × 0.4)	30–99% of FOB cost	702,000
C	Freight [(1,757,000 + 702,000) × 0.07]	Weight, distance, or 7% of (A + B)	172,200
D	Installed cost	(A + B + C)	2,632,000
E	Building (2,632,000 × 0.3)	15–35% of installed cost	789,600
F	Utility facilities, site development (2,632,000 × 0.2)	20–35% of installed cost	526,400
G	Direct cost of plant	(D + E + F)	3,948,000
H	Insurance, custom, taxes (3,948,000 × 0.07)	7% of direct cost	276,360
I	Contractor's fee (3,948,000 × 0.2)	10–27% of direct cost	789,600
J	Contingencies (3,948,000 × 0.15)	10–30% of direct cost	592,200
K	Procurement, supervisory, administration, and other owner expenses related to the plant (3,948,000 × 0.07)	5–15% of direct cost	276,360
L	Fixed capital investment	(G + H + I + J + K)	5,882,520
1	Depreciation of drying system (2,632,000 × 0.09)	7–9% of installed cost	236,880
2	Depreciation of building, utility facilities and site improve ([789,600 + 526,400] × 0.05)	5% of building, utility facilities and site improve costs	65,800
3	Interest charges on the investment (5,882,520 × 0.055)	3–5% of FCI	294,130
4	Constant taxes and rent (5,882,520 × 0.02)	2% of FCI	117,650
5	Plant protection, insurance (5,882,520 × 0.01)	1% of FCI	58,830
6	Maintenance (4,843,000 × 0.055)	5–10% of FCI	323,540
7	Executive salaries and other fixed costs (5,882,520 × 0.06)	6% of FCI	352,950
8	Fixed cost of drying (annual)		1,449,780
9	Unit fixed cost of drying		41.09/t

TABLE 53.6
Predicted Variable Cost of Drying (Data for the Presented Example)

It.	Cost	Cost Basis	Estimate (Annual, US\$)
A	Fuel	300 L/h \times 7200 h \times 46.5 ¢/L	1,004,400
B	Electricity	190 kW \times 7200 h \times 6.3 ¢/kWh	86,184
C	Labor	2 workers \times 8760 h \times \$27/h	473,040
D	Royalties	Neglected	0
E	Lost of raw material	5 kg/h \times 7200 h \times \$1.59/kg	57,240
F	Working capital servicing	(A + B + C + D + E) \times 4%	64,835
G	Variable cost		1,685,699
H	Unit variable cost of drying	(G/35,280 t)	47.78/t

According to the forecast for the year 2006, the cost of fuel oil (loco system) will probably approximate 46.5 ¢/L and the unit cost of electrical energy will be of about 6 ¢/kWh. The unit cost of raw materials in the year 2006 predicted on the basis of the price in the year 2003 as US\$1.48/kg (dry weight) as well as the values of the CE Index₂₀₀₃ and the predicted CE Index₂₀₀₆ is

$$1.48 \left(\frac{432.1}{402.0} \right) = \$1.59/\text{kg}$$

The unit labor cost with insurance, tax, social, and other personal expenses is expected at US\$27/h and the interest rate of about 4%. The variable cost items for drying of the laundry detergent slurry calculated according to the presented method are summarized in Table 53.6. The predicted variable cost is about US\$1,685,699 per annum and US\$47.78/t of dried product. Then, the expected unit cost of drying of the laundry detergent in the year 2006 will most likely be

$$\text{US\$41.09/t} + \text{US\$47.78/t} = \text{US\$88.87/t}$$

Because the fixed part of the unit cost includes the variable part of the maintenance cost (as in the assumption), the value to be compared with the costs of alternative technologies for drying of the laundry detergent should be the sum of both costs, i.e., US\$0.089/kg.

SUMMARY

Very often, the published cost data (a primary resource is *Chemical Engineering*) refer to a specific year and apply to a location of the plant in North America. It is therefore necessary to correct the cost estimate for the inflation index and also for different geographical locations. Local cost of energy, as well as the local regulations (environmental, taxes, cus-

tom) and legislative requirements, must be considered carefully on a case-by-case basis to estimate the drying costs at different locations using the published cost data for the United States. Also, no warranty can be made concerning the accuracy of the published correlations or data on costs of dryers, ancillary equipment, as well as fuel and energy costs, but these data are very useful for the shortcut methods of cost analyzing and for preliminary prediction of costs of a drying process on optimistic and pessimistic levels.

ACRONYMS

CE Index	Chemical Engineering Plant Cost Index
FCI	Fixed capital investment
FOB	Free-on-board
M&S Index	Marshall and Swift Index
w/w	Weight by weight

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